Establishing The Environmental And Economic Benefits Of Applying Advanced Thermal Hydrolysis To Existing Anaerobic Digesters In The Western Cape, South Africa

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February 2021

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ABSTRACT

The treatment of sewage by wastewater treatment works (WWTW) generates a solids by-product stream requiring treatment and disposal. The waste sludges generated are rich in essential nutrients and energy, and present an opportunity to turn a waste stream into a resource. Retrofitting a thermal hydrolysis process (THP) to existing anaerobic digestion (AD) sludge treatment has proven itself globally as a reliable means of increasing treatment capacity and creating a final product that is non-hazardous. This allows for more sustainable sludge disposal options. However, THP has not yet been proven in a South African context.

This research carried out a comparative desktop case study of the existing AD facility at the Cape Flats WWTW in the Western Cape, South Africa. The facility's equipment is due for an upgrade and it was investigated if an improved process could be created. The base case of maintaining existing conventional mesophilic anaerobic digestion (MAD) was compared against the case of retrofitting THP. This would increase capacity and improve final product quality. The site would become a regional sludge facility importing additional sludge from some of the surrounding WWTWs. This would divert sludge from landfill and create more sustainable disposal options.

Steady-state models were developed for the conventional MAD and THP-MAD. These models were developed to include a kinetics section, stoichiometry section and a weak acid/base chemistry section. The kinetics section used hydrolysis as the rate limiting step when applying saturation kinetics. A stoichiometry section takes input from the kinetics conversion and used the elemental compositions of both substrate and biomass while predicting the amounts of other AD products formed. The weak acid/base chemistry predicted pH and took into account corrections for ionic strength and temperature, which were found to be particularly applicable in the case of high solids THP digestion with the elevated dissolved concentrations.

Many of the WWTW's in Cape Town make use of nitrification-denitrification biological excess phosphorous removal (NDBEPR) activated sludge (AS) treatment, often preceded by primary sedimentation. The modelling thus considered a 60:40 mixture of NDBEPR wasted activated sludge (WAS) and primary sludge (PS). AD modelling accounted for the breakdown of polyphosphate (PP) with the uptake of readily biodegradable COD to form poly3-hydroxybutyrate (PHB). The models also predicted the extent of spontaneous magnesium ammonium phosphate (struvite) precipitation inside the digester, and as well as the effect this would have on digester alkalinity and pH.

Results showed that when THP is retrofitted 2.5 times more sludge could be processed using the existing digesters' volume i.e. without building any additional digesters. This results in sludge treatment throughput increasing from 60 dry tonnes of solids per day for conventional digestion to 153 dry tonnes per through advanced THP digestion. Modelling has shown in each case that important AD parameters, such as free ammonia concentration, pH, alkalinity, and methane production are within the correct range for stable digester operation while sludge stability was achieved.

Major operating expenses and savings were evaluated. It was estimated that retrofitting THP created a saving of over R70mil/annum, largely due to savings in sludge disposal, and produce 2.7MW of surplus electrical energy. Carbon emissions were assessed for each case with THP digestion reducing significantly more emissions than conventional digestion. Additional investment required to upgrade conventional digestion to THP digestion specifically at the Cape Flats WWTW site would create a payback of between 5 to 6 years.

ACKNOWLEDGEMENTS

The following have given me support and motivation throughout this project and I want to acknowledge the following:

- My family, my partner and her family have all been incredibly supportive. From the messages of support and gestures of encouragement through the late nights, I want to say a big thank you to all of them.
- My supervisor and co-supervisor have been great teachers and guided me through the research process. Thank you for your patience and sharing your depth of knowledge with me through this study.
- Projects Assignments (Pty) Ltd have provided support and mentorship through my studies. Thank you for the opportunities you have provided and continue to provide.
- The City of Cape Town Wastewater Department must be thanked for their cooperation and information sharing which has enabled such an interesting project to go ahead.

TABLE OF CONTENTS

List of figures

List of tables

List of symbols and acronyms

A	Composition subscript for nitrogen in organics empirical
Ac	Dissociated Acetic Acid (CH3COO-)
ACP	Amorphous calcium phosphate
AD	Anaerobic Digestion
Aer	D Aerobic Digestion
Alk	Alkalinity
Alk H3PO4	Phosphate Alkalinity (as mgCaCO3/l)
AS	Activated Sludge
AT	Total acetate in acetate weak acid/base sub-system
atm	Atmospheres
B	Composition subscript for phosphorus (P) in organics empirical
BEPR	Biological Excess Phosphorus Removal
BNR	Biological Nutrient Removal
BO	Biodegradable organics
BPO	Biodegradable Particulate Organics
BSO	Biodegradable Soluble Organics
C	Carbon
˚C	Degrees Celsius
Ca	Calcium
CaCO3	Calcium Carbonate
Capex	Capital expenditure
CAS	Conventional Activated Sludge
CH4	Methane
CHP	Combined heat and power generation
CO2	Carbon dioxide
CoCT	City of Cape Town
COD	Chemical Oxygen Demand
COD	in the influent wastewater
CSTR	Continuously Stirred Tank Reactor
CT	Total carbon in carbonate weak acid/base sub-system
CXHYOZNAPB	Biomass empirical composition
d	Day
DAF	Dissolved air flotation
DB	Electrons accepting capacity of biomass
DS	Dry solids
Ds	Electron Donating Capacity of the Substrate
E	Fraction of the Biodegradable COD converted to Biomass
ELCA	Environmental life cycle assessment
ER	Endogenous Residue produced with biomass lysis
ƒ	Value that relates the pH and equilibrium (pKp2) in AD model
F/M	Food to Microorganism ratio
fC	or αC Total organic carbon (TOC) to mass (VSS or molar mass) ratio
fcv	or αCOD Chemical oxygen demand (COD) to mass (VSS or molar mass) ratio
fEG	Endogenous residue fraction of PAOs

fEH	Endogenous residue fraction of OHOs
fH	or αH Hydrogen (H) to mass (VSS or molar mass) ratio
ƒiOHO	ISS fraction of the OHOs (mgISS/mgOHOTSS)
ƒiPAO	ISS fraction of the PAOs (mgISS/mgPAOTSS)
ƒm,ƒd and ƒt	Activity Coefficient (mono-, di- and tri-valent) Ionic species
fN	or αN Nitrogen (N) to mass (VSS or molar mass) ratio
fO	or αO Oxygen (O) to mass (VSS or molar mass) ratio
fP	or αP Phosphorus (P) to mass (VSS or molar mass) ratio
FRBCOD	Fermentable Soluble Biodegradable Organic COD
fS'up	Fraction of unbiodegradable particulate (with respect to total)
fS'us	Fraction of unbiodegradable soluble (with respect to total) COD
FSA	Free and Saline Ammonia
fS'b	Fraction of biodegradable COD
fS'u	Fraction of unbiodegrable COD
fXBGP	P fraction of the PAOs (mgP/mgPAOVSS)
fXBGPBM	Biological cell P fraction of the PAOs (mgP/mgPAOVSS)
fXBGPP	Fractional Polyphosphate P content of the PAOs
fXBHPBM	P fraction of OHOs (mgP/mgOHOVSS)
g	Gram
GHG	Greenhouse gas
H	Hydrogen
H	Elemental Hydrogen
h	Hour
H+	Hydrogen ion
H2	Hydrogen molecule, denotes dissolved hydrogen concentration
H2C03* Alk	Inorganic Carbon Alkalinity (as mg CaCO3/l)
HCO3-	Bi-carbonate
HGH	High grade heat (heat source in excess of 100^{o}C)
HRT	Hydraulic retention time
HRT	Hydraulic Retention Time
IC	Inorganic Carbon
ISS	Inorganic Settleable Solids
IWA	International Water Association
K	Degrees Kelvin
Ka	Dissociation Constant for Weak Acid/Base
Kc1	Equilibrium Constant for H2CO3/HCO3- weak acid sub-system
Kc2	Equilibrium Constant for HCO3-/CO32- weak acid sub-system
KH	Henry's law constant
KLA	Oxygen mass transfer coefficient (l/h)
Ks	Half Saturation Constant (mol/ l)
Kspm	Thermodynamic Solubility Product
l	Litre
LCCA	Life Cycle Costs Analysis
LGH	Low grade heat (heat source under 100^{o}C)
MAD	Mesophilic anaerobic digestion
Me	Counter-ion metals (include cations of Mg, K and Ca)

MePO3	Polyphosphate
Mg	Magnesium
mg	Milligram
MgNH4PO4.6H2O	Struvite
min	Minute
ML	Mixed Liquor
MLE	Modified Ludzack–Ettinger (activated sludge system)
MLSS	Mixed Liquor Suspended Solids (gSS/l)
MLVSS	Mixed Liquor Volatile Suspended Solids (gVSS/l)
MMx	Molar Mass (g/mol) (where x refers to the relevant element)
MXa	Mass of active biomass (mgAVSS)
MXv	Mass of volatile suspended solids (mgVSS)
N	Elemental Nitrogen
N2	Di-nitrogen molecule
Nae	Effluent ammonia concentration (mgN/ l)
ND	Nitrification-denitrification
NDBEPR	Nitrification-denitrification Biological Excess Phosphorus Removal
NH4+	Ammonium (mgN/ l)
Nm^3	Normal cubic meter i.e. volume at conditions of 1atm and 20^{o}C
NO3-	Nitrate (mgN/ l)
Nous	Organic unbiodegradable soluble Nitrogen (mgN/ l)
NT	Total nitrogen in ammonia weak acid/base sub-system
O	Elemental Oxygen
O2	Oxygen molecule
OHO	Ordinary Heterotrophic Organism
OP	Orthophosphates
Opex	Operating expense
P	Elemental Phosphorus
PAO	Phosphorus Accumulating Organism
pCO2	Carbon dioxide (CO2) partial pressure
pH	Activity of Hydrogen ions
pKa	negative Log10 of dissociation constant (Ka) in acetate weak acid subsystem
PP	Polyphosphate
PS	Primary sludge
PS	Primary Sludge
PST	Primary Settling Tank
PT	Total phosphorus in phosphate weak acid/base sub-system
Q	Flow rate (l/d)
Qe	Effluent flow rate (l/d)
Qi	Influent flow rate (l/d)
Qw	Sludge waste flow rate (l/d)
RBCOD	Readily Biodegradable COD (mgCOD/ l)
rHYD	Rate of hydrolysis
Rs	Sludge age or sludge retention time (SRT, measured in days)
s	Second

Sb	Biodegradable organics in reactor (mgCOD/ l)
Sbi	Influent biodegradable organics (mgCOD/ l)
Sbp	Biodegradable particulate organics (mgCOD/l)
Sbpe	Residual biodegradable particulate organics (mgCOD/l)
Sbs	Biodegradable soluble COD (mgCOD/ l)
SCFA	Short chain Fatty Acid
SRT	Solids (or sludge) Retention Time (d)
SS	Steady State
SST	Secondary Settling Tank
Ste	Total effluent COD concentration (mgCOD/ l)
Sti	Total influent wastewater COD concentration (mgCOD/ l)
Sup	UPO in COD concentration (mgCOD/l)
Sus	Unbiodegradable COD in influent (mgCOD/ l)
T	Temperature (˚C or K)
TDS	Total Dissolved Solids
THP	Thermal hydrolysis process
TKN	Total Kjeldahl Nitrogen (mgN/ l)
TOC	Total Organic Carbon (mgC/ l)
TP	Total Phosphorus (mgP/ l)
TSS	Total Settleable Solids (mgTSS/ l)
UCT	University of Cape Town
UPO	Unbiodegradable Particulate Organics
USO	Unbiodegradable Soluble Organics
V	Volume
VFA	Volatile Fatty Acid
VSR	Volatile solids reduction
VSS	Volatile Suspended Solid (mgVSS/l)
WAS	Waste Activated Sludge
WRC	Water Research Commission
WW	Wastewater
WWTW	Wastewater Treatment Works
X	Composition subscript for carbon in organics' empirical formulation (i.e. CXHYOZNAPB)
Xa	Active biomass concentration (mgAVSS/ l)
XBG	Active biomass of the PAOs (mgVSS/l)
XBH	Active biomass of the OHOs (mgVSS/l)
XEG	Endogenous residue of the PAOs (mgERVSS/l)
XEH	Endogenous residue of the OHOs (mgERVSS/l)
Xi	Inert (unbiodegradable) organics concentration (mgUPOVSS/l)
XIo	Inorganic settleable solids concentration (mgISS/l)
XV	Volatile settleable solids concentration (mgVSS/ l)
Y	Composition subscript for hydrogen in organic empirical formulation (i.e. CXHYOZNAPB)
YH	Ordinary heterotrophic cell yield coefficient
Z	Composition subscript for oxygen in organic empirical formulation (i.e. CXHYOZNAPB)
ZAD	Acidogenic biomass concentration
μmax	Maximum Specific Growth Rate (/d)

List of terminology

Conventional digestion	Conventional mesophilic anaerobic digestion with no sludge pre-treatment
Digestate	Effluent stream flowing from anaerobic digestion
Indigenous sludge	Sludge generated from the adjacent WWTW, including primary sludge and/or waste activated sludge
THP digestion	Thermal hydrolysis pre-treatment followed by mesophilic anaerobic digestion

1. INTRODUCTION

1.1. BACKGROUND

The treatment of municipal wastewater at a wastewater treatment works (WWTW) generates a solids by-product stream of sewage sludge which requires further disposal (Russell, 2019), and in many developing countries the sludge is usually sent to landfills. However, as the population of a city increases so do its wastewater treatment requirements, and thus its sewage sludge production. In South Africa sewage sludge is classified as hazardous waste and in 2017 a total production of 632 749 tonnes of sludge was produced, of which over 85% was disposed in landfills (DEA, 2018).

The disposal of sludge in landfills results in the creation of harmful greenhouse gas (GHG) emissions released as the sludge decomposes to methane. Diverting waste from landfills has been identified as a means of reducing South Africa's contribution to greenhouse gas emissions in an effort to fight climate change (South African Department of Environmental Affairs, 2010). A six-year compliance timeframe in legislation from the Department of Environmental Affairs (2013) has recently lapsed and, as of September 2019, new South African laws dictate that sludge under 60% dry solids (DS) may no longer be disposed of in landfills and organics will soon be banned from landfills altogether. This requires costly means of increasing the sludge's DS, such as adding lime to the sludge, and alternate sludge disposal options urgently need to be sought.

An option is to dispose of sludge on "sacrificial land", an agricultural practice with restricted use (e.g. no foods crop production), but space is crucial in urban cities. Disposing of sludge on land, be it a landfill or sacrificial land spreading, consumes space and places strain on a city's space availability (Lam, Lee and Hsu, 2016). From 2012 the City of Cape Town forecasted that it has around 15 years of waste disposal space left (Western Cape Government, 2012). While the City of Cape Town explores various alternative sludge disposal measures e.g. anaerobic digestion, the increasing population growth in Cape Town requires that more sludge be processed than the existing facilities can handle. This introduces the need for an effective measure which will increase the sludge processing capacity of the current treatment plants in the short term, allowing time for long-term strategies to be developed.

Waste sludges generated from municipal wastewater treatment are rich in essential nutrients, such as nitrogen and phosphorous, and present an opportunity to convert a waste stream into a resource. Rather than dispose of this nutrient-rich sludge in landfills or incineration it would be beneficial to retain its nutrients within the municipal district. This would be beneficial as about 12% of the food processed in Cape Town is grown within the administrative area (Currie, Musango and May, 2017). Furthermore, fertilizers are energy intensive to produce (Gellings and Parmenter, 2004) and thus having a lower cost alternative derived from local biosolids would be worth considering. This would also then allow the retention of nutrients in a recycled economy, thus reducing dependency on sourcing and manufacturing nutrients elsewhere. This is particularly beneficial in the context of phosphorous (P) as macronutrient for crops. Further, phosphorous is finite resource and is mined from the earth, unlike nitrogen (N) which can be sequestrated from the atmosphere.

Disposing of sludge in landfills also misses an opportunity to produce renewable energy from the biodegradable organic material present in municipal sludge (Brent, 2016). Moving towards the generation of renewable energy from waste, sewage sludge can be processed into a fuel source via anaerobic digestion (a process which converts the biodegradable organics to combustible methane gas) or pyrolysis (Cao and Pawłowski, 2012). The generation of renewable energy from waste decreases reliance on fossil fuel derived power sources, thus further lowering greenhouse gas emissions further. The use of biogas (containing methane and carbon dioxide) in combined heat and power engines can significantly supplement the energy demands of wastewater treatment plants where the digestion takes place and potentially convert a site into a net energy producer (Carlsson, Lagerkvist and Morgan-Sagastume, 2016).

A more sustainable means of sludge disposal can be sought if specific processing is done to change a sludge's classification with regards to pathogens, stability and contaminants. Anaerobic digestion is a means of stabilising sludge to allow consideration for more sustainable disposal practices. However, pathogens in the sludge will still prohibit alternative uses than disposal to landfill or sacrificial land e.g. a useful alternative might be as a biofertilizer for unrestricted crop production (Herselman and Moodley, 2009). If a pathogen killing technology (e.g. thermal hydrolysis) is added in combination with to anaerobic then a complete pathogen kill will result in conjunction with stabilisation (Collivignarelli *et al.*, 2019). This then will allow the processed sludge to be considered as a biofertilizer for use on food crops, provided other contaminants are not present in the sludge.

The inclusion of a thermal hydrolysis process (THP) upstream of an existing anaerobic digester will increase the system's sludge digestion capacity (Perez-Elvira, Fdz-Polanco and Fdz-Polanco, 2010). This has been proven in numerous cases globally (Barber, 2016) and allows an increase in capacity without building any new large structures which might take up valuable space (e.g. more anaerobic digesters or sludge drying beds). The Integrated Waste Management Plan for the City of Cape Town has identified waste-to-energy as an important concept in both meeting the city's solid waste disposal and energy requirements (City of Cape Town, 2017). A centralised regional sludge processing facility would help achieve this. Further, the Water Research Commission (WRC) has been investing in studies of how to make anaerobic digestion and advanced digestion e.g. using THP easily understood in a local context and to promote its application for energy efficiency in wastewater treatment (Musvoto *et al.*, 2018). The City of Cape Town has one operational sewage sludge digestion facility at the Cape Flats WWTW. However, the city has over a dozen WWTW's generating a combined sludge quantity that far exceeds the capacity of the Cape Flats digesters, thus operating the digesters as a regional facility would have marginal benefit. It is expected that the concept of using THP technology towards increasing the capacity of the existing digesters and turning the site into a regional sludge processing facility would improve the City's sludge management operations.

The retrofitting of THP to sewage sludge digestion in the Western Cape has not been investigated with regard to quantifying the economic and environmental benefits. This research was done in the form of a desktop case study using the existing Cape Flats digesters as a base (termed "conventional digestion" in this study) onto which THP would be retrofitted to create a regional facility, termed "THP digestion" in this document. The conventional digestion case of maintaining the status quo of the Cape Flats digesters was compared to that of THP digestion using mathematical models and evaluative performance indices for strategic scenarios to increase the anaerobic digestion capacity at the Cape Flats WWTW. The environmental and economic evaluation of the selected strategic scenarios was done using the International Water Association (IWA) Benchmark Simulation Modelling (BSM) task group (Gernaey *et al.*, 2014) performance indices that have been adjusted towards a South African context (De Ketele, Davister and Ikumi, 2018). Ultimately, this study allows the City of Cape Town (CoCT), and other municipalities, to understand the drivers behind applying THP technology and in what context to consider its financial and environmental benefits for the management of sewage sludge.

1.2. RESEARCH PROBLEM

Research that can make anaerobic digestion (AD) more easily applied in South Africa is of value, as AD fundamentals are well enough understood, and the feasibility of the technology has proven itself to be beneficial for sludge treatment. Further research is then required into what technology exists that can not only enhance the anaerobic digestion process but also improve the digested sludge quality to allow a more sustainable disposal means. Thermal hydrolysis has proven itself in various installations around the world. There are over 50 working installations, many of which have been designed and built as sludge beneficiation centres. The applicability of this technology in a South African context is yet to be proven, and thus its financial and environmental benefits in a local installation are unknown (Musvoto *et al.*, 2018).

It has been proposed to convert the AD facility at the Cape Flats WWTW into a regional sludge facility by retrofitting THP to increase capacity without building any new digesters, and at the same time improve final sludge quality. This is done by operating high solids digestion (>10%DS) at higher solids loading rates and shorter sludge retention times (SRT) than conventional digestion (Barber, 2016). However, it is uncertain to what extent the THP will change the sludge fed to digestion. Further uncertainty exists around the feasibility of high-solids digestion at Cape Flats: the impact this may have on important digester operating parameters and the by-products that may be formed in the digestion significant amounts of THP sludge.

Increasing capacity allows large quantities of sludge from surrounding WWTWs to be imported to a regional digestion facility. In Cape Town many WWTW's use activated sludge (AS) treatment employing a nitrification-denitrification biological excess phosphorous (NDBEPR) removal process. The waste activated sludge (WAS) from NDBEPR AS contains high amounts of polyphosphate, which when digested in AD, releases high amounts of P and metals (magnesium, potassium and calcium). This impacts digester operating parameters and also increases mineral precipitation potential (Ikumi and Ekama, 2019). It is uncertain what impact THP and high solids digestion might have to the AD process when digesting large amounts (>50% fraction of feed) of NDBEPR WAS, and how this might change digester operation and effluent characteristics of the dewatering liquor returning from the AD to the adjacent WWTW.

The economic and environmental impacts, when comparing the THP digestion increased capacity case to conventional digestion require investigation. A previous study was done by Musvoto *et al* (2018) for the Water Research Commission (WRC) where it was concluded that implementing THP digestion is more economically beneficial than conventional digestion. However, Musvoto *et al* (2018) looked at a greenfield comparison between conventional digestion and THP digestion. That differs from this current research which considers a brownfields installation (as digesters already exist at Cape Flats WWTW). Further, in the study done by Musvoto *et al* (2018) each case processed the same amount of sludge (50tonDS/d). This ultimately lead to different AD volumes being required for each case (smaller volume for the THP digester case due to a higher loading rate). The difference in this research is that as the digesters are existing the volume is kept the same in both cases, and rather the initial capacity of the conventional digestion case is compared to the increased capacity of THP digestion possible due to retrofitting of THP upstream of digestion. Similarly another study considering THP with AD in a South African context was done by Petrie *et al* (2016). The study focused on improving energy efficiency by varying plant configurations. They concluded that THP digestion can provide long-term economic improvement over conventional digestion. However, this was for the same treatment throughput in each case. Further, the other work discussed here did not look at the breakdown of operating cost items and the net saving benefits additional investment in THP creates over maintaining the status quo. No other literature was found that compared conventional and THP digestion in a South African context.

The research proposed here will compare conventional digestion and THP digestion using the same volume digesters in each case. This research will provide local authorities a useful tool for considering the applicability of this technology as a means to improving their sludge management practices, especially when deciding to retrofit THP to an existing installation. It will also provide insight into what the impact is of importing and digesting large quantities of NDBEPR sludge and what changes the THP high solids digestion brings in relation to conventional AD. Finally, it quantifies operating costs associated with additional investment and to what extent THP digestion can be economically retrofitted to an existing installation.

1.3. RESEARCH QUESTIONS

The overarching research question for this study is: *Is it environmentally and economically sustainable to implement high solids digestion through the retrofit of thermal hydrolysis to mesophilic anaerobic digestion of sewage sludge in the City of Cape Town rather than maintaining the current practice of conventional mesophilic anaerobic digestion?*

Further questions expanding to continue this theme are:

- How do important operating parameters in AD compare in THP mesophilic anaerobic digestion (MAD) to those typically observed in conventional MAD, and using a case study at Cape Flats WWTW, what impact do these differences have on the AD operation and digester effluent?
- What are the anticipated economic and environmental benefits to be realised by the stakeholders with the application of THP technology?
- What key drivers do interested parties need to consider before implementing a THP project of retrofitting the technology to conventional anaerobic digestion?

By answering the above questions, the research will work towards achieving its objectives and generating outputs.

1.4. RESEARCH HYPOTHESES

It is hypothesized that, considering the current sludge disposal means available in Cape Town, it is more economically and environmentally sustainable to retrofit thermal hydrolysis process (THP) to an existing anaerobic digestion sewage sludge facility rather than simply maintaining conventional anaerobic digestion facilities as is. The retrofitting of this technology will:

i. Increase the capacity of existing conventional anaerobic digestion facilities.
ii. Generate financial payback and permanent ongoing long-terms savings.
iii. Create to a more environmentally beneficial sludge management method.

It is expected using THP to create a regional facility will be a beneficial sludge management system when integrated into the City of Cape Town's sludge management strategy.

1.5. RESEARCH OBJECTIVES

In light of the need for this research discussed above the following objectives for this research are listed below.

1. Review the fundamental changes THP generates when applying high solids digestion of both primary sludge and NDBEPR WAS and incorporate these changes as

adjustments while developing steady-state models to enable the comparison of THP MAD and conventional MAD.
2. Identify major operating costs and environmental benefits associated with treating sludge using MAD and how these are impacted by increased throughput when creating a regional treatment facility by upgrading an existing AD facility using THP.
3. Perform a case study on the Cape Flats WWTW AD facility, whereby the modelling tool from (1.) is used to investigate the economic and environmental benefits retrofitting THP can achieve over conventional MAD, highlighting major aspects which create project drivers for stakeholders.

1.6. RESEARCH SCOPE AND LIMITATIONS

The following are the scope and limitations of the project to fulfil the Master of Engineering (MEng) research project requirements.

- The thesis shall involve the utilisation of steady state equations that virtually replicate processes in the AD system to build an MS Excel spreadsheet model, hence the model shall not be a complex dynamic one as those found in simulation programs. The existing anaerobic digestion steady state theory (Sötemann *et al*., 2005; Harding., 2009; Ikumi, 2011; Ikumi and Ekama 2019) will be applied for estimating the capacity of the existing municipal sludge anaerobic digestion facility at the Cape Flats WWTW.
- It is assumed that the above-mentioned theory will also hold true for any new technology applied to the existing facilities and only the following will be impacted by the retrofitting of new technology:
  - Potential increase in throughput of existing anaerobic digesters.
  - Quality of final sludge produced (i.e. stability, pathogen levels, odour, etc)

1.7. DOCUMENT OUTLINE

The following section is intended to give the reader an overview of the structure of this research report and how each section contributes towards achieving the research objectives listed in Section 1.5.

Chapter 1 has served as an introduction to the research that will be carried out.

Chapter 2 serves to provide a literature review to give background to key aspects which must be considered in sludge management and how using anaerobic digestion is possible as a means to address sludge management challenges. Laws and guidelines for sludge usage are reviewed with the focus to identify what treatment is required to improve sludge disposal options. This includes anaerobic digestion and comparing it's sludge treatment ability against the guidelines for sludge disposal options. This includes reviewing a steady-state modelling method. This helps in identifying a means to build a steady-state model. The literature review goes on to explain THP and what benefits it may bring over conventional digestion. This is important as it provides information used to evolve a steady-state model for conventional digestion to one that can be used for THP digestion. Figure 2-1 shows the conceptual framework followed for this research and how various concepts investigated build on each other. These concepts are expanded on progressively through chapters 2,3, and 4.

Chapter 3 explains the methods used in the research. This chapter identifies a case study and motivates why this is a suitable case. The chapter then explains the methods that will be used to develop and evaluate the steady-state models. This will ultimately allow the environmental and economic assessment of what benefits may be realised by retrofitting THP technology to a conventional digestion process.

Chapter 4 goes into the detail used to build each model. The various input parameters are defined, and equations presented to calculate outputs are discussed.

Chapter 5 presents the results of the modelling exercise. It presents the data in a comparative form showing the comparison between conventional digestion and THP digestion. The first half of the chapter focuses on the impact THP and high solids digestion have on AD. The second half of the chapter then evaluates each case with regards to operating variables and their associated costs. The results are discussed and data presented in a way that conclusions can be drawn.

Chapter 6 lists the main conclusions from the research and compares these against how well the objectives have been realised.

Chapter 7 provides recommendations on aspects for further investigation.

2. LITERATURE REVIEW

2.1. CONCEPTUAL FRAMEWORK

The following literature review section gives an overview of the concepts in focus for this study. Current laws and regulations impacting sludge disposal are discussed along with current practices. Anaerobic digestion and thermal hydrolysis technology for processing sludge are discussed with the aim of improving sludge disposal options, as well as increasing the amount of sludge processed. The review includes the environmental and economic benefits of such practices. Figure 2-1 below gives an illustration of the study concepts and how they interlink with one another.

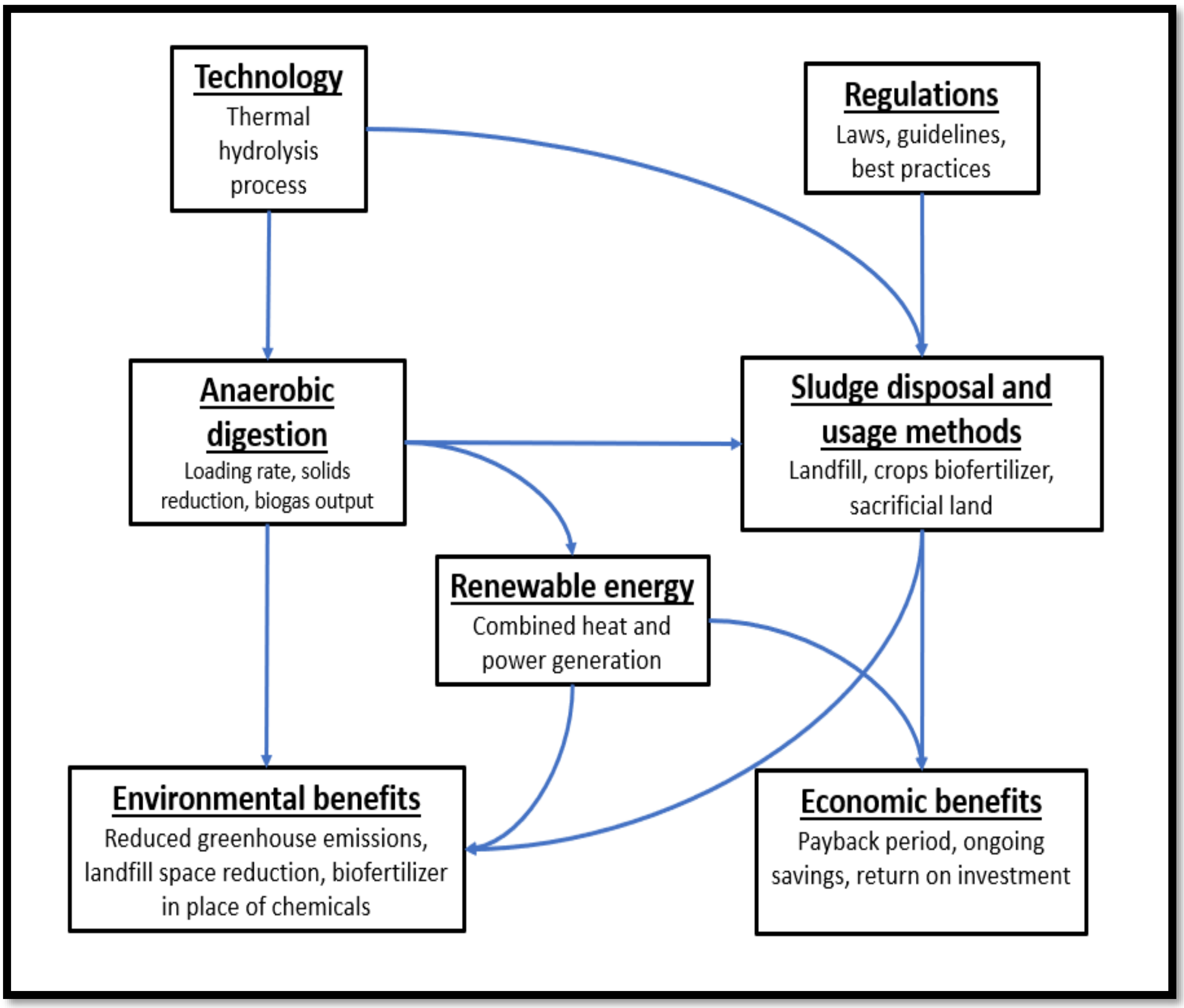

Figure 2-1: Conceptual framework linking the concepts investigated in this research

2.2. SLUDGE DISPOSAL LAWS AND REGULATIONS

When considering sludge disposal options in South Africa there are various laws which need to be considered. The Department of Environmental Affairs (2013) issued a set of regulations, particularly item 5 (q) R 634, which explains the type of substances eligible for landfill. One of the criteria listed is that moisture content shall not exceed 40%. Typically, sludge produced from WWTW's has a moisture content regularly exceeding 70%. To reduce moisture content and dry it further requires costly thermal practices, or vast space to create drying beds.

The regulations also refer to a limit on the calorific value of waste disposed to landfill. As of 2019 waste with a calorific value greater than 20MJ/kg will not be allowed to landfill, and by

2025 a limit of 10MJ/kg will be in place. According to a study done by Kim & Parker (2008) dry sludges have calorific values of 23 MJ/kg for primary sludge, WAS 19MJ/kg and digested sludge around 17MJ/kg. For some dried sludges, calorific value is likely to be the contributing factor and by 2025 most sludges will be restrained on a calorific basis. Therefore, alternate means of sludge disposal will need to be sought in the near future, even if moisture restrictions can be adhered to.

2.3. SLUDGE AS A RESOURCE

Biosolids sludges produced from sewage treatment contain useful nutrients which have been removed during the wastewater treatment process. A study in the UK showed that our diet results in almost half of the P content found in wastewater at around 44%, with food additives contributing a further significant portion of 29%, followed by laundry products at 14% and dishwashing detergents 9%, then small amounts from personal care products and food waste making up the balance. In some areas of the world dosing to reduce the lead content of tap water can make up to 6% of P in the wastewater (Comber *et al.*, 2013). Significant amounts of phosphorous are mined each year and fossil P resources are finite. Recycling P from biosolids can reduce the demand for fossil P and create a more sustainable approach in recycling nutrients rather than depleting finite reserves (Reijnders, 2014).

The nitrogen content of sludge found in sewage WWTW's originates mostly from the WWTW's processing of ammonia and organic nitrogen in the raw wastewater. The presence of nitrogen in wastewater is largely from eating habits, the food digestion, personal hygiene from bathing and washing of clothes (Patterson, 2003). This nitrogen is present in the domestic sewage sent to WWTW's. The nitrogen content in anaerobically digested sewage sludge is therefore sourced from the either the raw wastewater (from primary sludge) or from the growth of WAS in activated sludge reactors.

The application of biosolids from sewage treatment can provide valuable nitrogen (N) and phosphorous (P) macronutrients to crops, for example wheat and barley (Weggler-Beaton, Graham and McLaughlin, 2003). Rather than disposing of sludge in landfills it would be beneficial to retain nutrients within municipal districts. In Cape Town this could help create a nutrient recycle economy since 12% of the food consumed in Cape Town is grown within the administrative area (Currie, Musango and May, 2017). Furthermore, fertilizers are energy intensive to produce (Gellings and Parmenter, 2004) and thus having a lower cost alternative derived from local biosolids would be worth considering and would also reduce demand on relying on external nitrogen sources outside of the area.

2.3.1. Usage and disposal guidelines

A study assessing the use of liquid sludges applied to land in South Africa by Badza (2020) looked at supplementing nutrient requirements for crops as well as water requirements. The study found that liquid sludge would be beneficial at supplying nutrients to the soil, primarily focusing on nitrogen and phosphorous macronutrients. The study also concluded it would not be beneficial to use liquid sludge as a water source as it would supply an excess of nutrients, ultimately polluting the soil. This suggests that applying sludge to land does not have to be in a liquid form and so to reduce costs the transport of dewatered sludge cake would prove equally beneficial. The research concluded sustainable practices are to apply dry sludge to supply either nitrogen of phosphorous requirements, with the balance of nutrients supplied from other fertilizer sources if needed.

According to the Water Research Commission's Guidelines for the Utilisation and Disposal of Wastewater Sludge (2009) sludge must first be classified before it can be disposed of or applied to land. These guidelines are approved by the Minister of Water Affairs and Forestry and by the Minister of Environmental Affairs and Tourism. These guidelines require that the sludge is ranked under three categories: microbial and pathogen activity (Class A, B or C), stability (Class 1, 2 or 3) and pollutant content (Class a, b or c; made up of mostly heavy metals). For instance, a sludge of Class A1a is the highest category and can be applied freely to most agricultural crops, whereas a sludge of Class B2a needs to be used within restrictions (e.g. used only on animal feed crops and not food for human consumption and ploughed beneath the soil within a certain period of application to land). The regulations state that the treatment of sludge requires a volatile solids reduction (VSR) of at least 38% on a 90-percentile basis for treated sludge's stability to be classed as Class 1. The guidelines were purposefully developed as a user-friendly document for regulatory authorities, managers, practitioners and operators responsible for sludge management. The development of the Sludge Guidelines was also supported by an extensive stakeholder consultation processes in each province through the country. Therefore, complying with these guidelines ensures that all environmental requirements for safe beneficial use and/or disposal of municipal sludges are met and should provide to the sludge producer, an array of options for sludge disposal.

2.3.2. Processing sludge to beneficial use

Wastewater treatment plants apply various stabilisation and treatment processes to the sludge they generate in order to ensure it is of a suitable quality for specific disposal options (Badza, Tesfamariam and Cogger, 2020). Common stabilisation processes include aerobic and anaerobic digestion, which will improve the stability criterion in classification. Pathogen treatment and disinfection can be done by heat treatment e.g. pasteurisation or thermal hydrolysis, which will improve the microbial pathogen classification rating.

2.3.3. Sludge management in the Western Cape, South Africa

The City currently disposes of its sludge via a combination of landfill and application as a fertilizer to restricted land ("sacrificial land"). Testing data gathered by the City of Cape Town shows that heavy metal content of the sludge is within the limits required of a "Class a" pollutant level sludge. If pathogen content and stability criteria are then met the City's sludge would be a ClassA1a sludge allowing for essentially unlimited land application.

However, due to the current sludge treatment methods (AD, drying beds, dewatering) the land application of sludge is currently constrained as the sludge may only be applied to specific land used to grow animal feed, and it is not allowed to be placed on land used for crops grown for human consumption. Disposing of sludge on land, be it a landfill or sacrificial land spreading, consumes space and places strain on a city's space availability (Lam, Lee and Hsu, 2016). From 2012 the City of Cape Town forecasted it has around 15 years of waste disposal space left (Western Cape Government, 2012). While the City of Cape Town explores various alternative sludge disposal measures e.g. anaerobic digestion, the increasing population growth in Cape Town requires that more sludge be processed than the existing facilities can handle.

2.4. ANAEROBIC DIGESTION

Anaerobic digestion is used globally as a common means of sewage sludge treatment. This is because the fundamentals of anaerobic digestion are well enough understood and the feasibility of the technology has proven itself to be beneficial for sludge treatment. (Musvoto *et al.*, 2018). The below section gives an overview of the anaerobic digestion process.

2.4.1. Biological sludge treatment process

In anaerobic digestion, biological processes are carried out by bacteria which convert complex sludge to intermediate products, which are in turn converted to biogas and water. These subprocesses occur simultaneously in the anaerobic digester. First, hydrolysis breaks down complex carbohydrates, proteins and fats/lipids to sugars, amino acids and fatty acids. Acidogens then convert these by acidogenesis to hydrogen, carbon dioxide, ammonia, acetic acid and propionate. The propionate is converted to acetate and hydrogen by acetogens carrying out acetogenesis. Finally, acetoclastic methanogens convert acetic acid to methane and carbon dioxide, and hydrogenotrophic methanogens convert the hydrogen and carbon dioxide to methane and water. At low digester loading rates the acid formation step is likely to be rate limiting, but as loading rates increase methanogenesis may become the rate limiting step (Speece, 1983). In the digestion of sewage sludge the hydrolysis-acidogenesis step is the rate limiting process and is used to determine the biodegradable COD removal, methane production and biomass growth (S. W. Sötemann *et al.*, 2005). Therefore, if the hydrolysis rate can be improved then a higher throughput can be digested in the same volume digester.

2.4.2. Anaerobic digestion as a means of sewage sludge treatment

Typically primary sludge and wasted activated sludge produced from WWTW's are fed to an anaerobic digester where in controlled conditions are degraded to generate biogas made up of mostly methane and carbon dioxide, with the methane component being high enough for combustion. Thus, the biogas forms an energy source. Anaerobic digestion in sewage treatment reduces the volume of sludge discharged from a WWTW while reducing its biodegradable portion and thus stabilising the sludge (Kor-Bicakci and Eskicioglu, 2019). This allows alternate uses to landfill disposal, such as land application as fertilizer. Anaerobic digestion is therefore a sustainable and worthwhile means of sludge management and improves its disposal options.

2.4.3. Important process operation variables

Solids retention time

The solids retention time (SRT) in anaerobic digestion is an important process variable for which an optimal range exists. If the SRT is too low <4days then slow growing methanogens can be washed out of the digester resulting in a build-up of organic acids causing inhibition of methane production and a drop in pH, which can further worsen performance. High SRT values result in large digesters and hence greater capital and operating costs. Most sewage sludge anaerobic digesters operate between 15-30 days SRT for mesophilic conditions (Lee, Parameswaran and Rittmann, 2011).

Loading rate

The loading rate for a digester is defined as the specific amount of volatile solids per unit volume each day. For mesophilic sewage sludge digestion loading rates are typically in the range of 1.5-3.0kgVSS/$m^3.d^{-1}$ (Merwe-Botha, Borland and Visser, 2019). This is usually done at sludge feed dry solids concentrations of 3-6% (Higgins *et al.*, 2017). The loading, along with SRT, essentially define the volume of digester capacity required to treat a specific load of sludge and if loading rate can be increased then so can the amount of sludge processed.

Volatile solids reduction

In anaerobic digestion the volatile portion (VSS) of the TSS is broken down into soluble compounds and converted to AD products. Depending on the feed sludge type (PS or WAS) and operation a volatile solids reduction (VSR) of between 40-60% is achievable (Merwe-Botha, Borland and Visser, 2019). For conventional sewage digestion of mixed sludge the VSR tends to be on the lower end of this range, especially as the fraction of WAS is increased. Table 2-1 shows the volatile solids reduction possible from conventional mesophilic AD at an 18-day HRT for each type of sludge and for an equal mix of PS and WAS.

Table 2-1: Volatile solids destruction for each sludge type during conventional digestion (Jolly and Gillard, 2009)

PS	WAS	Volatile solids destruction
100%	0%	58%
0%	100%	32%
50%	50%	45%

Gas production

The conversion of biodegradable organics in AD creates various products. These products include the formation of methane (CH_4) gas and carbon dioxide (CO_2) gas. In conventional digestion operating at >15-day SRT with a 50/50 mixture of PS and WAS a gas production of around 385$Nm^3/tonVSS_{fed}$ can be produced with an average methane content of around 60%, with the balance being carbon dioxide. This translates into a gas production of 230$Nm^3CH_4/tonVSS_{fed}$ (Lee, Parameswaran and Rittmann, 2011).

Mixing

The continuously stirred reactor (CSTR) is a common operating method for anaerobic digestion. The digester must be well-mixed to ensure a homogenous distribution of feedstock and microorganisms, even distribution of temperature and achieving the correct SRT for all digester contents. As mixing consumes energy any means to improve the ease of mixing will help reduce the energy consumed by the operating plant (Lindmark *et al.*, 2014).

Temperature

Anaerobic digestion reactors are designed according to a specific temperature operating range. Two common temperature ranges are mesophilic digestion at 30-40°C and thermophilic digestion at 50-60°C (Lier and Pol, 2001). Kim *et al* (2002) explains that due to its stability mesophilic seems to be the more commonly applied operating temperature range. An external heat source is required to maintain digester temperature and often the biogas produced is used as a fuel source to generate the required heat.

pH

The pH of an anaerobic digester is important for the biological sub-process making up the overall digestion process. A literature review by Inglesby (2011) describes mesophilic digesters typically operate at around pH 7 as this best suits the methanogen bacteria which are the most pH sensitive of the various bacteria. If digesters are overloaded with feedstock pH can drop due to an acid build-up. A lower pH results in methanogen slow down which ultimately results in volatile acids not being consumed as fast as they are created. Acidogens still operate at lower pH's and continue to produce acids, and thus cause a further pH reduction. This can ultimately cause methanogen inhibition altogether. To rectify, feed rates must be lowered and alkalinity dosing e.g. sodium bicarbonate or sodium hydroxide can be applied.

A variable impacting digester pH is the split between undissociated and dissociated acetate species in the influent, which is governed by the degree of influent hydrolysis and influent pH. The biochemistry in the digester uses dissociated acetate to generate alkalinity by forming bicarbonate and methane. Thus, the higher the digester's influent pH and hydrolysis, the higher the fraction of dissociated acetate species, the higher the alkalinity generation and therefore the higher the digester pH (S. W. Sötemann *et al.*, 2005).

Ratio of carbon to nitrogen (C/N)

A low carbon to nitrogen ratio results in excessive ammonia production in anaerobic digestion which can raise the pH to above 8.5. This is toxic to methanogenic organisms and as a result may cause inhibition (Musvoto *et al.*, 2018). Conversely a high ratio can result in a lack of alkalinity being produced and a low digester pH. An optimal ratio exists around 20 to 30.

Ammonia concentration

The free and saline ammonia (FSA) concentration in AD is largely established from the breakdown of nitrogen containing organics. The FSA concentration in conventional digestion can range from around 500-1500mgl/l (Jeong *et al.*, 2019a).

Alkalinity and VFA concentration

Alkalinity is important to balance the concentration of volatile fatty acids (VFA's) produced from the breakdown of substrate. Alkalinity established during conventional digestion is typically in the order of 2000-5000mg/l as $CaCO_3$. The ratio of VFA concentration to alkalinity is referred to as the Ripley Ratio (RR). This should be maintained below 0.3 to allow the digester to operate at optimum pH. An RR starting to exceed 0.5 can be an indication of digester failure (Merwe-Botha, Borland and Visser, 2019).

2.4.4. Struvite precipitation in anaerobic digesters

The precipitation of struvite, also known as magnesium ammonium phosphate ($MgNH_4PO_4$), is common in anaerobic digesters containing significant levels of dissolved free and saline ammonia, orthophosphates and magnesium. Both ammonia and orthophosphates are released from the digestion of organics containing nitrogen and phosphorus present in municipal sewage sludges, and particularly so when digesting NDBEPR WAS. Magnesium levels in wastewater entering WWTW's can often be high enough that when sludge fed from the WWTW is subject to anaerobic digestion, magnesium combines with the ammonia and orthophosphates to form struvite. However, the limiting reagent is typically magnesium, as significant levels of FSA and OP are released during the digestion. The reaction forming struvite occurs according to the following stoichiometric process:

$$Mg^{2+} + NH_4^+ + H_2PO_4^- + 6H_2O = MgNH_4PO_4 \cdot 6H_2O + 2H^+ \qquad (2\text{-}1)$$

The formation of struvite in WWTW's tends to precipitate in equipment and can cause process plant fouling issues such as blocked pipelines and imbalances in rotating equipment e.g. pumps, dewatering centrifuges, etc. (Loewenthal, Kornmuller and van Heerden, 1994)

Struvite has the ability to be a useful fertilizer in agriculture. The nitrogen and phosphorous bound in the molecule provide essential nutrients for plant growth. A further benefit is due to the low solubility of struvite in neutral pH water a slow release of nutrients takes place, preventing overdosing soils with nutrients. It has also been found to have low heavy metal content and low concentration of other pollutants, such as polychlorinated biphenyls (PCB's) (Siciliano *et al.*, 2020).

Loewenthal *et al.* (1994) developed an experimentally verified model for the predication of struvite precipitation from water associated with anaerobic digestion. This was based on weak acid/base equilibrium chemistry and accounts for changes state as water chemistry changes with varying struvite precipitation potential and carbon dioxide (CO_2) partial pressure, a reduction in the latter being found to have significant impact to increasing struvite precipitation potential. The mechanism for precipitation is such that when CO_2 is released from solution, often by turbulence (pipe bends, inlet/outlet to processes, rotating equipment) and/or a change in system pressure, a corresponding increase in pH creates a state of supersaturation with respect to dissolved struvite, causing precipitation. The algorithms developed not only predict whether precipitation will occur, but also quantify the mass of struvite expected and the final water chemistry state.

2.5. RENEWABLE ENERGY SOURCE

The biogas generated from anaerobic digestion can be viewed as a renewable energy source. This biogas would have otherwise been generated in anaerobic conditions beneath landfills through decomposition of sludge, where in a purpose-built plant the decomposition is done at high rate in a controlled fashion and biogas collected for energy beneficiation. This energy can be used for to supplement the energy needs of the wastewater treatment plant. Gas production from anaerobic digestion has an average energy content of 21.5MJ/m^3 of biogas, and varies depending on the methane content. The combustion of biogas in a combined heat and power (CHP) engine can recover around 40% of the energy in the biogas as electrical energy and another 40-50% energy recovery as heat (Clarke Energy, 2020). This can help supplement the energy requirements of a WWTW.

An important aspect to a useful energy source is that its supply is continuous and can be relied upon. Sewage sludge is generated in constant quantities from WWTW's and there is an abundance of waste solids (Malla, 2011). This would allow the steady and continuous production of biogas, and thus a constant energy supply.

2.6. ADVANCED ANAEROBIC DIGESTION

The rate of AD is limited by the hydrolysis step. A means to improve digestion is to pre-treat sludge prior to AD This could be done via mechanical, chemical or biological means, or a combination of these. The aim is to cause hydrolyse organics into more soluble material and disrupt the sludge floc structure. This increases biodegradability, and solids reduction in AD, and enhances biogas production and digester effluent properties, such as dewaterability of the digested sludge (Carrère *et al.*, 2010). According to Carlsson (2012) thermal and ultrasonic pre-treatments are the most commonly researched, followed by chemical pre-treatment. Together these make up around 80% of studies on pre-treatment over the last several decades. Based on this thermal, ultrasonic and chemical enzyme treatment are briefly compared here. Motivation is then given to the most suitable pre-treatment for this study.

2.6.1. Thermal hydrolysis

Thermal hydrolysis process (THP) is a physical pre-treatment of sludge. The technology uses direct steam injection into the sludge to heat it to around 160-180^{o}C at 600kPag pressure for 20-30minutes. The pressurised mixture is then released rapidly causing high shear forces on the sludge. This will the combination of high temperature ruptures cells releasing material previously inaccessible during conventional AD. This hydrolysed material is then fed to AD (Barber, 2016).

2.6.2. Ultrasound

The ultrasound principle is physical form of sludge pre-treatment. It works by applying ultrasound waves of varying intensity to sludge. This generates low pressure zones in the sludge which causes liquid components to form microbubbles. The bubbles move towards through the sludge and increase in magnitude and eventually implode. This causes strong shear forces within the sludge causing the breakdown of sludge structure. Local temperatures and pressures can reach up to 500bar and 1000ºC. This is essentially cavitation and causes hydrolysis of the sludge (Bougrier *et al.*, 2006; Neumann *et al.*, 2016).

2.6.3. Enzyme treatment

Enzyme pre-treatment is a form of biological pre-treatment. Various hydrolytic enzymes are added to the process of anaerobic digestion to increase the transformation of polymeric compounds into more biodegradable substances. This brings with it benefits of increased biogas production and improved dewaterability of final cake. Common enzymes used are commercially available protease, amylase, carbohydrase and cellulase (Neumann *et al.*, 2016).

2.6.4. Pre-treatment selected for this study

The table shown here comapars some of the outputs the various pre-treatment discussed can provide. All methods increase both degree of solubilisation of feed sludge and result in increased biogas output during AD. Dewatering Is improved too.

Table 2-2: Pre-treatment methods to improve AD

	THP	**Ultrasound**	**Enzyme treatment**
Application/dosage to sludge as pre-treatment	170-190ºC at 20-30min	1000 – 16000kJ/kgTSS at 20-80min	Protease, Amylase, Carbohydrase, Cellullase; 0,06–18% (w/w TS)
Increase in solubilisation	40% (Bougrier *et al*., 2006)	22% (Pérez-Elvira *et al.*, 2009)	20% (Neumann *et al,* 2016)
Increase in biogas yield (biodegradation in period)	57% (Xue *et al.,* 2015)	42% (Pérez-Elvira *et al.*, 2009)	12% (Neumann *et al,* 2016)
Loading rate in AD	6kgVSS/m^3/d (Higgins *et al.*, 2017)	3kgVSS/m^3/d (Lizama *et al.*, 2018)	-
Disinfection	Yes, complete (Higgins *et al.*, 2017)	Partial (Tyagi and Lo, 2011)	-
Dewaterability increase	Yes, 30% DS (Barber, 2016)	Yes, 25% DS (Tyagi and Lo, 2011)	Yes, 27% (Jolly and Gillard, 2009)
Operating facilities (full-scale)	39 (Barber, 2016)	11 (Jolly and Gillard, 2009)	10 (Jolly and Gillard, 2009)

As shown in Table 2-2 above THP offers AD at a high loading rate of up to 6kgVSS/m^3/d. This is suitable for increasing the throughput of an existing facility. THP is has good references for complete pathogen destruction which is particularly important for a creating a Class A final sludge. THP allows for a high dry solids final dewatered sludge cake. This is beneficial in that it reduces transports requirements for the final dewatered sludge. This is useful for a final product that is likely to be transported long distances e.g. biosolids as fertilizer to outlying agricultural areas. THP as a form of pre-treatment is proven in any many cases on a full-scale. It is therefore a low-risk option for applying it as a new technology in South Africa where it has not been implemented on a full-scale scale before. THP was thus selected for this study as a suitable form of pre-treatment to be compared against conventional AD. It is also suitable for the purpose of this research's case study described in section 3.1.

2.7. THERMAL HYDROLYSIS OF SEWAGE SLUDGE

One of the more common methods of improving anaerobic digestion performance is using thermal hydrolysis to pre-treat the sludge before it is fed to the digestion process. There are numerous well-established technology providers of thermal hydrolysis, which have successfully installed multiple plants around the world. In 2016 there were already over 39 operating THP plants globally (Barber, 2016) and this number has since risen to over 50 (Musvoto *et al.*, 2018). Large water authorities such as Thames Water in the UK and elsewhere use THP extensively, due its multiple benefits. THP is therefore proven on a large scale and presents a low process risk. Further, the Water Research Commission has been investing in studies of how to make anaerobic digestion and advanced digestion e.g. using THP easily in a local context to promote its application for energy efficiency in wastewater treatment (Musvoto *et al.*, 2018).

A summary of the main drivers to installing THP are:

- increased loading rate, allowing more sludge to be processed through existing digesters, and in new installations smaller digesters are required, in comparison to conventional digestion
- improved dewatering, thereby creating a dryer final sludge cake and decreasing transport requirements
- increased energy production, realised from an increase in sludge biodegradability within the digester's SRT and subsequently greater biogas production per unit mass of sludge processed
- sludge sterilization, which allows for creating a Class A sludge and increased disposal options

2.7.1. Process overview

The following description of THP is drawn from an overview of the technology given by Jolly & Gillard (2009).

A THP module will consist of one pulper, batch reactors and a flash tank. The first tank, the pulper, will operate at near atmospheric pressure and continuously receive 15%-18% DS thickened sludge from upstream processes. Thickening to greater than this may cause heat transfer limitations and a more dilute sludge would increase energy requirements. The thickened sludge is pre-heated in the pulper tank and then fed to the batch reactors.

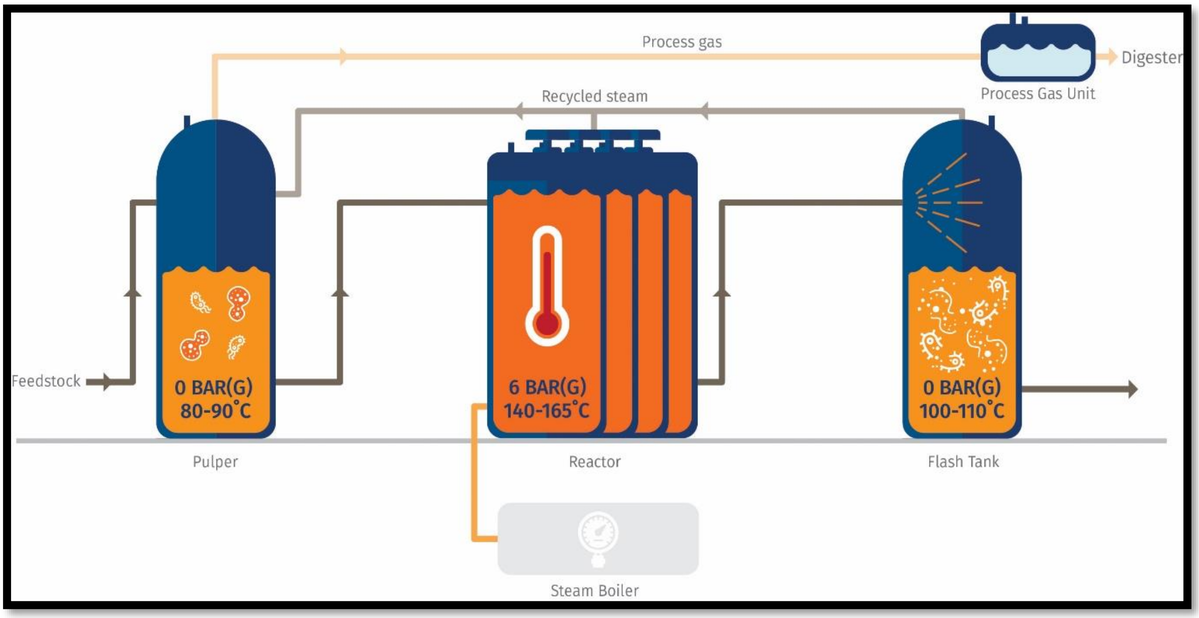

Figure 2-2: Schematic of the thermal hydrolysis process (Cambi, 2021)

Once in the reactors the sludge will be pressurised and heated to 165ºC and 600kPag using direct steam injection. The sludge will be held under these conditions for 20-30minutes where it will undergo hydrolysis. These are deemed optimal temperature and duration for THP. Organic matter will be hydrolysed into soluble compounds and the sludge will be sterilised with all pathogens being destroyed. The direct steam injection adds some dilution to the process flow.

Downstream of the THP reactors a flash tank will operate at near atmospheric pressure and will receive the hydrolysed sludge. The transfer will be done using the pressure difference between the reactors and flash tank and as a result, strong forces in the sludge will cause it to rupture due to the pressure differential. This is known as steam explosion and will break up the sludge cells and fibres. This will allow for the release of cell contents and entrained moisture, and as a result will substantially increase the overall volatile solids destruction possible in the downstream anaerobic digestion process. The steam released during the steam explosion step will be recycled from the flash tank back to the upstream pulper tank where it will be used to pre-heat incoming sludge.

Although the reactors are batch operation themselves, the feeding sequence of the multiple reactors is done in such a way that the THP process allows for a continuous process flow. The THP process provides excess heat to that required for mesophilic AD and sludge coolers are required downstream of THP to remove any excess heat from the hydrolysed sludge prior to feeding to digestion. Further, dilution water (potable water or treated effluent) is added to the hydrolysed sludge to control the feed solids to digestion to not exceed 10-12%. This is done to prevent ammonia inhibition which could result from an excessively high solids concentration in the downstream anaerobic digester.

2.7.2. Change in sludge physical and chemical characteristics

Increased solubilisation

Jeong *et al* (2019b) used a microscope to observe changes to WAS after THP treatment and noted destruction of cell walls. This resulted in the release of dissolved substances increasing the available material for conversion to methane, thus increasing biodegradability and biogas output of AD. It was also found THP is an effective method for solubilizing WAS.

THP creates an increased solubilisation of total COD, with significant increases in biodegradable soluble COD (bSCOD). Part of the bSCOD are VFAs. Donoso-Bravo (2011) observed that THP creates an increase in VFAs, and further, the proportion of VFA made up by acetic acid increased from around 30% in the raw sludge to 70% of VFA's being acetic acid in the treated sludge. Jeong *et al* (2019b) observed over 60% of the VFA's in THP sludge are acetic acid and a pH reduction of treated sludge coincided with the increase in VFA concentration, thereby confirming that thermal pre-treatment was effective at promoting hydrolysis and solubilization.

Most of the studies are focused on either THP of WAS or of a mixture of WAS and PS. There seems to be limited literature on the effect of THP on PS by itself. However, Wilson and Novak (2009) carried out THP experiments on WAS and PS separately and compared the results of each. They found that while WAS increased COD solubility from 1.2% to 17.8% the THP of PS increased solubility from 8.6% to 17.5%. Further, the VFA concentration on the PS hydrolysate was found to be 4 to 7 times greater than that in the WAS. This study showed that THP has useful effects increasing solubility of both PS and WAS. In the comparison they also found significantly more ammonia was released from the WAS compared to the PS, which was expected due to WAS having a higher protein content than the PS.

These changes increasing COD solubility ultimately allow increased AD performance due to COD becoming more readily available. Table 2-3 shows a summary of observations from various studies done on THP and the successive anaerobic digestion of the THP treated sludge.

Table 2-3:Summary of THP effects on sewage sludge digestion over various studies

Reference	***Conventional digestion*** (Lee, Parameswaran and Rittmann, 2011)	***Xue et al*** *(2015)*	***Han et al*** (2017)	***Jeong et al*** (2019a)	***Donosso-Bravo et al*** (2011)	***Higgins et al*** (2017)
Sludge type treated	50/50 PS/WAS mix	WAS	WAS	WAS	WAS	55%PS/45 %WAS
% TSS solids	5%	16.7%	10.1%	7.0%	7.7%	10.5%
THP temperature (^oC)	n/a	160	165	160	170	160
THP time (min)	n/a	30	50	30	30	30
Increase in digestion rate	n/a	57%	127%	66%	60%	-
Methane yield (Nm^3CH_4/tonVS S_{fed})	230	-	296	295	305	340
Biodegradability increase (within SRT period)	n/a	11%	-	-	20%	-
Solubilisation of COD - THP feed	2-3%	5%	-	3%	-	-
Solubilisation of COD - THP outlet	n/a	43%	45%	27%	30%	18%

rbCOD - THP outlet	n/a	-	-	-	-	19%
VFA - THP feed (g/l)	310	1450	250	200	320	-
VFA - THP outlet feed to digester (g/l)	n/a	1550	4210	400	500	-
VFA - Digester effluent VFA (g/l)	30	29	280	-	-	-
FSA - THP feed (g/l)	98	1100	270	228	130	-
FSA - THP outlet digester feed (g/l)	n/a	1450	1060	490	140	-
FSA- digester effluent (g/l)	640	2449	3020	-	-	2650
Alkalinity - THP feed (mg/l as CaCO3)	-	-	670	-	-	-
Alkalinity - THP outlet digester feed (mg/l as CaCO3)	n/a	-	4230	-	-	-
Alkalinity - Digester effluent (mg/l as CaCO3)	2290	-	17820	-	-	-
Ripley ration (VFA/Alk)			0.02			
pH of digester	6.95	7.82	8.03	-	-	7.7
Steady-state/batch	Steady-state	Batch	Steady-state	Batch	Batch	Steady-state
Duration (days)	15	28	20	-	21	15
VSR (THP + AD)	48%	49%	49%	-	-	54%

Comments	Steady-state lab-scale mesophilic AD tests on mixed	BMP tests on low temperature THP and high temperature THP sludge; significant viscosity reduction;	High solid digestion with of raw sludge and THP sludge; mesophilic and thermophilic; 200day experiment time period, PolyP is released to OP during THP step (and not the digestion step).	BMP tests done on THP sludge at different temperatures. pH reduction of treated sludge coincided with the increase in VFA concentration; WAS analysis with microscope images of intracellular matter.	21days time for biodegradability tests, Lab and pilot scale THP at different times 0-30min in 5min intervals at same temp 170ºC; total coliforms destroyed.	No control test on untreated sludge; THP at multiple temperature over 130ºC to 170ºC; experiment for 200days; observed if substrate was already highly biodegradable THP did not improve it further.

The following comments apply to Table 2-3.

1. The conventional digestion study is included as reference for comparison purposes. This is not the control against which other THP studies are carried out. Each THP study had their own control and the data shown in this table, for example % increase, was in relation to each studies' own control.
2. Some of the studies varied THP temperature and time. The data presented corresponds to the temperature and time shown.
3. Where data is not shown it was not presented in the literature source.
4. VSR is volatile solids reduction of the sludge through THP followed by AD.

It can be seen that ammonia production from THP can range from 140-1450mg/l before the sludge is fed to AD. A common theme is that THP digesters operate at higher FSA concentrations 2500-3000mg/l than conventional digestion of 500-1500mg/l. This is largely due to the higher solids concentration at which THP digesters can operate (Wilson and Novak, 2009) , made possible due to the reduced sludge viscosity improving mixing and pumping. THP increases the feed FSA to digestion due to solubilisation, and this trend is also seen to apply to VFAs and alkalinity.

Phosphates have been observed to increase through the THP unit process before digestion. Han *et al* (2017) found that most of the polyphosphate (PP) stored in PAO's of NDBEPR WAS released to OP during the THP process. The downstream digestion had a marginal increase in OP in relation to the OP generated through the THP process.

The pH of THP digesters tends at pH 7.5 to 8 tends to be higher than that of conventional digestion around pH 7. This is often due to higher alkalinity from the increased concentration of AD products – bicarbonate and phosphates.

Increased digestion rate

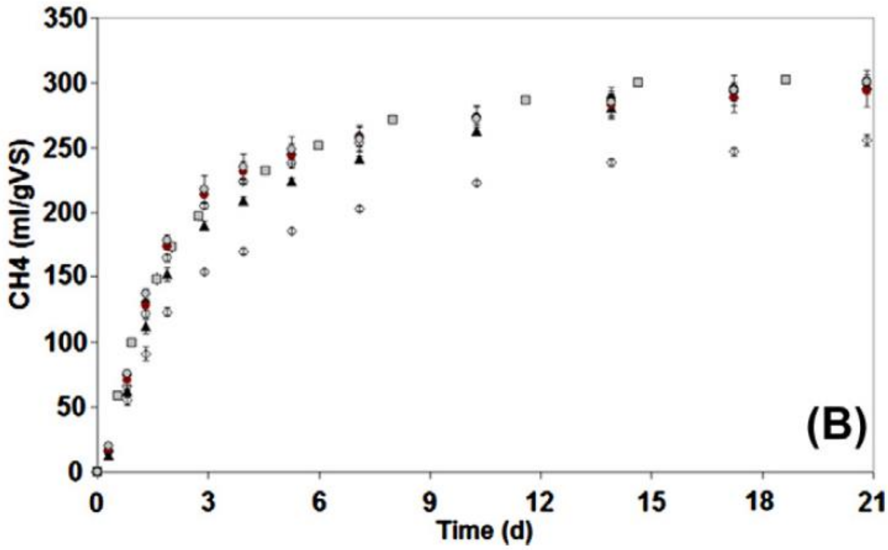

Figure 2-3: Biogas production from biomethane potential test taken from Donoso-Bravo et al (2011)

The plots shown in Figure 2-3 are the results of biomethane potential (BMP) tests done on untreated sludge and THP treated sludge by Donoso-Bravo *et al* (2011). The lowest curve of the set shown is from raw untreated sludge and the curves above are for increasing THP treatment time from zero to 30min (for which the curves show a steeper rate and higher overall biogas production). The observed rate increase of the THP sludges over untreated sludge is due to more soluble biodegradable organic matter being available after THP pre-treatment. A higher biomass specific growth is thus possible and with that an increased biogas production rate (Donoso-Bravo *et al.*, 2011). The increased rate allows digester SRT to be reduced from 18-20days to 12-14days, making effectively the same amount of biogas in less time. This suggestion was also made after BMP tests on THP sludge done by Xue *et al* (2015).

Increased biodegradability within SRT

A further observation is the increase in total biogas produced over the test duration. This implies that part of the inert organic matter, which could not be degraded in the test with raw sludge, turned into available organic matter, both particulate and/or soluble. The increases in biodegradability found in the literature suggest that biodegradability is increased only for the digester hydraulic residence time (HRT). In Table 2-3 the increase in the biodegradability of WAS due to THP varies from 11% to 20%.

During the literature review it was not found if the organics were in fact already relatively biodegradable but simply had been made more accessible to AD biomass through THP solubilisation i.e. slowly biodegradable substances becoming readily biodegradable. It was unclear if these substances under conventional digestion, if given more time (say 40days SRT), would have biodegraded anyway and thus produced the same amount of biogas as the THP sludge. Further, it was also not conclusive to what changes may have occurred to stubborn unbiodegradable substances i.e. substances difficult to breakdown that may require >100days digestion for any measurable change. Therefore, for the purposes of this research the concept of increased biodegradability is taken as the increase in biodegradability during the time period of the digester SRT.

Change in unbiodegradable fractions

Thermal hydrolysis of wastewater sludges can increase the soluble unbiodegradable COD, unbiodegradable dissolved organic nitrogen and colour. The production of these compounds increases with increasing THP temperature and exceeding 180-200°C is not beneficial in this regard. Colour changes are thought to be due formation of recalcitrant materials such melanoidins and other Maillard reaction products (MRPs), from the Maillard reaction between sugars and proteins that occurs at high temperatures in THP (Higgins *et al.*, 2017). WWTW's using UV for disinfection of final effluent should consider high THP temperature can cause tea-coloured UV absorbing material (Wilson and Novak, 2009). When WAS was subject to THP it was found that refractory COD, which is understood to be unbiodegradable COD, can increase by 11kg per tonne of dry solids dewatered in final dewatering (Oosterhuis *et al.*, 2014). Another study by Figdore *et al* (2011) found there to be up to 3000mg/l soluble refractory COD in the effluent leaving THP digestion. This same effluent had an unbiodegradable organic nitrogen content of 132gN/l. Another study found the average soluble unbiodegradable COD to be around 800mg/l Zhang *et al* (2016). Xue *et al* (2015) found that over 180°C there is a significant increase in unbiodegradable soluble COD and suggested this is most likely due to the production of melanoidins. It is recommended to operate THP around 160°C to reduce the formation of these by-products.

2.7.3. Link to AD systems and capacity increase

This section discusses the impact THP pre-treatment of sludge will have on the subsequent AD.

THP application for mixed sludge

The thermal hydrolysis process (THP) is applied to sewage sludge as a pre-treatment to improve anaerobic digestion and improve the final sludge quality for sludge management purposes. The conventional application is to apply THP to all sludge proceeding to anaerobic digestion e.g. a blend of PS and WAS subject to THP before AD. If the final sludge produced from AD does not need to be pathogen free, then the option exists to THP WAS only and bypass PS around the THP straight to AD. This would be done to save capital and operational costs using a smaller THP system. However, although the benefit of THP applied to primary sludge is less significant compared to the improvements to WAS, overall digestion performance and loading with WAS only will not be as great as having mixed sludge undergoing THP (Zikakis *et al.*, 2019). In the case where sludge is intended to go to land application a major requirement is to have pathogen free sludge, and THP pre-treatment of all sludge fed to anaerobic digestion ensures this.

Reduced sludge viscosity

THP lowers the viscosity of sludge which allows a higher solids feed concentration and increased digester loading rate while still being able to mix digesters effectively. Liu *et al* (2012) investigated the effects of THP on WAS at 175°C and 60min retention followed by anaerobic digestion. It was reported that the viscosity of WAS reduces from 13500 to 1658 mPas when undergoing thermal hydrolysis. However, this was done by heat only and no steam explosion was created, which would further change the properties of sludge, most likely improving viscosity reduction further. Higgins *et al* (2017) observed that this viscosity reduction improves as THP operating temperatures increases from 130°C to 170°C which allowed for a significantly improved solids loading rate to anaerobic digestion, and therefore sludge throughput. The ideal THP temperature is 165ºC to maximise benefits of the technology. The reduction in viscosity is also likely to contribute to increased biodegradability of particulate COD due to easier mixing of digester contents (Jeong *et al.*, 2019b). In comparison to low solid anaerobic digestion, high solid anaerobic digestion (TS > 10%) is more attractive because of the relatively smaller reactor volumes, lower energy requirements for heating, less material handling. The reduction in sludge viscosity is often used in existing digesters to operate at increased capacity due to a higher loading rate. Several full-scale commercial THP digestion applications have successfully operated at up to 12%DS in the same way as conventional digestion operates at 5%DS (Xue *et al.*, 2015).

Increased biogas production

An increase in biogas production is observed from in numerous studies as listed in Table 2-3. Digestion of THP sludge shows for both for both WAS and a mixture of PS and WAS an increases in methane production of 15%-25% over that in conventional digestion. This results in methane production of 290-340$Nm^3CH_4/tonVSS_{fed}$. A review of various THP studies by Korbicaki (2019) found that methane production from WAS after thermal pre-treatment increased by 24%. A similar value of 18-26% increase in methane production was found by Haug (1978). It is therefore a common theme in literature that THP increases specific gas production per mass of VSS fed, within the digester's SRT.

Increased loading rate

A study carried out by Oosterhuis *et al*. (2014) sows that digesters using THP pre-treatment can be operated at a solids loading of up to 2-3 times higher than conventional sludge. Loading rates typically up to 5-6$kgVSS/m^3/d$ can be achieved, with full scale commercial plants operating at up to 7$kgVSSm^3/d$ (Higgins *et al.*, 2017) at a SRT of 15-20days, with one plant operating up to 10$kgVSS/m^3/d$ (Pook *et al.*, 2013) . This loading can be applied at a sludge feed dry solids of 9-12%. Further, a 62% greater volatile solids reduction can be achieved in anaerobic digestion with hydrolysis pre-treatment per mass unit of WAS. This means retrofitting of THP to an existing installation can comfortably double throughput without having to build any new digesters while increasing specific gas production per tonne of product fed to digestion.

While operating at an increased loading rate it is suggested that anaerobic digesters following thermal hydrolysis are optimized at a retention time of 10-12 days, as by then approximately 95% of biogas potential of conventional digestion at 20 days can be realised (Xue *et al.*, 2015). Longer retention times encourage protein degradation which increases ammonia, alkalinity and pH, and do not result in a statistically significant increase in biogas production (Barber, 2016).

In a study done to review the upper limits of organic loading rate by Pook *et al* (2013) organic loading rates of up to 9$kgVSS/m^3/d$ were applied with a 6 day sludge age, achieving a volatile solids destruction of 45% and gas production of 313$Nm^3/tonVSS$ of methane and total biogas production of 455$Nm^3/tonVSS$. This was a full scale THP digestion facility in Chertsey, UK, operating at 24tonTSS/day throughput.

Increased volatile solids reduction

THP has been found to increase volatile solids reduction (VSR) from that of conventional digestion. Where conventional digestion for WAS achieves around 32% VSR, THP typically achieves 48% to 54% (See Table 2-3). This makes THP a good technology to assist in meeting the criteria required for sludge stabilisation as discussed in Section 2.3.1.

Increased nutrient concentration – ammonia and phosphorous

Due to the higher feed concentrations of solids the THP digestion process results in higher concentrations of dissolved orthophosphate (OP) and free and saline ammonia (FSA). A study by Figdore *et al* (2011) on the treatment of side-stream tested FSA levels of around 2200mgN/l while another by study by Zaoli Gu (2018) found FSA levels up to 2500mg/l. Barber (2016) reported that FSA could reach as high as 3500mg/l in digestion with 300mg/l free ammonia without any inhibition noted in full scale THP facilities. As summarised by Flores-Alsina *et al* (2021) THP shifts bacterial population in the AD towards acetate oxidizers instead of acetoclastic methanogens. This produces more hydrogen and carbon dioxide, shifting AD population towards hydrogenotrophic archea. This allows operation at these higher FSA concentrations of 2500-3500gFSA/l where conventional digestion typically operates around 500-1500mg/l. While THP increases the FSA concentration in pre-treated sludge before it is fed to AD, it has been found the FSA concentration in the AD is mostly governed by the organic loading on the digester. It has been suggested to operate THP under 170^{o}C to reduce the possibility of FSA inhibition in the AD (Wilson and Novak, 2009). Depending on feed sludge properties dilution water is added to the hydrolysed sludge after THP to control the AD feed loading and resulting FSA concentration (Barber, 2016).

Duan *et al* (2012) reported that during high solids digestion when free ammonia nitrogen (FAN) in the digester exceeds 400mg/l moderate inhibition may start and above 600mg/l inhibition of the digestion process will occur. The digestion of sludge mixture of equal parts PS and WAS which has undergone THP is expected to have a free ammonia concentration of under 200mg/l while the same mixture of untreated sludge in conventional digestion would have around 25mg/l free ammonia. This is for a THP digester FSA of 2400-3500mg/l (Barber, 2016). Higgins *et al* (Higgins *et al.*, 2017) found the FAN to be 134mg/l when the FSA was 2650ng/l at pH 7.68. Knowing the total free and saline ammonia (FSA), temperature and pH of the digestion the FAN concentration can be estimated by the below equation from Emerson *et al* (1975) .

$$FAN = \frac{17}{14} \cdot FSA \cdot \left(10^{0.09018+\frac{2729.82}{272.16+T}-pH} + 1\right)^{-1} \qquad \text{mg/l (2-2)}$$

Where,

- FSA is the free and saline concentration of ammonia, mg/l
- pH is the digester pH
- T is the digester temperature in ^{o}C

Siciliano *et al* (2020) noted during digestion dissolved phosphorous can reach up to 800mg/l and Kumari *et al* (2019) reported phosphorous in AD effluent can be as high as 3000mg//l. These high concentrations then return nutrients back to the adjacent WWTW increasing load on the AS reactors. Han *et al* (2017) reported that for the digestion of NDBEPR WAS polyphosphate contained in PA's is released as phosphate during THP, and most of the dissolved phosphorous in the subsequent anaerobic digester liquor is generated during the upstream THP process. It is common that some form of side stream treatment is required to reduce both the N and P nutrients loads before dewatering liquor from digestion is returned to the WWTW.

2.7.4. Benefits for sludge management

Besides the benefit of capacity increase being able to process more sludge with the same digester volume, there are various other benefits to using THP as a pre-treatment unit process.

Improved dewaterability

The final sludge produced from anaerobic digestion with thermal treatment improves in dewaterability. The disintegration of sludge cell walls due to THP allows the final sludge from anaerobic digestion to dewater to above 30% dry solids which is a significant improvement over the 22% typically achieved in sludge from conventional digestion (Higgins *et al.*, 2017). This will reduce transport costs of the final product as less moisture is carried in the sludge, therefore reducing its overall weight. This is expected to bring both financial benefits and environmental benefits.

To achieve 30% dryness in THP cake around 15kg polyelectrolyte per tonne of dry solids is required in the final dewatering step (Oosterhuis *et al.*, 2014) compared to dewatering conventional sludge requiting around 5-10 kg polyelectrolyte per tonne dry solids (Slim, Devey and Vail, 1984; Saveyn *et al.*, 2005; Wei *et al.*, 2018) . However, when considering all operational costs in the THP case the thickening pre-dewatering step prior to THP must also be considered, where typically 3.5kg per tonne of dry solids polyelectrolyte usage is sufficient (Higgins *et al.*, 2017).

Pathogen reduction and stabilisation

The combination of thermal hydrolysis pre-treatment followed by anaerobic digestion ensures a pathogen free final sludge is generated (Perez-Elvira, Fdz-Polanco and Fdz-Polanco, 2010). The complete pathogen kill experienced during the high temperature thermal hydrolysis process ensures complete elimination of Faecal Coliform and Ascaris that will meet South African guidelines for Class A microbial requirements. Studies by independents (Higgins *et al.*, 2008) have also concluded that there is no Re-Occurrence and Sudden Increase (ROSI) in pathogen presence post THP treatment of the final dewatered sludge .The sludge can thus meet the designation criteria for Class A. As the sludge is also stabilised by anaerobic digestion it will meet the designation criteria of Class 1 in the regulatory sludge classification system, discussed in section 2.3. This means that as long as the sludge’s heavy metal content stays in within limits then the sludge can be applied freely to land. Therefore, THP positively improves sludge disposal options.

3. RESEARCH METHODS

Steady-state anaerobic digestion mass balance models (Sötemann *et al*. 2005; Ikumi *et al*., 2015) were used and tailored to suit the anaerobic digestion process at Cape Flats in order to develop a deeper qualitative and quantitative understanding of the process and to allow for a basis of comparative technical evaluation between the AD sludge treatment options. Models were developed for each of the two cases and results compared:

(i) the case where conventional mesophilic anaerobic digestion (AD) is employed without thermal hydrolysis pre-treatment (THP), called “convetional digestion”
(ii) the case where the sludge is pre-treated using THP followed by mesophilic AD, called “THP digestion.”

The two cases are described in more detail in section 3.1.1and 3.1.2. The comparison considered both the economic and environmental aspects on the basis of formulated performance indices described by De Ketele *et al*. (2018) along with additional evaluative methods developed in this study to add depth to this research.

3.1. CASE STUDY – CAPE FLATS WWTW REGIONAL SLUDGE ANAEROBIC DIGESTION FACILITY

The Cape Flats WWTW has three identical operating anaerobic digesters. These are currently used to treat PS and WAS generated at the Cape Flats WWTW. The anaerobic digestion facility has limited spare capacity to import some sludge from other WWTW’s. However, the site has previously been considered by the City of Cape Town to be run as a regional sludge processing facility where significantly more quantities of sludge from surrounding WWTW’s would be imported to join that produced at Cape Flats. Retrofitting THP would increase the capacity of the existing installation and allow for more sludge to be imported, thus maximising digester usage and diversion from landfill.

This study included a case study of the Cape Flats digestion facility. This was considered useful as the Cape Flats anaerobic digesters have been in operation for several decades and are due for an upgrade soon. The City plans to spend capital maintaining and upgrading this facility and this case study will provide useful insight into economic and environmental benefits.

Further, THP has not previously been installed in South Africa yet. Thus, providing a case study with a South Africa context will be useful to any parties considering its application.

Cape Town currently has limited municipal sewage sludge digester capacity and increased sludge treatment capacity in the form of AD would benefit the city. A possible solution is to increase the capacity of the existing installation at the Cape Flats WWTW. This would allow the creation of a regional facility where the increased capacity would allow sludge to be imported from some surrounding WWTW’s. Increasing capacity through applying an advanced digestion pre-treatment process (such as THP) does not require building any further digestion facilities. Other associated infrastructure such as pipelines, roads, mixing pumps, buildings and electrical infrastructure also exist on the site. All this infrastructure would have had to otherwise be constructed if one had to consider building a regional facility on new site, and therefore using existing infrastructure will save overall capital costs and construction time for the project.

Due to the increasing sludge production from the Cape Flats WWTW the digestion capacity would benefit from an expansion in capacity if it were to be used as a regional site. This could take the form of building additional digesters or modifying the existing process to increase its throughput. As discussed in the literature review THP can increase throughput and thus this case study will be useful in determining what capacity the digestion facility could be uprated to by using THP. As the site has limited space and is positioned in an ecologically sensitive area, THP makes sense as a means to increase capacity, as it will take up significantly less space than building additional digesters (Barber, 2016). THP also makes sense to this case as the pathogen kill effect on the final sludge means it would no longer need to be disposed of in landfill but could be distributed directly to agriculture as a Class A1a biofertilizer.

One of the challenges of a digestion facility is the generation of side stream nutrient loads which require treatment. These will increase if the site were to expand to a regional sludge processing facility. The Cape Flats WWTW adjacent to the digesters is the largest in Cape Town at 200Ml/d treatment capacity and is running at 70-80% treatment capacity (Western Cape Government, 2012). This provides some spare capacity to help treat the nutrient-rich side streams generated from the digestion process, thus making the site suitable for the overall processes under investigation in this research. However, it may be likely that additional nutrient treatment of the digestate dewatering liquor will be required from the increased nutrients loads due to imports. The site also serves as a source of treated effluent which is required for various unit processes in each design case e.g. polyelectrolyte solution make-up, dewatering equipment washwater, etc.

3.1.1. Conventional digestion case

The existing facility has three identical CSTR anaerobic digesters operating at steady-state with a continuous feeding and a flow-through regime. The digesters use mixing pumps for mixing. For specific operating criteria used in this study to model this process, such as sludge age and loading rate, please refer to Section 4.3.

The Cape Flats AD site has an import facility where sludge cake from surrounding WWTW's is trucked in to maximise digester throughput. It must be pointed out the sludge contribution from each source, whether imported or indigenous (generated at the Cape Flats WWTW), is not the focus of the study, but rather the capacity of the anaerobic digesters is the focus. Prior to digestion indigenous sludge is thickened using gravity thickening for PS and dissolved air floatation (DAF) for WAS. The sludge is then fed to AD. The digested material is fed to drying beds or dewatered and trucked offsite for disposal. The offsite disposal method for the conventional digestion case is application to sacrificial land. This is land that has limited agricultural use due to the application of sludge, such as only growing crops for uses other than human consumption e.g. animal feed. Biogas from AD is fed to steam boilers providing heat to the mesophilic digestion process. Any excess biogas is flared. For the purpose of this study it will be assumed that no excess gas is flared, but rather in both the conventional and THP digestion cases excess biogas is used in CHP engines to provide electrical power and heat recovery. Further, phosphorous treatment in the form of struvite precipitation and nitrogen treatment by anammox are included in each case to treat the nutrient loads in the digester side-stream returning to the WWTW (see Section 3.3.6 for a further discussion in this regard). Figure 3-1 gives a schematic of the conventional digestion process flow.

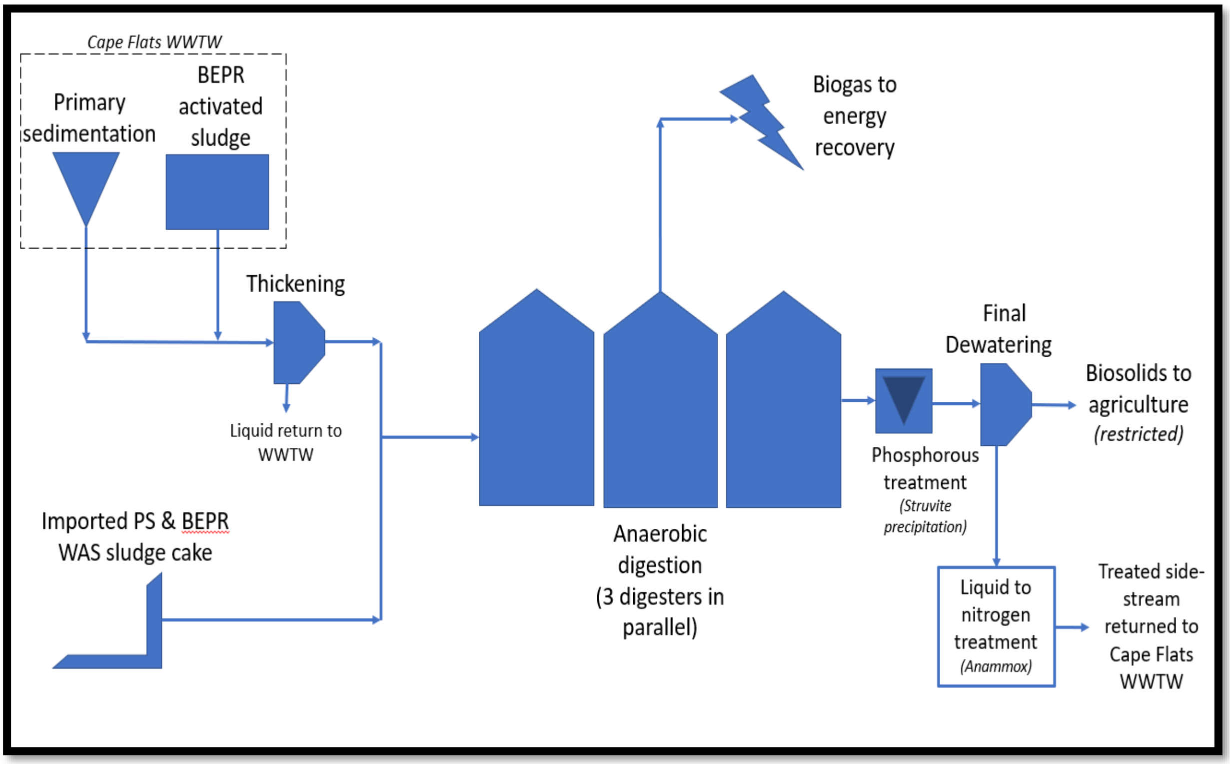

Figure 3-1: Schematic of Cape Flats WWTW's proposed regional sludge beneficiation facility using conventional mesophilic anaerobic digestion, referred to in this study as "conventional digestion."

3.1.2. THP digestion case

The THP digestion case is illustrated in Figure 3-2 below. This case is purely hypothetical and does not yet exist but is a proposed alternative to maintaining the conventional digestion case. The major difference in this case to conventional digestion is that THP sludge pre-treatment is implemented prior to mesophilic anaerobic digestion. As discussed in section 2.7 of the literature review the heat for digestion is provided for by THP and any excess heat is removed by the THP's sludge coolers. Another difference is pre-dewatering is required prior to THP (in addition to gravity/DAF thickening) in order to create a sludge cake. It is assumed for the purposes of this investigation that the total suspended solids (TSS) flux to pre-dewatering is equivalent to the conventional digestion capacity. This can be justified in that the digesters were built for the indigenous sludge production prior to the site being considered for a regional facility.

Besides these differences the unit processes are essentially the same type but are sized for each case's respective throughput (THP digestion is expected to process more sludge and generate more biogas with greater nutrient treatment requirements than conventional digestion and thus has proportionally greater capacity for various unit processes).

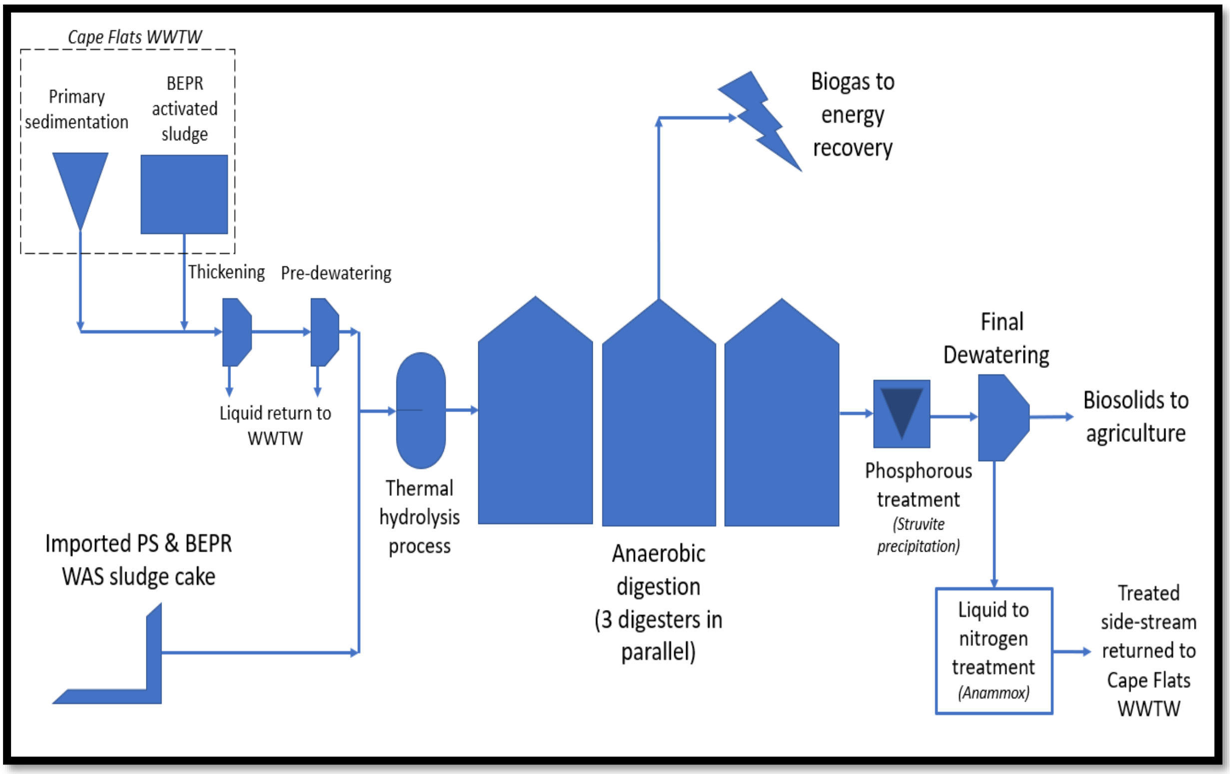

Figure 3-2: Schematic of Cape Flats WWTW's proposed regional sludge beneficiation facility using thermal hydrolysis pre-treatment followed by mesophilic anaerobic digestion, referred to in this study as "THP digestion."

3.2. STEADY-STATE MODEL FOR ANAEROBIC DIGESTION

When modelling wastewater treatment processes one can use either steady-state or dynamic models. Wu (2015) describes that steady-state models are made up of equations which are simple to use, and are typically used for sizing unit processes. The outputs from a steady-state model may also be used as inputs for dynamic models. In contrast, dynamic models are complex and require intricate mathematical processes for solving. Dynamic models are useful for optimising a system.

A suitable steady-state model for anaerobic digestion is that developed by Sotemann (2005). This model uses three elements to model the system. The first is determining the COD removal portion by modelling the rate limiting hydrolysis kinetic process and obtaining a conversion of organics to products. Following this, stoichiometry is used to model bioprocesses and using the conversion from the COD removal in the previous step the quantities of various AD products can be determined. These include alkalinity generated, ammonia released and biogas composition. Developments by Ekama (2009) extended this to include the modelling of the orthophosphate released. Finally, the pH of the digester can be determined using weak acid/base chemistry. This step uses the alkalinity generated and biogas composition together with the carbonate system.

The feed sludge to anaerobic digestion, whether PS or WAS or a combination thereof, is characterised with respect to its C, H, N, O and P components. The sludge's COD is broken into biodegradable and unbiodegradable fractions, as well as VFA concentration. For each sludge type hydrolysis rates are selected from literature, and can be sourced from Ikumi (2014). The modelling process then allows the tracking of various components through the system and also allows for a full characterisation of the digester's effluent, which is useful for modelling downstream processes i.e. P-treatment by struvite precipitation and N-treatment by anammox.

3.3. OPERATIONAL COST INDICES

In research towards energy projects that have an intended payback one of the most challenging aspects is to define the various on-going operating expenses and revenue streams associated with running the plant (Fuller and Petersen, 1995). Thus, to create a more useful outcome this research generated various operational expenses (opex) incurred when running each case, as well as the potential savings that can be generated from the beneficiation process e.g. improved sludge disposal options and sale of electricity.

3.3.1. Overview of cost indices

When comparing the two cases, some major cost items will be compared. The International Water Association (IWA) Benchmark Simulation Modelling Task Group provides a guideline to calculating various operational cost indexes (OCI) that can be applied to maintain consistency when comparing similar processes (Gernaey *et al.*, 2014). These evaluation criteria provide a simple, accurate and objective means for comparison and have been developed specifically for comparing operational strategies.

While various cost indices are available from the publication the following where found to be relevant to this study.

- PE - Pumping energy
- SP - Sludge disposal costs
- ME - Mixing energy
- MP - Methane production
- HE – Heat energy required to heat the anaerobic digester

For completeness the following additional indices have been developed for the purpose of this study and will be included in the valuation in addition to the above:

- PY - Polyelectrolyte consumption for dewatering
- NE – Energy required for treatment of nitrogen
- MG - Magnesium oxide consumption for orthophosphate treatment
- CO_2e - carbon credits saleable generated from emissions reduction

The original operational cost index formula generates a dimensionless value which can be used to compare relative costs. Terms are given a weighting factor according to their cost impact, and terms which present a cost saving are given a negative factor e.g. methane production. The operational cost index relevant with weighting factors given by literature is given by the IWA (Gernaey *et al.*, 2014) is:

$$OCI = PE + 3SP + ME - 6MP + HE^{net}$$

The HE^{net} term refers to the net heating energy required to maintain the temperature in the anaerobic digester taking into account the heat produced by methane.

For the objectives of this study some terms will be modified, and additional terms will be added. The energy recovered from methane in combined heat and power (CHP) engines will be separated into electrical energy (MP_{elec}) and heat energy (MP_{heat}), as these cannot be used interchangeably e.g. heat energy will not supply electrical energy. Note the heating cost term cannot be negative as heat is not used in any other parts of the site. Thus, if the heating term is less than zero, it is rather set to zero.

Indices for pumping and mixing energy given by the IWA are useful in WWTW's where flowrates are high (>10Ml/d) with low pressure drop and tanks are large with low solids concentrations (PS <2%DS and WAS <1%DS). In contrast the flows in an AD plant are relatively small in comparison and solids concentrations are higher when compared a WWTW e.g. thickened solids feed to digestion (4-12%DS), digester effluent (2-7%DS), dewatered sludge cake (15-30%DS). Further, the AD facility comprises of multiple specialist equipment unit process not covered by the IWA indices, such as dewatering equipment (centrifuges and belt presses), polyelectrolyte preparation and dosing, biogas treatment scrubbers, CHP engines and various other process operations. Therefore, the pumping energy (PE) and mixing energy (ME) terms will rather be combined together in one term and called operating energy (OE). OE will be determined using specific energy consumption sourced from literature for similar facilities.

The total operating cost will be calculated in Rand (ZAR) rather than a dimensionless value by applying a relevant unit cost to each term. Further details of how each term is calculated are given in respective sub-sections of Chapter 3, but the overall operational cost (OC) applied for this study was:

$$OC = \Big[\big((OE + NE - MP_{elec}) \times Electrical\ energy\ cost\big) + \Big(\max\big(0; (HE_{fuel} - MP_{heat})\big) \times Heating\ cost\Big) + (SP \times Disposal\ cost) + (PY \times Polyelectrolyte\ cost) + (MG \times Magneisum\ hydroxide\ cost)\Big] - CO_2e \cdot Tax_{CO_2}e$$

Where the following applies, with more detail available in the respective sections that follow.

- OC is the net operational cost in ZAR/annum considering expenses and savings (sale of electricity and carbon credits)
- The electrical cost will be taken as R1.02/kWh (Eskom, 2020).
- The cost of heating will be that of diesel fuel which would be consumed for steam heating if CHP was not available, taken as R11.86 per litre (South African Petroleum Industry Association, 2020).
- Disposal cost in ZAR/tonne sludge
- Polyelectrolyte cost in ZAR/kg dry polyelectrolyte powder
- Magnesium hydroxide cost in ZAR/kg MgO powder
- The carbon tax rate and potential savings from carbon credits are discussed in Section 3.3.8 below.

3.3.2. Operational energy

This section describes how the operational energy (OE) is calculated for the processes shown in process flow diagrams Figure 3-1 and Figure 3-2, and will cover all operations except the downstream nitrogen treatment. During the anaerobic digestion of organics the biogas produced is often used in CHP engines to generate electrical power. However, a fraction of the power generated is consumed by the operation of processing the organic waste into useable energy. This parasitic load varies depending on the type of waste and technologies used and can range from 18 to 54% for plants using anaerobic digestion and CHP (Usack *et al.*, 2019). A study done by Havukainen *et al* (2014) on assessing energy efficiency in biogas production plants assessed parasitic load to be 38% when gas upgrading is added to AD and CHP in for injection into a local gas grid. For AD and THP only the parasitic load was found to 29% of the electrical energy produced by CHP.

More specifically a report on sustainable AD of municipal wastewater sludge by Bachmann (2015) suggests a parasitic operating energy consumption of 0.29kWh/kgTSS when using AD and CHP. A case study carried out by Caposciutti *et al* (2020) on optimising power generation from municipal wastewater sludge using AD and CHP found a similar value of 0.27kWh/kgTSS.

For the purposes of this study the following equation has been derived to determine the amount of electrical energy required for operation.

$$OE = 365 \cdot f_{elec} \cdot Q_i \cdot XT_i \qquad \text{kWh/annum (3-1)}$$

Where,

- OE is the operating electrical energy in kWh/annum
- f_{elec} is the specific energy consumption factor taken as 0.28kWh/kgTSS based the literature review above (in this Section 3.3.2)
- Q_i is the digester feed flowrate in Ml/d
- XT_i is the digester feed TSS concentration in mgTSS/l

3.3.3. Methane production

The methane produced in anaerobic digestion is listed as a negative term in the overall operating cost calculation as it serves as an energy source, rather than an energy consumer, and thereby creating a saving. The methane produced can be used to generate heat and power in a combined heat and power (CHP) biogas engine. Therefore, the energy in the methane gas flow from the digester will be represented as the energy recovered from the CHP engine that can be used in the process.

Electrical energy can be used to power the unit operations at the anaerobic digestion facility and the WWTW, and any surplus electrical power can be exported to the grid. Therefore, all the electrical energy can be recovered.

The heat generated in CHP is made up from high grade heat (+/-400^{o}C) and low-grade heat (+/-85^{o}C). The high grade thermal energy can be used to raise steam required for THP and heating applications, while the low grade heat can be used to pre-heat the boiler feed water to steam generation and provide heat for the downstream side-stream treatment process e.g. nitrogen treatment for anammox or AMTREAT high-rate MLE (Van Hulle *et al.*, 2010; Bungay, 2012).

CHP engines can typically recover the following energy outputs from the energy in the biogas (Clarke Energy, 2020):

Table 3-1: Combined heat and power energy recovery (Clarke Energy, 2020)

Energy recovered	**Symbol**	**Energy recovery** (kW recovered / kW in biogas feed)
Electrical efficiency	ε_{elec}	40%
Thermal efficiency to high grade heat (+/- 400°C)	ε_{HGH}	25%
Thermal efficiency low grade heat (+/-85°C)	ε_{LGH}	25%

The electrical energy recovered MP_{elec} will be calculated by following equation derived for this study, in kWh/annum:

$$MP_{elec} = LHV_{CH_4} \cdot (\varepsilon_{elec}) \cdot \frac{1000 \cdot 365}{3600} \cdot \int_{t_{start}}^{t_{end}} Q_{CH4}\, dt \tag{3-2}$$

The energy recovered as heat energy will be calculated as follows, in kWh/annum:

$$MP_{heat} = LHV_{CH_4} \cdot (\varepsilon_{HGH} + \varepsilon_{LGH}) \cdot \frac{1000 \cdot 365}{3600} \cdot \int_{t_{start}}^{t_{end}} Q_{CH4}\, dt \tag{3-3}$$

Where,

- Q_{CH4} = average methane production, Nm^3/d
- ε_i = energy recovery factors for CHP as presented in Table 3-1
- LHV_{CH4} = lower heating value of methane of form $8.026x10^8$J/kmol taken from Perry (2008) is converted to 33.4 MJ/Nm^3 using the ideal gas law pV=nRT by, given in MJ/Nm^3:

$$LHV_{CH_4} = 8.026 \times 10^8 J \cdot kmol^{-1} \cdot \frac{101325Pa}{8.314J \cdot K^{-1} \cdot mol^{-1} \times 293K} \cdot \frac{kmol}{1000mol} \cdot \frac{MJ}{10^6 J} \tag{3-4}$$

The balance of energy from the combustion of methane in CHP is lost as heat to the atmosphere.

3.3.4. Heating energy

The heating energy is taken as the amount of energy required to heat the sludge from ambient conditions of 20^{o}C to the anaerobic digester operation temperature of 37^{o}C. This will be denoted as $HE_{digester}$. In the case of THP digestion the heating energy required for the hydrolysis process must also be considered. This will be denoted as HE_{THP}. For the THP case the heat required for anaerobic digestion is provided by the upstream THP process. Thus, the greater of $HE_{digester}$ and HE_{THP} will be taken as the heating energy input required in the THP digestion case.

For conventional digestion it is assumed that the sludge reaches the desired temperature within the digester hydraulic retention time and that heating supply is constant. Heat losses to the surroundings via the digester walls are neglected in the calculation, as they would equate to the same value in the comparison of each case, thus cancelling each other out. Return liquors are assumed to lose heat to the environment before being recycled to the adjacent WWTW, and therefore do not return any energy to the system. These assumptions are consistent with similar studies (Gernaey *et al.*, 2014; De Ketele, Davister and Ikumi, 2018). The heating requirement for the conventional digestion case is in kWh/annum is:

$$HE_{digester} = 24 \cdot C_p \cdot (T_e - T_i) \int_{t_{start}}^{t_{end}} \frac{m_i}{3600} \cdot dt \tag{3-5}$$

Where,

- m_i = influent total mass flow rate to the digester, kg/h
- C_p is the specific heat capacity of the suspended solids and water mixture, assumed to be the same as that of water 4.182kJ/kg
- T_e is the digester operating temperature of 37^{o}C
- T_i is the digester influent feed temperature 20^{o}C

In the case of THP the heat supplied for anaerobic digestion to the sludge comes from the upstream steam injection process where steam is directly injected into the THP reactors. The sludge leaving THP has more than the required heat to reach mesophilic anaerobic digestion temperatures, and therefore requires cooling of hydrolysed sludge from 86^{o}C to 37^{o}C prior to feeding to anaerobic digestion (Barber, 2016). This cooling is heat that is lost to the atmosphere, however, it must be included in the overall energy balance as an input in the THP case because it is necessary for reaching the conditions required for the THP process.

The heat required for THP is that required for steam production. This can make use of the lower grade heat (LGH) required to heat the feed water and higher grade heat (HGH) required to overcome latent heat of vaporisation and raise steam. The heat required for THP is derived for this study as, given in kWh/annum:

$$HE_{THP} = \frac{\left(C_p \cdot (T_{THP} - T_i) + H_{vap}\right)}{3600} \cdot S \cdot \int_{t_{start}}^{t_{end}} m_{solids} \cdot dt \tag{3-6}$$

Where,

- m_{solids} is the average TSS flux through the plant in tonnes TSS per hour
- C_p is the specific heat of water 4.182kJ/kg (Sandler, 2006)

- T_{THP} is the THP reactor temperature of 165°C
- T_i is the infeed sludge temperature to THP of 20°C
- H_{vap} is the latent heat of vaporisation of steam at the required pressure for THP of 600kPag which is 2064kJ/kg (Sandler, 2006)
- S is the mass flow of steam, kg steam/tonne dry solids, and is given by Barber (Barber, 2016):

$$S = 10.82 \times \frac{\Delta T}{911 \cdot \varepsilon_{steam}} (134 \times DS^{-1.05}) \qquad \text{Kg steam/ tonne dry solids (3-7)}$$

Where,

- ΔT is the internal temperature difference between the THP reactor with steam pressure of 600kPag and 165°C and the THP flash recovery vessel operating at near atmospheric pressure and 102°C. This gives a ΔT of 63°C.
- DS is the dry solids concentration fed to THP of 17%, as discussed in Section 2.7
- ε_{steam} is steam generation efficiency typically around 85% for a steam system with a biogas boiler (John Thompson, 2020)

Once the heat required for the process is determined it will be compared against the heat re recoverable from CHP. If a deficit exists between the heat from CHP and that required by the digestion process, then additional heat will be sourced from a supplementary diesel boiler. The additional heat ($HE_{fuel,i}$) will be calculated as follows, given in kWh/annum:

$$HE_{fuel,i} = \frac{HE_i - MP_{heat,i}}{\varepsilon_{boiler}} \qquad (3\text{-}8)$$

Where,

- $HE_{fuel,i}$ is the heat energy required from the combustion of fuel in the boiler for each case. For the purposes of this study the fuel source will be diesel.
- ε_{boiler} is the efficiency of a steam boiler, taken as 80% (John Thompson, 2020)
- HE_i is the heat required by the process in each case, either $HE_{digestion}$ for conventional digestion (given by equation (3-5) or HE_{THP} for THP given by (3-6).
- $MP_{heat,i}$ is the heat available from CHP, given by equation (3-3).

Once $HE_{fuel,i}$ is known the cost of diesel fuel consumption ($Q_{fuel,i}$) can be calculated as follows:

$$Q_{fuel,i} = \frac{HE_{fuel,i}}{LHV_{fuel} \cdot 1000} \cdot 3600 \qquad \text{l/annum (3-9)}$$

Where,

- $Q_{fuel,I}$ is the flow of fuel in litres per annum

- $HE_{fuel,I}$ as given by equation (3-8)
- LHV_{fuel} is the lower heating value of the fuel, in this case diesel which is around 36MJ/l (McAllister, Chen and Fernandez-Pello, 2011)

3.3.5. Sludge disposal cost

In the IWA indices the sludge production for disposal is listed with a higher weighting factor of 3 when compared to the other cost contributors, suggesting it is a significant cost item. The following equation is suggested by the IWA (Gernaey *et al.*, 2014) for calculation:

$$SP = M_{TSS}(t_{end}) - M_{TSS}(t_{start}) + \int_{t_{start}}^{t_{end}} TSS_x(t) \cdot Q_x(t) \cdot dt \qquad \text{tonne/annum (3-10)}$$

In the above equations the $M_{TSS}(t_{end})$ and $M_{TSS}(t_{start})$ terms representant the mass of sludge in the system at the beginning and start of the evaluation period. This is to account for any accumulation of sludge. However, in this study these terms will assume to be identical with the system being at steady-state and maintaining a constant hydraulic flow and solids concentration, and therefore they will sum to zero accumulation.

The remaining term will represent the solid flux discharged from the system and will include both VSS and ISS, accounting for all organics, inert solids and any struvite formed. Sludge disposal costs are based on each tonne of dewatered sludge cake (containing TSS+moisture), and not only solids TSS flux, and thus the dryness of the dewatered cake must be accounted for. A factor for dry solids percentage (DS) will be included in the calculation. This will convert the flux of dry solids to a quantity of "wet" dewatered cake for disposal. This also allows one of THP's claimed strengths to be applied in this investigation which is increasing the % dry solids achievable in the final dewatering step after anaerobic digestion. Thus, an applicable equation for this study has been derived as:

$$SP = \frac{1}{DS} \int_{t_{start}}^{t_{end}} \frac{TSS_e(t) \cdot Q_e(t)}{10^6} \cdot dt \qquad \text{tonne/annum (3-11)}$$

Where,

- SP is the sludge production in tonne per day
- DS is the dry solids fraction of the final dewatered sludge cake, listed for each sludge type in Table 3-2
- TSS_e is the sludge dry solids concentration of the digested sludge stream which has passed through p-treatment, mgTSS/l
- Q_e is the flowrate fed to dewatering, which is taken as the same as the digester flow through rate, m^3/d

The following information provided in Table 3-2 below lists the cost for both sludge disposal and transport of sludge from its source to the disposal site. This data will be used to determine disposal volumes and cost calculations for sludge disposal. Typically primary sludge is taken to landfill where WAS and digested sludges are applied to farmlands. The dry solids percentages shown here are in line with what the City of Cape Town typically achieves during sludge dewatering in their plants. The dry solids for THP digested sludge are taken from Section 2.7 of the literature review.

Table 3-2: Sludge disposal costs

Sludge cake type	**Primary**	**WAS**	**AD conventional**	**AD THP**
Dewatered cake dryness, DS (% dry solids w/w)	23%	15%	22%	30%
Disposal cost per tonne	R 812[(a)]	R 81[(b)]	R 81[(b)]	R -
Transport cost per tonne[(c)]	R 53	R 53	R 53	R 53
Total cost per tonne (transport+disposal)	R865	R135	R135	R53
Disposal destination	Landfill	Agriculture (restricted)	Agriculture (restricted)	Agriculture (unrestricted)

The superscripts in Table 3-2 are detailed as:

a) Disposal cost per tonne taken as special hazardous waste as issued by the City of Cape town's solid waste tariff notice (Ladouce, 2020)
b) WAS and digested sludge are considered stable enough for restricted use in some agricultural applications, thus attracting a lower disposal cost than primary sludge.
c) Transport cost data sourced directly form the City of Cape Town gives the cost of sludge transport to be R32/km using 30 tonne truck loads (Sotemann, personal email, June 2020) It is assumed the sludge will be disposed in either Vissershok landfill or surrounding farmlands, all within a 50km radius from its source. This results in a cost of R53 per tonne of sludge.

Sludge is classified as a hazardous waste (Herselman and Snyman, 2009) and thus a hazardous landfill tariff is applicable, as selected for primary sludge above. In some cases, a delisting process can be followed where sludge is tested to determine its impact to its receiving environment against various thresholds, and if deemed acceptable, its classification can be delisted from hazardous waste to general waste. This is possible with WAS and digested sludge, and hence a lower disposal rate is lower for these sludges. However, for primary sludge, even if it is delisted it will still bring a significant cost to dispose of in landfill at R613 per tonne as per general waste charges, as there are currently no alternatives for primary sludge (where delisted WAS and digested sludge can be applied in agriculture through restricted use).

The final biosolids product produced in the case of THP digestion is a Class A1a sludge as described in section 2.3, allowing for unrestricted use on land. This attracts no disposal cost over and above transport cost, and in fact may possibly provide a source of revenue. The potential to generate revenue comes from using the biosolids as a fertilizer instead of using conventional chemical fertilizers, and thus the biosolids if of sufficient quality, become a marketable product that can sold (Merwe-Botha, Borland and Visser, 2019). For the purposes of this study a conservative approach is taken, and the assumption is made that this product will not be sold but would rather be given away for free to the public or used by municipalities for parks and recreation sites e.g. sports fields, parks, communal gardens.

For the disposal cost calculation, the capacities of each digestion case are considered. This is done because each case can process a different amount of sludge i.e. THP digestion can process more sludge than conventional digestion using the same digester volume. The difference in throughput between conventional digestion and THP digestion is thus considered. This difference is sludge that would otherwise required disposal instead of beneficiation via THP digestion. Any reduction in cost for sludge diverted from disposal to THP digestion must be considered as a saving in the case of THP. Alternatively, this can be viewed as a cost in the case of conventional digestion.

The sludge that was not able to be treated by conventional digestion is the difference in capacity between the two digestion cases.

$$SP_{untreated} = MX_{i,THP} - MX_{i,conventional} \qquad \text{tonneTSS/annum (3-12)}$$

Where,

- $SP_{untreated}$ = the quantity of feed sludge untreated and sent for disposal, which is the difference between the feed capacity of THP digestion and conventional digestion
- $MX_{i,THP}$ is the influent solids flux to THP digestion, tonneTSS/annum
- $MX_{i,conventional}$ is the influent solids feed flux to conventional digestion, tonneTSS/annum

The quantity of sludge not treated by conventional digestion will be split into PS and WAS fractions for cost calculation.

$$SP_{untreated,PS} = X_{i,PS} \cdot SP_{untreated} \qquad \text{tonneTSS/annum (3-13)}$$

$$SP_{untreated,WAS} = X_{i,WAS} \cdot SP_{untreated} \qquad \text{tonneTSS/annum (3-14)}$$

Where,

- $SP_{untreated,PS}$ is the quantity for untreated primary sludge sent for disposal, tonneTSS/annum
- $SP_{untreated,WAS}$ is the quantity of untreated WAS sent for disposal, tonneTSS/annum
- $X_{i,PS}$ is the primary sludge feed fraction to digestion in this study, which is given by Table 4-2 as 0.4
- $X_{i,PS}$ is the WAS feed fraction to digestion in this study, which is given by Table 4-2 as 0.6

3.3.6. Nutrient treatment

Dissolved nitrogen and phosphorous are two major nutrients requiring treatment in the retuned anaerobic digester liquor to reduce additional load on the adjacent WWTW. For the purposes of this study a mechanism was developed to compare the extent of nutrient treatment required between and the two cases and the resulting impact on operating costs. These mechanisms are a function of throughput and digester effluent quality and have been created from a literature review of common practices. For the purposes of this study these are considered as side-stream treatments only, where their purpose is to treat digester effluent before it returns to the WWTW. No influent wastewater from the WWTW is treated in these processes. The boundaries of the side-stream treatment are shown within the red squares in Figure 3-3 and Figure 3-4 for conventional digestion and THP digestion respectively.

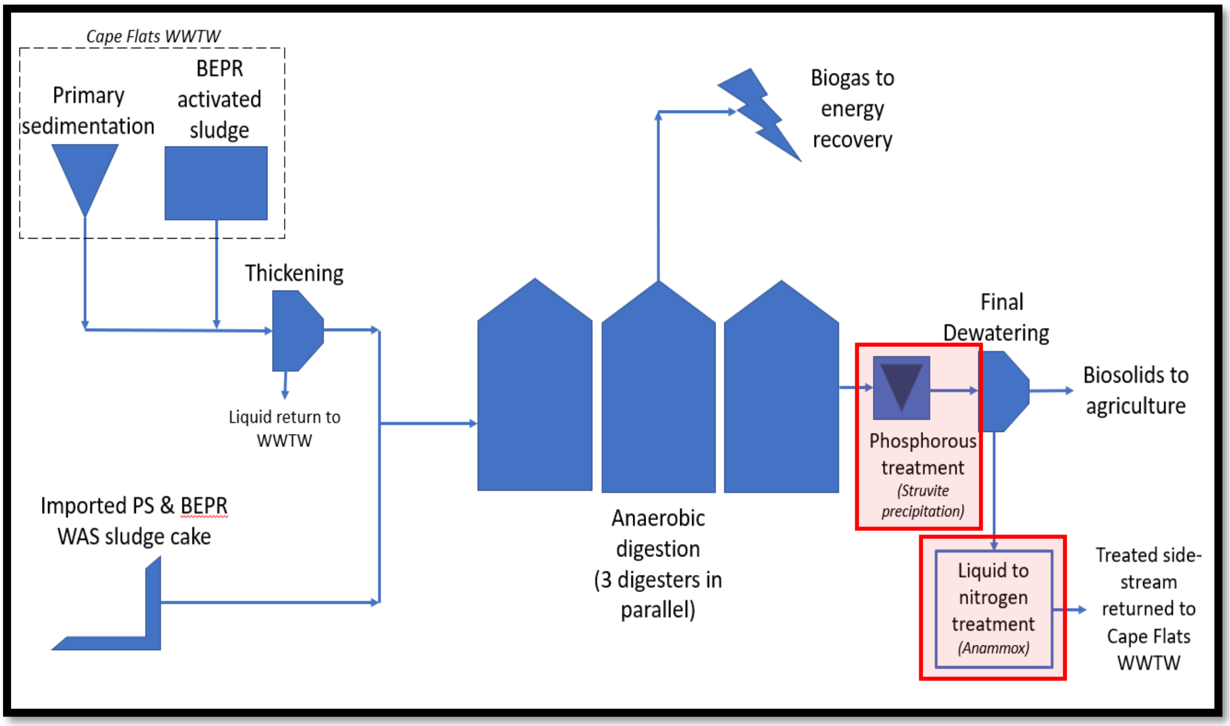

Figure 3-3: Nutrient side-stream treatment for conventional AD model

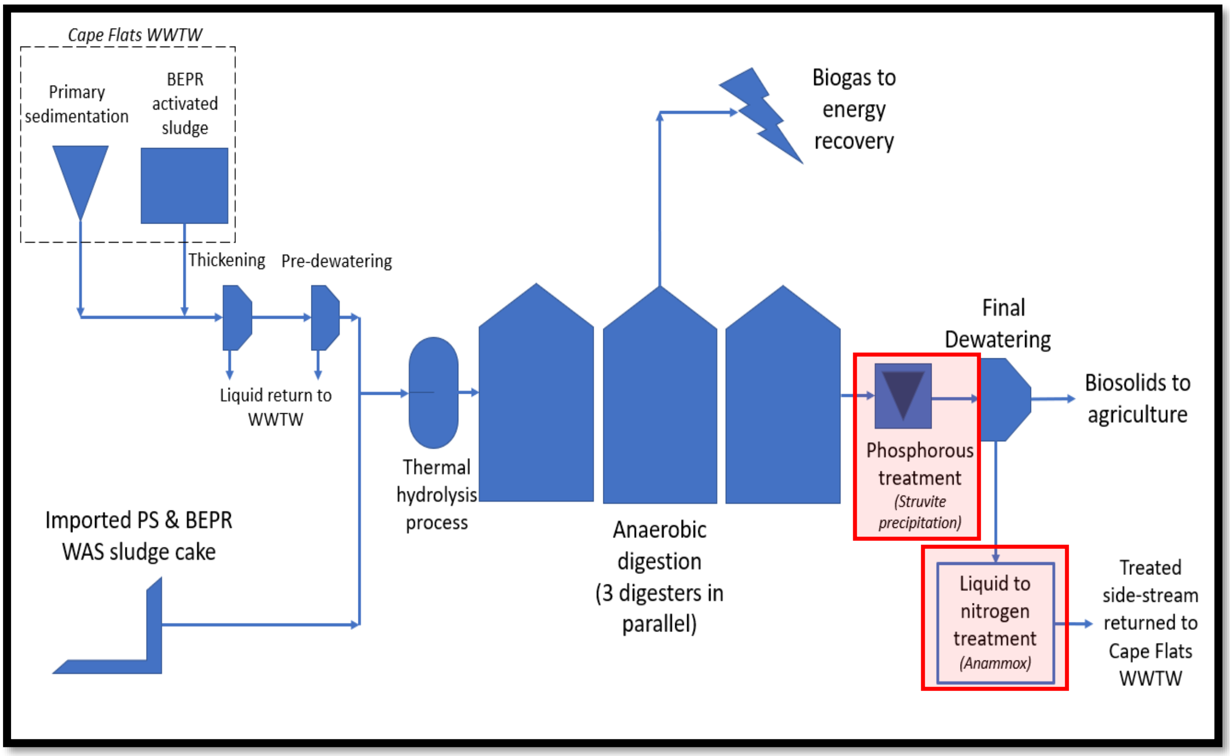

Figure 3-4: Nutrient side-stream treatment for THP digestion model

Phosphorous treatment

The anaerobic digestion of biodegradable organics containing phosphorous results in the release of soluble orthophosphates. The phosphorous in the biodegradable organics has mostly come from biological growth in activated sludge was well as BEPR. Primary sedimentation also contributes small fraction to the phosphorus removal. Phosphorous is removed from the WWTW's wastewater flow via sludge production. Thus, digesting this sludge and returning untreated digester effluent to the WWTW creates a recycle loop and phosphorous is not removed from the system. This will eventually result in a compromised effluent quality and so there is a need for phosphorous treatment of the digester effluent.

A review of commercially available phosphorous treatment technology by Ghosh *et al* (2019) describes the phosphorous treatment of anaerobic digester effluent via struvite precipitation. This is a process of precipitating struvite by dosing an external magnesium source to the effluent from the digester, causing precipitation of struvite in addition to any that may have formed spontaneously inside the anaerobic digester. Phosphorous treatment via struvite precipitation will ensure a high degree of phosphorus recovery while making the phosphorus available as a nutrient in the dewatered sludge. Alternatives include chemical precipitation processes, for example ferric chloride dosing to precipitate metal phosphates. However this has a significantly higher operating cost than struvite recovery (Dewaele, 2015). Further, phosphorous removal via metal precipitates creates a mixed sludge where it is difficult to remove organics from the precipitants, which when applying sludge to land may results in heavy metal contamination (Siciliano *et al.*, 2020).

Performing struvite precipitation under controlled conditions has an operational benefit in ensuring precipitation does not occur in other areas of the plant which can lead to blockages and equipment malfunction in downstream processes. The struvite precipitation process will be done upstream of final dewatering, rather on centrate, to prevent spontaneous struvite precipitation blocking pipes and damaging pumps and dewatering equipment upstream of the dewatering step. This is something that has been known to be an issue at Cape Flats in the past.

In the struvite precipitation process the digestate is fed from the anaerobic digesters to a struvite reactor. The reactor can be a continuous stirred tank configuration or a fluidised bed. The process allows the dissolution of carbon dioxide, which is often assisted with aeration blowers. The use of aeration simultaneously allows for the control of the pH to an ideal range for struvite precipitation without the addition of chemicals which may have been required to increase alkalinity. Energy consumption can be up to 24kWh/kgP; however this is considered exceptionally and around the upper limit of technologies. The ideal pH for precipitation is around 7.6 to 9 (Siciliano *et al.*, 2020). Above this pH ammonium ions speciate to ammonia gas and may become volatile and escape, thereby reducing the availability of ammonium for struvite precipitation.

In the struvite reactor, digestate is combined with a magnesium source in specific ratios according to P concentration and mixed for a sufficient retention time to allow struvite to precipitate within the reactor. The N required is usually in stoichiometric excess. Having the precipitation process immediately after the digester not only mitigates downstream equipment fouling but also benefits from temperature being in the ideal range of 25^oC to 35^oC for struvite precipitation. In the precipitation reactor the struvite precipitate mixes with the solids from anaerobic digestion and the mixture is pumped to the dewatering plant. The dewatered sludge is a mixture of organics from anaerobic digestion and struvite. This mixture is suitable for land application as both the organics and struvite provide essential nutrients for plant growth. The liquid fraction from sludge dewatering, now only rich in nitrogen, is then pumped downstream to further treatment (e.g. anammox).

Another review of P-treatment by struvite precipitation by Siciliano *et al* (2020) discussed recoveries of around 85% of P are common. Magnesium sources regularly used in full scale operation include magnesium chloride ($MgCl_2$), magnesium oxide (MgO) and magnesium metal (Mg), and to a lesser extent, various emerging low-cost options (magnesite mineral rock, seawater bittern, incineration ash). For this study it was decided to use MgO as a magnesium source. Although it is not as soluble as $MgCl_2$, MgO is a popular reactant as it helps increase solution pH without adding salinity to the water. This is useful as struvite precipitation is highly dependent on pH and ionic strength (Loewenthal, Kornmuller and van Heerden, 1994). Further, magnesium sources contribute over 75% of the operating cost in struvite precipitation plants and MgO is lower cost option than $MgCl_2$. The MgO is delivered to site as a powder and made up into a solution for dosing into the struvite reactor.

The dosage ratio of Mg:P is another important variable. Both reviews suggested the ideal ratio lies between an Mg:P of 1:1 to 1.3:1. Above this, any further excess of Mg causes magnesium phosphates to occur, thereby reducing the struvite production. Having the Mg:P too low however, allows competing ions to form other compounds. Calcium (Ca) is a major competing ion to Mg and Ca:Mg ratios should be kept less than 0.2:1 to maximise struvite precipitation (Siciliano *et al.*, 2020). With regards to Mg dosing for P removal to form struvite, Mg:P ratios lower than 1.05:1 were found to allow some formation of hydroxyapatite $Ca_{10}(PO_4)_6(OH)_2$ to mix with struvite production. Irrespective of which ratio is selected, it is unavoidable that small amounts of insoluble phosphate salts are formed along with struvite in the precipitation process. For the purposes of this study a ratio of 1.1:1 of Mg:P was selected.

The cost for MgO with a purity of 96% was sourced from local suppliers to be around R10.65/kg. The MgO consumption is calculated from the following formula derived for this study:

$$MG = M_R \cdot \frac{MM_{MgO}}{\varepsilon_{MgO} \cdot 1000} \cdot \int_{t_{start}}^{t_{end}} \frac{P}{MM_P} \cdot Q_e \cdot dt \qquad \text{kg/annum (3-15)}$$

Where,

- MG is mass of magnesium hydroxide used for P-treatment, kg/annum
- M_R is the Mg:P molar ratio of 1.1:1 for dosing
- MM_{MgO} is the molar mass of magnesium hydroxide (MgO), 40g/gmol
- ε_{MgO} is the purity of MgO used, 96%
- P is concentration of orthophosphates in the digester effluent, mg/l
- MM_P is the molar mass of phosphorous, 31g/gmol
- Q_e is digester effluent flowrate, m^3/d

Nitrogen treatment

The organic nitrogen from the biodegradable organics is released as free and saline ammonia (FSA) during anaerobic digestion. This FSA is in the aqueous form and thus most of it passes through the final sludge dewatering step and ends up in the dewatering liquor which is returned to the WWTW. This TKN load in the return liquor is commonly referred to as a side-stream load and ultimately adds to the TKN load on the WWTW.

A successfully used side-stream nitrogen treatment technology is anaerobic ammonium oxidation, also known as anammox. While anammox is still being investigated in a South African context and is yet to be operated, anammox has successfully been implemented in numerous full scale operations globally (Van Hulle *et al.*, 2010). In this process, ammonium is used as an electron donor and nitrite as an electron acceptor. The ammonium and nitrate are converted anaerobically into nitrogen gas, as well as a small portion of nitrate. The use of the application over conventional activated sludge nitrification-denitrification requires almost no organic carbon and aeration requirements are around 60% less. Sludge production is about 90% less, while the anammox process achieves over a 90% reduction in the FSA load returned to the WWTW (Lackner *et al.*, 2014). It is thus a suitable treatment for a high strength FSA stream with low biodegradable organics, as produced by anaerobic digestion.

Average energy requirements for anammox are reported at around 1.2kW/kgN treated (Lackner *et al.*, 2014). Using the nitrogen load leaving each of the digestion cases the energy cost of treating the effluent the impact of treating the nitrogen load can be calculated from the following:

$$NE = Q_e \cdot \left(X_{TKN,BO} + X_{FSA}\right) \cdot P_{AN} \qquad \text{kg/annum (3-16)}$$

Where,

- NE is the energy required to treat the nitrogen load, kWh/annum
- Q_e is the flowrate leaving the digester, m^3/d

- $X_{TKN,BO}$ is the remaining biodegradable organic TKN, kgN/m^3
- X_{FSA} is the FSA concentration in the dewatering liquor, kgN/m^3
- P_{AN} is the power consumption for nitrogen treatment via anammox, 1.2kWh/kgN

3.3.7. Polyelectrolyte usage

The use of polyelectrolyte aids sludge dewatering processes. However, it brings with it a cost that should be considered in the overall operation. In the case of conventional digestion there is a single dewatering step of digested sludge after anaerobic digestion. In the case of THP digestion there are two dewatering steps requiring polyelectrolyte. These are thickening pre-dewatering prior to THP and final dewatering after anaerobic digestion. It is assumed for the purposes of this investigation that the flow to THP pre-dewatering is equivalent to the conventional digestion feed capacity. This can be justified in that the digesters were built for the indigenous sludge production prior to the site being considered for a regional facility.

The polyelectrolyte required for sludge dewatering is stated in section 2.7.4 of the literature review as a specific mass of polyelectrolyte required per mass unit of TSS being dewatered to the desired sludge cake dryness. Thus, in order to determine the amount of polyelectrolyte required the solids flux of the stream being fed to dewatering must be multiplied by the specific polyelectrolyte usage. The following equations 3-17 and 3-18 have been derived for the purpose of this research to determine the polyelectrolyte usage required.

The following equation is used to determine the polyelectrolyte dry powder consumption for the conventional digestion. This is for the dewatering of digested sludge and reported in kg/annum:

$$PY_{conventional} = \frac{pw_{conventional}}{1000} \cdot \int_{t_{start}}^{t_{begin}} Q_e \cdot TSS_e \cdot dt \tag{3-17}$$

Where,

- $pw_{conventional}$ = specific polyelectrolyte powder usage in grams powder per kg dry TSS given in Table 3-3, g/kg
- Q_e is the effluent flowrate from the digester in Ml/d
- TSS_e is the effluent concentration from the digester in mgTSS/l

For the THP digestion case both the thickening and final dewatering steps must be accounted for. The following equation will be used to determine the polyelectrolyte dry powder consumption in kg/annum:

$$\begin{aligned} PY_{THP} = pw_{THP,thickening} \cdot \int_{t_{start}}^{t_{begin}} FX_{i,convetional} \cdot dt + \\ \frac{pw_{THP,final}}{1000} \cdot \int_{t_{start}}^{t_{begin}} Q_e \cdot TSS_e \cdot dt \end{aligned} \tag{3-18}$$

Where,

- $pw_{THP,thickening}$ = specific polyelectrolyte powder usage in the thickening step prior to THP in grams powder per kg dry TSS given in Table 3-3, g/kg

- $pw_{THP,final}$ = specific polyelectrolyte powder usage in the final dewatering step of sludge after THP digestion in grams powder per kg dry TSS given in Table 3-3, g/kg
- Q_e is the effluent flowrate from THP digestion in Ml/d
- TSS_e is the effluent concentration from THP digestion in mgTSS/l
- $FX_{i,convetional}$ is the pre-dewatering feed solids flux equal to the capacity of conventional digestion, kg/d, given by:

$$FX_i = SLR_i \cdot V \qquad \text{kg/annum (3-19)}$$

Where,

- SLR_i is the solids loading rate applied to the anaerobic digester, $kgTSS.m^{-3}.d^{-1}$
- V is the volume of the anaerobic digester, m^3

See Section 4.3 for the more information on solids loading rates (SLR) and reactor volume (V).

The following polyelectrolyte consumption and corresponding sludge dryness will be used in the calculations, as discussed in Section 2.7.4 of the literature review. The cost of polyelectrolyte was sourced from a local supplier.

Table 3-3: Polyelectrolyte usage for dewatering processes

	THP thickening	**THP final dewatering**	**Conventional final dewatering**
Dosage, pw_x, kg powder/kgDS	5	15	8
Sludge cake dryness	17%	30%	22%
Polyelectrolyte cost, R/kg	R40.18/kg	R40.18/kg	R40.18/kg

Organics (VSS) may possibly require a greater polyelectrolyte dosage than inorganics (ISS), which tend to dewater better than organics. A lower dosage of polyelectrolyte may be possible since the sludge has a high final ISS fraction due to struvite precipitation from P-treatment. However, as no literature was found to give the necessary details a conservative approach was taken and it was assumed all components making up TSS (VSS+ISS) require the same amount of polyelectrolyte dose.

3.3.8. Carbon credits

In 2019 the South African government put in place a carbon tax bill. This implements a taxation on greenhouse gas emitters and is based on the mass of carbon dioxide equivalents (CO_2e) released to atmosphere. However, this bill also allows for negative emitters of greenhouse gases to create carbon tax offsets, also known as a carbon credit. A negative emitter can sell carbon credits to a tax paying emitter and in that way reduce the emitters reduces emissions tax that would have had to be paid. The current marginal carbon tax rate is at R127 per tonne of carbon dioxide. However, depending on the industry, an emitter can qualify for various tax rebates to lower the effective tax rate incurred. The function of the rebates is largely to allow companies a period to invest in emissions reduction technology whereafter these rebates will reconsidered after 2022 (National Treasury of South Africa, 2019).

An example of negative emissions is renewable energy. The emission is considered negative as it reduces reliance on fossil fuels and thus reduces the amount of carbon dioxide equivalents that would have been emitted. A form of renewable energy is burning biogas in combined heat and power generation. The selling of carbon credits produced from a renewable energy source must comply with the following:

- be from the production of less than 15MW.
- if generation exceeds 15MW then the sale of power must be less than R1.09kWh.

Diverting degradable organic waste from landfill also creates negative emissions which would have otherwise decomposed to methane. An example of diverting organics from landfill is sending sludge to a regional anaerobic digestion facility. Methane created from these organics is thus no longer is released to atmosphere. For this study it is assumed all methane produced from anaerobic digestion is captured and used for heat and power generation. The CO_2 portion in the emissions from combustion of the recovered gas i.e. CO_2 in the biogas prior to combustion are not considered significant, as the CO_2 emissions are of biogenic origin (IPCC, 2006).

A carbon credit is equivalent to one tonne of carbon dioxide not emitted. Once credits are earned, they can be traded both locally and internationally. AS mentioned previously, the current value of carbon tax is R127 per tonne of carbon dioxide equivalent (CO_2e). Carbon credits are expected to trade at around 15% below the tax rate which prices carbon credits at R108 per CO_2e (Engineering News, 2020).

If the carbon credit creator is a carbon taxpayer then the credits are effectively used internally by that taxpayer to offset their own reductions and reduce their own emissions tax. It cannot be sold to another taxpayer, as this would then result in double accounting of the benefit.

The act mentions that initially most industries will receive up to a 90% rebate. This means that 90% of their emissions will not be taxed, but the balance will still be taxed at the marginal tax rate. There are numerous industries which will need to pay tax, and as long as the purchase price of carbon credits is lower than the marginal tax rate then there will be a market for companies to benefit from buying carbon credits. Emissions producers both locally and internationally can buy carbon credits from a South African project e.g. petroleum refining, steel and iron processing, food and beverage manufacturing, amongst others (as shown in the act).

Before a value can be applied to the carbon credits generated from biogas beneficiation any methane produced first needs to be converted to CO_2e. This is done by applying a Global Warming Potential (GWP) factor. Annexure 1 of the Carbon Tax Act gives the methane GWP as 23 i.e. 1kg of CH_4 is equivalent to 23kg of CO_2, essentially implying methane is 23 times more significant as a greenhouse gas than carbon dioxide.

When considering methane production in landfill the methane passes through an aerated layer and as a result a small fraction of up to 10% is oxidised to carbon dioxide (Towprayoon *et al.*, 2019). As this is a comparison to landfill this is taken into account in determining the net benefit of diverting sludge from degradation in landfill. The following formula will be used to calculate the CO_2e that is a result of sludge no longer decomposing in landfill (and preventing the release of methane to the atmosphere):

$$CO_2e_m = \frac{\left(GWP_{CH_4} \cdot (1 - OX) \cdot \rho_{CH_4}\right)}{1000} \cdot \int_{t_{start}}^{t_{end}} Q_{CH_4} \cdot dt$$ tonCO₂e/annum (3-20)

Where,

- CO_2e_m is the carbon equivalent negative emissions in ton/annum from no longer releasing methane form landfill
- GWP_{CH4} is the Global Warming Potential for methane which is 23
- OX is the oxidation factor for methane oxidised to carbon dioxide that would have occurred in the aerated layers of the landfill, given as 0.1 by the IPCC.
- Q_{CH4} is the flow of methane produced in Nm^3/d
- ρ_{CH4} is the density of methane, kg/Nm^3

$$\rho_{CH_4} = \frac{MM_{CH4} \cdot p}{R \cdot T \cdot 1000}$$

Where,

- MM_{CH4} is the molar mass of methane, 16g/gmol
- p is pressure under normal conditions, 101 325 Pa
- R is the ideal gas constant, $8.314 J.mol^{-1}.K^{-1}$
- T is temperature in degrees Kelvin

Various assumptions are made in the estimation of carbon credits.

- This study will ignore the CO_2 in the biogas which would be the same whether the sludge is decomposed in landfill or if it was subject to treatment in the regional facility.
- Transport for the is the same in both cases as the extra sludge from surrounding WWTW's that would have been diverted from landfill needs to be transported to the regional facility. Therefore, the emissions due to transport will be assumed to cancel out in each case.
- The remaining organic fraction of sludge not treated by anaerobic digestion i.e. the unbiodegradable organics and the residual biodegradable organics would eventually decompose in landfill whether they remained as part of the digested sludge or had gone to landfill in the first place, and thus are not considered in this comparison.

The CO_2e is also recovered from using renewable energy to create power and reduces the carbon emissions incurred by local power utility, Eskom, in power generation. However, the combustion of methane still creates carbon emissions which must be accounted for and is subtracted from the benefit, as shown in the second term below:

$$CO_2e_p = P_{elec} \cdot \frac{EF}{1000} - Q_{CH_4} \cdot \rho_{CH_4} \cdot C \cdot 365$$ kg/annum (3-21)

Where,

- CO_2e_p is the carbon credits in ton/annum saved from the generation of power using renewable energy, equivalent to the CO_2e that would be emitted from coal fired power generation
- Q_{CH4} is the methane flow combusted in CHP, Nm^3/d

- C is the stoichiometric mass ratio for the mass of CO_2 formed per mass of CH_4 combusted, which is 2.75
- EF is the missions factor as given by Eskom (2020) as 1.04kgCO_2e/kWh
- P_{elec} is the electrical power produced from combustion of biogas in CHP engines, kWh/annum, given by:

$$P_{elec} = LHV \cdot (\varepsilon_{elec}) \cdot \frac{1000 \cdot 365}{3600} \cdot \int_{t_{start}}^{t_{end}} Q_{CH4}\, dt \qquad \text{kg/annum (3-22)}$$

Where,

- Q_{CH4} = methane production, Nm^3/d
- ε_{elec} = electrical energy recovery factor for CHP of 40%, as presented in Table 3-1
- LHV = lower heating value of methane of form 8.026×10^8J/kmol taken from Perry (2008) is converted to 33.4 MJ/Nm^3 using the ideal gas law by equation (3-4):

3.4. CAPITAL COST

A financial feasibility study into waste-to-energy projects by Purser (2011) found that when technology unit processes are integrated as part of a system that their financial feasibility may change. Thus, this research not only considers the capital expense (capex) of the core technology components, but also the additional unit processes that are required support the main technology in each case.

A brief economic payback comparison of the two cases was carried out. This looked at the difference in capital investment required in conjunction with savings in operational cost to motivate additional capital investment THP would bring over conventional digestion. The capital cost to construct each plant would be different, as each process requires different unit operations. Further, from the operational costing there is difference in annual operating costs between the two cases, due to not only the differing unit processes, but also the difference in throughput. This difference incurs an operational saving from THP digestion over conventional digestion and this saving was used to estimate a payback that the additional capital in THP investment would bring.

3.4.1. Reference capital costs

A costing study was carried out by the author in addition to this study to assess the scope and costs associated with retrofitting THP technology to the Cape Flats WWTW anaerobic digesters. This was done from 2018 to 2020 at Project Assignments (Pty) Ltd (a multidisciplinary engineering consultancy specialising in the design and project management of process engineering in the wastewater industry). This costing study generated capital costs for each of the two cases described in Section 3.1. During the costing study some of the project deliverables generated for each case included process flow diagrams (PFDs), piping and instrumentation process diagrams (P&IDs), mechanical equipment layouts and general arrangement drawings, electrical cable routings and single line diagrams, civil and structural designs and environmental impact assessments. This allowed a costing to within 15% accuracy to be carried out using proprietary in-house methods and in consultation with equipment specialists. The results of this costing exercise concluding in June 2020 are shown in Table 3-4.

Table 3-4: Capital cost of reference facility

	Conventional digestion	**THP digestion**
Capital cost	R539mil	R861mil

Some of costing in the conventional digestion case assessed the capital required to maintain existing infrastructure and upgrade equipment which may have reached the end of its usable life. In the THP digestion case most of the same costs applied as in conventional digestion but with the inclusion of additional unit processes required to support the THP process. Further, the THP digestion case was sized at a greater throughput and thus unit processes were sized and costed accordingly.

Some of the major unit processes and infrastructure required for each case is listed in the Table 3-5.

Table 3-5: Equipment and plant infrastructure required for each case

Unit Process	**Conventional digestion**	**THP digestion**
Indigenous sludge thickening	✓	✓
Pre-dewatering	-	✓
Polyelectrolyte make-up and dosing	✓	✓
Import sludge facility	✓	✓
Thermal hydrolysis process	-	✓
Anaerobic digester refurbishment	✓	✓
Struvite precipitation (AD effluent P-treatment)	✓	✓
Dewatering equipment & cake silos	✓	✓
Utilities and services	✓	✓
Odour control	✓	✓
Biogas treated (prior to CHP)	✓	✓
Boilers	✓	✓
CHP engines	✓	✓
Electrical infrastructure improvement	✓	✓
Site refurbishments	✓	✓
Capital cost	R539mil	R861mil

3.4.2. Estimation of capital costs

The capital cost for each case was determined from the inputs listed in Table 3-6 and scaled based on the capacity of each case in this study. The capital cost was also adjusted for inflation.

The adjustment for capacity was done using the 0.6 rule as described by Tribe (1986). This approach is suitable for process plants and is thus suitable for this study. The equation below shows how this capacity rule is applied when scaling similar processes.

$$Cost_B = Cost_A \cdot \left(\frac{Capacity_B}{Capacity_A}\right)^n \qquad \text{Rand (3-23)}$$

Where,

- $Cost_B$ is the cost of the new capacity plant, Rand
- $Cost_A$ is the cost of original plant against which a comparison is being made, Rand
- $Capacity_B$ is the capacity of the new plant, kgDS/d
- $Capacity_A$ is the capacity of the base plant against which the comparison is made, kgDS/d
- n is a dimensionless value used to indicate economies of scale, in this case n shall be 0.6

The time value of money will be accounted for by adjusting for inflation. This will use a cost indices method as described by Whitesides (2012) where the capital value is estimated using a ratio of cost indices. For the purposes of this study Consumer Price Indices (CPI) were sourced from Stats SA (Stats SA, 2021) are listed in Table 3-6 and used the equation below:

$$Cost_t = Cost_i \left(\frac{I_t}{I_i}\right) \qquad \text{Rand (3-24)}$$

Where,

- $Cost_i$ is the initial capital cost, Rand
- $Cost_t$ is the capital cost at time "t", Rand
- I_i is the cost index at the time corresponding to the initial capital cost, dimensionless
- I_t is the cost index at time "t", dimensionless number

Combining the above two equations yields the equation used to estimate capital cost in today's value, denoted as $Cost_{Bt}$, derived for this study as:

$$Cost_{Bt} = Cost_A \cdot \left(\frac{I_t}{I_i}\right) \cdot \left(\frac{Capacity_B}{Capacity_A}\right)^n \qquad \text{Rand (3-25)}$$

The table below summarises the capital cost and indices used to make the time related and capacity adjustments.

Table 3-6: Capital cost calculation inputs

	Notation	**Conventional Digestion**	**THP Digestion**
Unit capital cost in 2020	$Cost_A$	R539mil	R861mil
Capacity of reference plant	$Capacity_A$	60 000kgDS/d	143 000kgDS/d
Cost index average for June (Stats SA, 2021)	I_i	114.9	114.9

Cost index Dec 2020 (Stats SA, 2021)	I_t	117.1	117.1
Cost indice for capacity adjustment	n	0.60	0.60

As this is a comparison the focus is to find the additional capital required to upgrade the conventional digestion case to THP digestion at increased capacity. The additional capital required (∆Cost) will be found from the difference between the cost of THP digestion ($Cost_{Bt,THP}$) and the cost of conventional digestion ($Cost_{Bt,Conventioal}$), given by:

$$\Delta Cost = Cost_{Bt,THP} - Cost_{Bt,Conventional} \qquad \text{Rand (3-26)}$$

Conventional digestion is used as a base case onto which additional capital must be spent to create the THP digestion case. This could then be used with the anticipated saving due to THP digestion to estimate a payback for the additional capital.

3.5. DEVELOP ANAEROBIC DIGESTION MODELS

Based on the steady state anaerobic digestion equations developed by Sötemann *et al*. (2005), two parallel anaerobic digestion models will be produced, one for conventional anaerobic digestion and the other for the THP advance digestion, from which sludge outputs can be estimated. This model is useful as it allows an estimation of volatile solids removal and gas production based on retention time and reactor volume for a given sludge feedstock criteria. While this theory does not cover THP digestion specifically, a modified model will be created using the increased hydrolysis and loading rates observed from experimental data found in the literature review. The modifications made to the model to adapt it from conventional digestion to THP digestion are based on the discussions in Section 2.7 of the literature review and are suitable for use as it compares conventional digestion and THP digestion, which is one of the aims of this research. The equations developed by Sötemann (2005) focuses on primary sludge and MLE WAS. However, it does not take into account the digestion of BEPR WAS. Thus, the model will be adapted to suit the digestion of BEPR WAS and the associated likelihood of struvite precipitation. The weak acid base chemistry will also be corrected for the influence of ionic strength, which is especially relevant for high solids digestion such as THP. The details of the model and supporting literature are discussed in further detail in Section 4.

4. MODEL DESCRIPTION

This section explains the theory that was used to generate the steady-state anaerobic digestion models. It also lists the differences between the conventional digestion model and THP digestion model. All equations are listed, and motivation is given to why these are relevant for this research.

4.1. BOUNDARIES FOR EACH MODEL

4.1.1. Conventional digestion

The model developed for the conventional AD case covers the scope shown within the boundaries of the red rectangle in Figure 4-1. This covers the AD process only and excludes the balance of plant.

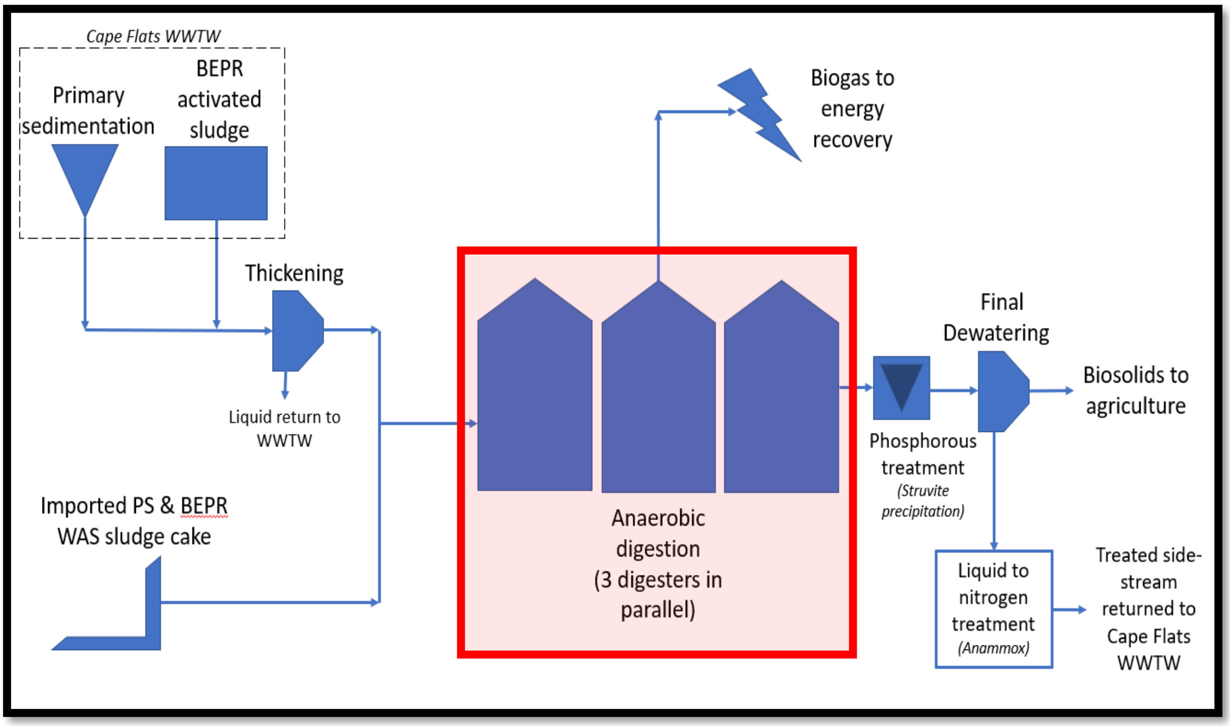

Figure 4-1: Boundaries of conventional AD model

4.1.2. THP digestion

The model for the THP digestion case considers the THP unit process and AD as a single unit process combined in one step. The model covers the scope as shown within the boundaries of the red rectangle in Figure 4-2.

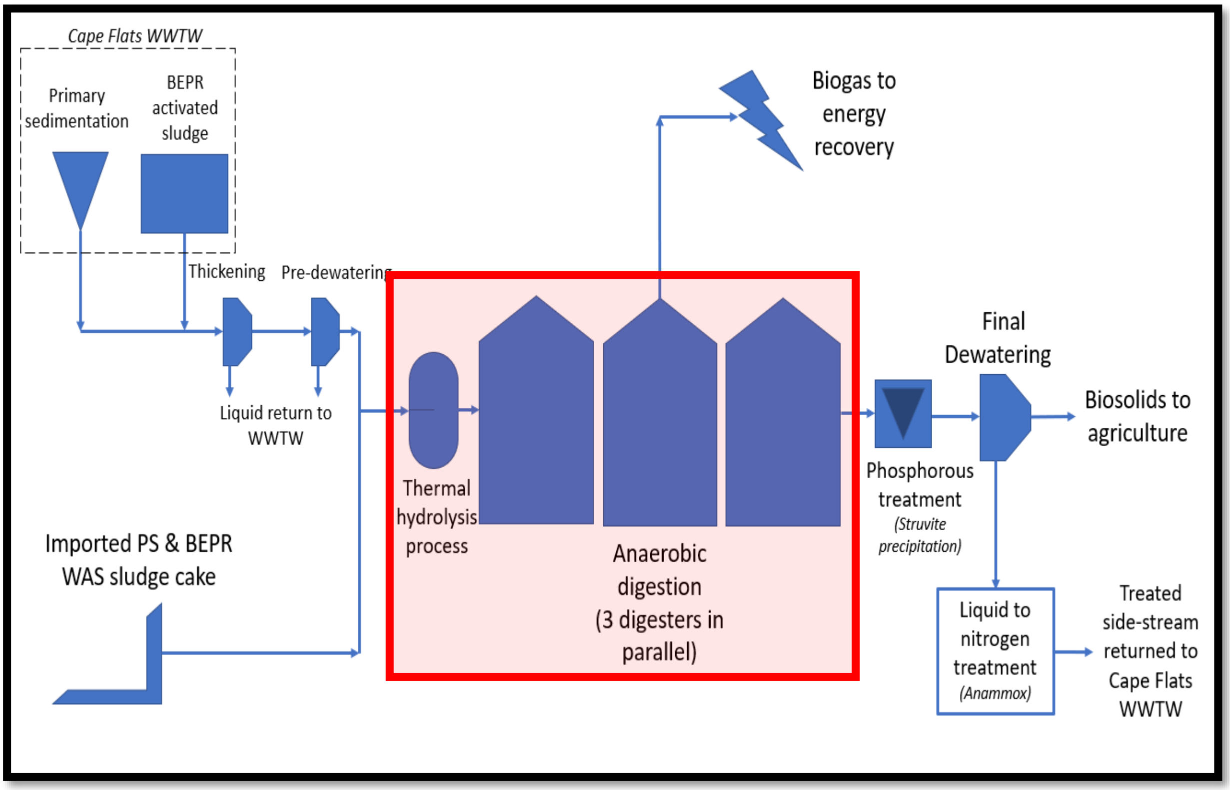

Figure 4-2: Boundaries of THP digestion model

4.2. COMPONENTS

The following components will be tracked through the digester model:

- COD – the influent COD will be split into biodegradable organic (BO) COD and unbiodegradable organic (UO) COD, which accounts for both the soluble and particulate fractions of each.
- VSS – this is made up of particulates of biodegradable and unbiodegradable organics in the feed, and biomass formed during anaerobic digestion.
- ISS – the inorganic suspended solids will simply pass through the digester but are tracked to allow for determination of the total suspended solids (TSS). In addition, polyphosphate (PP) contributing to the inlet ISS but is broken into soluble components, thereby reducing ISS. However, inside the digester some struvite is formed which adds to the ISS. These changes are accounted for in the model.
- TSS – this is the sum of the VSS and ISS described.
- TOC – this will be contained in the influent organics and released through digester kinetic and stoichiometry.
- TKN – this will be contained in the influent organics and released through digester kinetic and stoichiometry.
- FSA – the free and saline ammonia (FSA) portion of the TKN will also be tracked.
- TP – the total phosphorous (TP) fed to digestion will be split into organic bound nitrogen, which is part of the sludge biomass, and polyphosphate chains, which are stored in the active in the active PAO fraction.
- OP – the TP mentioned above will consider the amount of orthophosphate (OP) present.
- Methane – component of the biogas produced

- Carbon dioxide – component of biogas that will be tracked
- Hydrogen ion concentration – this is determined in the form of pH

4.3. PARAMETERS

The following parameters will remain fixed in each model. Some of these inputs will differ for conventional digestion to those used in THP digestion, based on information found from the literature review.

4.3.1. Digester operation

The volume of the digestion space is fixed and is the same in both instances. This is because the case study is investigating existing digesters. This is calculated using the existing digester's dimensions.

Table 4-1: Digester volume

Parameter	Value
Height	20m
Diameter	20m
Volume per digester	6283m^3
Number of identical digesters	3
Total digestion volume, V_{ad}	18850m^3

The three digesters are identical and work as continuous stirred reactors (CSTR). The digesters have an aspect ratio of 1:1 which is ideal for a CSTR type operations.

Table 4-2: Capacity and loading rate

	Conventional Digestion	THP Digestion		Comment
Influent solids flux	60463	153855	kgTSS/d	2.5 times increase
Solids retention time (SRT)	14.7	13.5	days	-
Feed % WAS	60%	60%		-
Feed % PS	40%	40%		-
Feed solids concentration	4.7%	11%	gTSS/l	-
THP loading rate	-	6.2	kgVSS/m^3.d^{-1}	-
Digester loading rate	2.5	5.93	kgVSS/m^3.d^{-1}	2.4 times increase

Table 4-2 shows the operating parameters of each digestion case. The City of Cape Town has many WWTW using NDBEPR AS and a regional facility would most likely receive significant quantities of the NDBEPR WAS. Further, many of the city's WWTW's use primary sedimentation and PS has little option for disposal other than landfill. Thus, both PS and NDBEPR WAS would most likely be fed to a regional facility at Cape Flats and this study therefore includes both PS and NDBEPR WAS in the feed to AD.

As the digesters are of the continuous flow-through type the sludge retention time (SRT) is essentially the same as the hydraulic retention time (HRT). The solids flux capacity was determined by selecting a loading rate and solids feed concentration according to what is typically seen in commercial scale plants. For conventional digestion the sludge feed solids concentrations was set at 47g/l (4.7%DS) which is typical feed solids concentration for conventional digestion (Merwe-Botha, Borland and Visser, 2019) and loading rate at 2.5kgVSS/m^3.d^{-1} as taken from Section 2.4.3 of the literature review. For THP digestion the solids feed concentration was set at 110g/l (11%DS). The loading rate of untreated sludge to THP was 6.2kgVSS/m^3.d^{-1} and after solubilisation (as discussed in section 4.3.4) the applied loading of hydrolysed sludge to the AD was 5.9 kgVSS/m^3/d. This is typical for THP digestion as found in Section 2.7.3 of the literature review.

4.3.2. Sludge mass fractions

The sludge mass fractions for the unbiodegradable and biodegradable organics are assumed to not be altered by THP and thus are kept the same for both conventional and THP digestion. Sludge input data for both PS and WAS for this study was taken from operating WWTW's in the City of Cape Town and is presented in the work done by Ikumi (2011). PS data was from samples from the Athlone WWTW. This is suitable as Athlone WWTW is one of the city's largest WWTW and would most likely be a sludge contributor to the Cape Flats regional facility.

The WAS data was taken from the laboratory treatment of settled wastewater from Mitchells Plain WWTW treated in a nitrification-denitrification biological excess phosphorous removal (NDBEPR) activated sludge (AS) process. A 10-day sludge age was used and excess P and acetate were dosed to enhance BEPR. The process achieved a removal of 36mgP/l from the wastewater. This resulted in good growth of PAO's containing PP in the WAS. Considering that typical settled municipal wastewater influent can have 6-15mgP/l (Ekama *et al.*, 1984) the AS run by Ikumi (2011) is representative of good NDBEPR. This WAS input data was thus deemed suitable for the current research investigating the AD of NDBEPR WAS containing PP. Further, Mitchells Plain WWTW runs a full-scale NDBEPR activated sludge UCT process treating settled wastewater and being located near to the Cape Flats WWTW site would most likely contribute sludge to the regional sludge treatment facility. Many of the city's WWTW use NDBEPR AS and thus this data would be representative of the WAS treated at the regional facility.

The VSS/TSS of the feed sludge is taken from Ikumi (2011) will be as shown in Table 4-3 below.

Table 4-3: VSS/TSS fractions of raw sludge (prior to THP pre-treatment)

	PS	**WAS**
VSS/TSS	0.81	0.74

The VSS component of the sludge is made up from biodegradable and unbiodegradable portions. For the steady-state model used in this investigation it is assumed the VSS term represents the soluble and particulate fractions in the feed. This can be motivated by the fact that the sludge in conventional digestion is thickened, and the soluble fraction is negligible for the purposes of this exercise. For the sludge fed to THP the pre-dewatering step thickens the sludge even further to a dewatered sludge cake, thus further reducing the relative contribution of soluble substances. For all intents and purposes the VSS fraction here can be viewed as the organic fraction of the sludge and unbiodegradable particulate organics (UPO) will be referred to as total unbiodegradable organics (UO) and biodegradable particulate organics (BPO) will be referred to as total biodegradable organics (BO).

Table 4-4: Mass fractions of sludge components

		PS		**WAS**		
		UO	BO	UO	BO	
COD content of organics	fcv	1.510	1.479	1.451	1.435	gCOD/g mass
Mass fractions						
Carbon content of organics	fc	0.522	0.466	0.517	0.514	gC/g mass
Hydrogen content of organics	fh	0.057	0.085	0.061	0.062	gH/g mass
Oxygen content of organics	fo	0.306	0.404	0.372	0.247	gO/g mass
Nitrogen content of organics	fn	0.061	0.033	0.036	0.138	gN/g mass
Phosphorous content of organics	fp	0.054	0.012	0.013	0.040	gP/g mass
Total		1.000	1.000	1.000	1.000	

The sludge mass fraction shown in Table 4-4 were taken from Ikumi (2011). The WAS mass fractions are that of the biomass and are taken to be the same for the OHO‘s and PAO’s.

4.3.3. Polyphosphate content of WAS

The WAS generated from NDBEPR AS will contain a portion of PAO’s in addition to OHO’s. The active fraction of the PAO’s will contain polyphosphate chains and thus contribute more phosphorous to the overall phosphorous release during AD (Wentzel *et al.*, 1990).

Table 4-5: P content make-up of raw WAS TSS (Ikumi, 2011)

PAO P			
Organically bound P in PAO biomass	f_P	4.0%	gP/gPAO
Polyphosphate P in PAO's	f_{XBGPP}	6.3%	gP/gPAO
Total P of active PAO's (f_P+f_{XBGPP})	f_{XBGP}	10.3%	gP/gPAO
WAS P			
P content of active WAS VSS	-	8.4%	gP/gVSS
P content of WAS VSS	-	6.0%	gP/gVSS
P content of WAS TSS	-	4.5%	gP/gTSS

Table 4-6 shows the data given by Ikumi (2011) the mass fractions for polyphosphate:

Table 4-6: Polyphosphate elemental fractions (Ikumi, 2011)

Molar fractions of PP			
Magnesium content of PP	c	0.30	molMg/molPP
Potassium content of PP	d	0.33	molK/molPP
Calcium content in of PP	e	0.03	molCa/molPP
PP content of PAO			
Molar ration of PP in PAO		0.23	molPP/molPAO

The value of q of 0.23molPP/molPAO results in a total P content (both biomass and PP) in the PAO's of around 10.3% gP/gPOAVSS.

4.3.4. Sludge Biodegradability

The modelling approach used for this study does not discriminate between soluble and particulate biodegradable organics and lumps all biodegradable organics in the feed together as biodegradable organics. As discussed in section 4.3.2, the sludge fed to digestion will be thickened. This will result in a relatively low contribution of the soluble biodegradable organics to the COD in the AD feed (<1%). The contribution of biodegradable matter will be mostly from particulate COD. This is even more significant in the case of THP where pre-dewatering of sludge to a cake is done prior to input to the process as described in Section 2.7.

For WAS the active fraction is considered the biodegradable fraction. This is the fraction of the WAS that is made up of active ordinary heterotrophic organisms (OHO's) and active polyphosphate accumulating organisms (PAO's). These active fractions exclude the other VSS components, such as endogenous residue and unbiodegradable particulate organics (UPO) (Wentzel and Ekama, 1997). Table 4-7 shows the active fractions taken from Ikumi

(2011) of OHO's 14% and PAO's 33% giving a total combined active fraction in the WAS of 47%.

Table 4-7: WAS active fractions

% Active OHO's in WAS VSS	14%
% Active PAO's in WAS VSS	33%
% Total active WAS VSS	47%

Table 4-8 shows the biodegradable ($f_{S'bs}$) and unbiodegradable fractions ($f_{S'us}$) of the sludge fed to digestion. The raw sludge data are sourced form Ikumi (2011) and for THP these have been modified due to the effects of THP pre-treatment. THP increases the biodegradability of the WAS by 11%-20%, as discussed in the Section 2.7.2 of the literature review. For the purposes of this study an increase in WAS biodegradability of 15% was applied. This was done via a mass balance by reducing the unbiodegradable fraction and allocating that reduction to increasing the biodegradable fraction, all while the keeping the total system mass the same.

Table 4-8: Unbiodegradable fraction of feed sludges

Digestion	**Conventional digestion**		**THP digestion**	
Unbiodegradable COD fraction ($f_{S'u}$)	0.30	0.54	0.31	0.46
Biodegradable COD fraction ($f_{S'b}$)	0.70	0.46	0.69	0.54

For primary sludge a value of 0.30 was taken as measured by Ikumi (2011). This biodegradability will be assumed to represent all primary sludge within the City fed to the regional digestion facility. The decrease in PS biodegradability is due to the effects of THP creating a small fraction of unbiodegradable soluble organics (discussed further in section 4.5.1). Although THP increase solubilisation of PS (see section 2.7.2) there is limited literature for the increase in biodegradability of PS due to THP. Therefore, this study assumed that the biodegradability of PS did not increase, and the only effect considered is the conversion of some BO to a small fraction of USO. However, other factors such as solubilisation were assumed allow for the increase in loading rate without overloading the AD.

4.3.5. General assumptions

The following assumptions will be made:

- The three anaerobic digesters operating in parallel act as one digester with the same volume as the three combined.
- No water leaves in the biogas (or all water is condensed from biogas and returned to digester to leave via digestate).
- Effects of precipitation are ignored inside the digester other than struvite precipitation.
- Effects of sulphur are ignored.

4.4. VARIABLES

The following variables were calculated from each model:

- Effluent COD: Unbiodegradable and biodegradable
- Effluent AD biomass concentration
- Effluent solids: TSS, VSS and ISS
- Effluent TKN: FSA, biodegradable TKN and unbiodegradable TKN
- Effluent TP: OP, biodegradable TP and unbiodegradable TP
- Polyphosphate released and subsequent Mg, K and Ca effluent concentrations
- Ionic product and activity coefficients
- Struvite precipitated inside digester: Solubility product, ionic product and quantity of struvite precipitated
- pH
- Alkalinity generated
- Gas production: Quantity and composition

4.5. PROCESSES: KINETIC AND STOICHIOMETRIC EQUATIONS

4.5.1. Chemical oxygen demand

The influent sludge COD (S_{ti}) consists of unbiodegradable particulate organics (S_{upi}) and unbiodegradable soluble organics (S_{usi}), biodegradable particulate (S_{bpi}) and fermentable readily biodegradable organics (S_{bsfi}) and volatile fatty acids (S_{VFAi}). This can be represented by the following equation:

$$S_{ti} = S_{upi} + S_{usi} + S_{bpi} + S_{bsfi} + S_{VFAi} \qquad \text{mgCOD/l (4-1)}$$

For the purposes of this modelling exercise and in limiting the scope of this study, the anaerobic digestion model will assume that all particulate and soluble biodegradable organics as one combined entity. Thus, the biodegradable terms will be summed together and represented as one term, and the influent COD is represented as:

$$S_{ti} = S_{upi} + S_{usi} + S_{bi} \qquad \text{mgCOD/l (4-2)}$$

After digestion the effluent COD (S_{te}) consists of unbiodegradable particulates (S_{upe}) and unbiodegradable soluble (S_{use}), residual biodegradable organics (S_{bpe}) and the biomass of microorganisms (Z_{AD}).

Unbiodegradable soluble organics

As discussed in Section 2.7.2 of the literature review THP tends to create a small fraction of unbiodegradable soluble organics (USO). This has been accounted for in the model and this study assumed some of the biodegradable particulate organics (BPO) are converted to USO. The extent of this is given as 11kgUSO formed per ton dry solids in digester effluent (Oosterhuis *et al.*, 2014; Barber, 2016). The increase in USO was then deducted from the

BPO fed to digestion in order to conserve mass. The increase in USO was calculated as follows.

$$S_{use} = f_{USO,generated} \cdot \frac{XT_e}{1000} \qquad \text{mgCOD/l (4-3)}$$

Where,

- S_{use} is the digester effluent unbiodegradable soluble organics, mgCOD/l
- $f_{USO,generated}$ is the amount of USO generated per ton of digester effluent solids, given as 11kgUSO/tonDS (Oosterhuis *et al.*, 2014; Barber, 2016)
- XT_e is the digester effluent TSS concentration, mgTSS/l

The USO in the feed sludge contributes a small fraction to the overall COD, especially in thickened sludge where the COD is almost entirely from particulates. However, to ensure accurate tracking of COD through the model USO tracking is included. The USO concentration in the feed sludge is assumed to be the same as that found in typical raw wastewater fed to a WWTW. This is because USO moves through all unit processes in a WWTW unchanged. Therefore, the liquid fraction in the sludge fed to the AD will contain the same USO concentration. The following USO characteristics were taken for this study and are typical of influent wastewater at South African WWTW's, taken from Ekama *et al*. (1984). The USO generated during THP is assumed to have the same characteristics (fcv, fc, fn).

Table 4-9: Unbiodegradable soluble organics in feed sludge

COD concentration (mg/l)	47
gCOD/g,mass (fcv)	1.493
gC/g,mass (fc)	0.498
gN/g,mass (fn)	0.036
gP/g,mass (fp)	0.00

Effluent VFA

The COD of the fermentable readily biodegradable organics (S_{bsfe}) and the volatile fatty acids (S_{VFAe}) is assumed to be zero in the effluent in conventional digestion. It is assumed that the digesters in this study are run optimally. This means they are not overloaded with substrate, have good buffer capacity and are run at a pH close to 7. This will allow VFAs to be utilised extensively in the reactor and under steady state conditions and at long sludge age >13days acetoclastic methanogens are able to comfortably keep up with converting VFA's to AD products. Further, THP digesters operate at a higher pH shifting the speciation of acetate more towards dissociated acetate, which can get converted to methane and therefore lower VFA in effluent. In the case of THP digestion the effluent VFA concentration can range from 25-5000mg/l (Zhang *et al.*, 2016). Xue *et al* (2015) performed high solids digestion of THP sludge at 16%DS for 28days and measured a effluent VFA concentration of 29mg/l. Han *et al* (2017) ran digesters on THP sludge with a 20-day SRT at high solids concentration of 10% and measured effluent VFA concentration of 280mg/l. Further, the VFA/alkalinity Ripley ratio (Merwe-Botha, Borland and Visser, 2019) achieved was 0.02 which is significantly below the 0.3 maximum recommended for stable digesters as discussed in Section 2.4.3 of the literature review.

For the purposes of this study, and considering the digestion model, the residual biodegradable COD leaving the digester is assumed to be in particulate form and the soluble COD is considered to be negligible. The effluent COD can be represented by the following equation:

$$S_{te} = S_{upe} + S_{use} + S_{bp} + Z_{AD} \qquad \text{mgCOD/l (4-4)}$$

4.5.2. Hydrolysis kinetics

The steady state model is based on the slowest kinetic rate. For mesophilic anaerobic digestion the slowest process is hydrolysis. In high-solids digestion made possible by THP the hydrolysis rate is still the limiting step (Liu *et al.*, 2016).

Hydrolysis rate and residual biodegradable organics concentration after digestion are expressed using Saturation kinetics. This was chosen over alternatives, such as Monod kinetics. The Monod kinetics is more effective when applied with the assumption that the ratio of the active biomass mediating the process and the bulk liquid are at a ratio that could be calculated based on the substrate concentration in the bulk liquid (works well with soluble components, when the biomass is mixed in the reactor). Saturation kinetics however works better with particulates because ratio of biomass to particulates that biomass use within their active sites is different to the ratio of the biomass to substrate concentration in the bulk liquid. For the purposes of modelling a comparative exercise between conventional MAD and THP+MAD saturation kinetics is deemed to be a more suitable approach than Monod kinetics.

Table 4-10 shows the kinetic constants for anaerobic digestion of each sludge type taken from (Ikumi, Harding and Ekama, 2014). However, for THP digestion the maximum specific growth rate (K_M) was adjusted to cater for the changes caused by THP to the organics fed to AD.

Table 4-10: Kinetic constants

	Conventional digestion	**Conventional digestion**	**THP digestion**	**THP digestion**
Sludge type	WAS	PS	WAS	PS
Maximum specific growth rate (K_m)	1.95 gCOD	5.27 gCOD	3.12gCOD	5.27gCOD
Half saturation coefficient (K_s)	9.11 gCOD/l	7.98 gCOD/l	9.11gCOD/l	7.98gCOD/l

As discussed in Section 2.7 of the literature review, increased solubilisation from THP pre-treatment of WAS increases the rate of substrate usage in AD by 57%-127% when compared to that conventional digestion. Based on the literature findings this study used an increase in the maximum specific growth rate for WAS of 60%. This value was on the lower range of that found in literature and thus is deemed conservative. Although literature was found to show that THP increases solubilisation of PS, limited literature seems to exist on the impact this has on the rate of subsequent digestion. One might expect this to also increase the rate of PS digestion in AD, but to be conservative the rate was assumed to remain the same. However, it is clear THP benefits both PS and WAS to enable a higher digester loading rate and allows more sludge to be processed in the same digester volume, as discussed in Section 2.7.3.

Acidogen biomass yield (Y_{AC}) makes up the majority of biomass growth has the value 0.089gCOD biomass/gCOD hydrolysed. To account for other biomass groups the yield is increased from 0.089 to 0.113gCOD/gCOD hydrolysed (S W Sötemann *et al.*, 2005). These groups of organisms have a similar endogenous respiration rate (b_{AD}) of 0.041/d. The biomass unbiodegradable fraction from biomass death (Y_{ad}) is considered negligible in anaerobic digestion, and is assumed to be zero for the purposes of this modelling exercise.

For saturation kinetics the following equations will apply, taken from Sötemann *et al.* (2005):

Hydrolysis rate:

$$r_h = \frac{K_M \left(\frac{S_{bp}}{Z_{AD}}\right)}{\left[K_S + \left(\frac{S_{bp}}{Z_{AD}}\right)\right]} Z_{AD} \qquad \text{mgCOD/l.d}^{-1} \text{ (4-5)}$$

Residual biodegradable organics concentration:

$$S_b = \frac{S_{bpi}}{1 + \frac{\left[Y_{AD}K_M - \left(\frac{1}{R} + b_{AD}\right)\right][1 + b_{AD}R(1 - Y_{AD})]}{Y_{AD}K_S(\frac{1}{R} + b_{AD})}} \qquad \text{mgCOD/l (4-6)}$$

Considering that the reactor is stable under steady state conditions, various microorganism groups are present in the reactor and utilise the organics and intermediate products to produce the digester end products. In the reactor all microorganisms are consolidated into one term Z_{AD}, determined as follows:

$$Z_{AD} = \frac{Y_{AD}(S_{bpi} - S_{bp})}{1 + b_{AD}R(1 - Y_{ad})} \qquad \text{mgCOD/l (4-7)}$$

Methane production is represented as:

$$S_m = (1 - Y_{AD})Rr_h \qquad \text{mgCOD/l (4-8)}$$

The unbiodegradable particulate organics in the influent of the AD simply pass through the digester and emerge as part of the effluent flow, ultimately leaving with the dewatered sludge cake final product. Similarly, the unbiodegradable soluble organics entering AD pass through the digester unchanged and leave as part of the liquid fraction discharged from the final sludge dewatering process.

Unbiodegradable particulate organics, gCOD/l:

$$S_{upe} = S_{upi} \qquad \text{mgCOD/l (4-9)}$$

Unbiodegradable soluble COD:

$$S_{use} = S_{upi} \qquad \text{mgCOD/l (4-10)}$$

The fraction of hydrolysed COD that is converted to biomass, E, is defined by the following:

$$E = \frac{Y_{AD}}{1 + b_{AD}R(1 - Y_{AD})} = \frac{Z_{AD}}{S_{bpi} - S_{bp}} \qquad (4\text{-}11)$$

4.5.3. Stoichiometry for AD

The stoichiometry section of the model determines how many mols of each product is created from the breakdown of the number of mols of biodegradable organics. This then allows the concentrations of AD products to be calculated. The mols of biodegradable organics converted is determined from the amount of biodegradable COD converted to AD products in the kinetic section of the model (described in Section 4.5.2) and the molar mass of the biodegradable organics. The equations below give more detail to this process.

The stoichiometry of anaerobic digestion used to model this bioprocess is as listed in Harding *et al* (2011) (See also Ekama, 2009) :

$$\begin{aligned} C_xH_yO_zN_aP_b &+ \left[2x - z + a + b(2+f) - E\frac{\gamma_S}{\gamma_B}(2k - m + n + p(2+f)) \right. \\ &\left. - \frac{2\gamma_S}{8}(1-E)\right] H_2O \\ &\rightarrow \left[x - a + b(2-f) - E\frac{\gamma_S}{\gamma_B}(k - n + p(2-f)) - \frac{(1-E)\gamma_S}{8}\right] CO_2 \\ &+ \left[\frac{\gamma_S}{8}(1-E)\right] CH_4 + \left(E\frac{\gamma_S}{\gamma_B}\right) C_kH_lO_mN_nP_p + \left(a - nE\frac{\gamma_S}{\gamma_B}\right) NH_4^+ \\ &+ f\left(b - pE\frac{\gamma_S}{\gamma_B}\right) H_2PO_4^- + (1-f)\left(b - pE\frac{\gamma_S}{\gamma_B}\right) HPO_4^{2-} \\ &+ \left(a - b(2-f) - E\frac{\gamma_S}{\gamma_B}(n - p(2-f))\right) HCO_3^- \end{aligned} \qquad (4\text{-}12)$$

Where, the electron donating capacity of the biodegradable organics:

$$\gamma_S = 4x+y-2z-3a+5b, \qquad e^-/\text{mol (4-13)}$$

The electron donating capacity of the AD biomass:

$$\gamma_b = 4k+l-2m-3n+5p, \qquad e^-/\text{mol (4-14)}$$

And where f is the split of phosphorous to the $H_2PO_4^-$ species, with the balance HPO_4^{2-}. The total phosphorous is given by the summation of its various sub-species.

$$TP = [H_3PO_4] + [H_2PO_4^-] + [HPO_4^{2-}] + [PO_4^{3-}] \qquad \text{mol/l (4-15)}$$

For the phosphate system between 4 < pH <10 the concentrations of $[H_3PO_4]$ and $[PO_4^{3-}]$ are negligible compared to $[H_2PO_4^-]$ and $[HPO_4^{2-}]$ (Loewenthal, Ekama and Marais, 1989). Thus, the phosphorous split of f and (1-f) to $[H_2PO_4^-]$ and $[HPO_4^{2-}]$ respectively is justified.

The COD equivalent of influent biodegradable organics is given by:

$$S_{bo} = 8 \cdot \gamma_S \qquad \text{gCOD/l (4-16)}$$

The molar mass (MM) of the influent biodegradable organics is given by, gVSS/mol:

$$MM = 12x + y + 16z + 14a + 31b \qquad \text{gVSS/mol (4-17)}$$

If the biodegradable organics have their VSS composition known with respect to COD (f_{cv} gCOD/gVSS), TOC (f_c, gC/gVSS), organic nitrogen (f_n, gN/gVSS), organic phosphorous (f_p, gP/gVSS) then their molar composition can be determined from the equations below.

Assume a y value, say:

$$y = 7$$

$$z = \frac{y}{2}\left(\frac{1 - \frac{1}{8}f_{cv} - \frac{8}{12}f_C - \frac{17}{14}f_N - \frac{26}{31}f_P}{1 + f_{cv} - \frac{44}{12}f_C + \frac{10}{14}f_N - \frac{71}{31}f_P}\right) \qquad \text{(4-18)}$$

$$x = \frac{f_C}{12}\left(\frac{y + 16z}{1 - f_C - f_N - f_P}\right) \qquad \text{(4-19)}$$

$$a = \frac{f_N}{14}\left(\frac{y + 16z}{1 - f_C - f_N - f_P}\right) \qquad \text{(4-20)}$$

$$b = \frac{f_P}{31}\left(\frac{y + 16z}{1 - f_C - f_N - f_P}\right) \tag{4-21}$$

Biomass is produced from some of the COD consumed. The production of endogenous residue is low and it is assumed to have negligible concentration in the digester. The mass fractions of the AD biomass are shown in Table 4-11. These are taken from Ekama (2009).

Table 4-11: AD biomass mass fractions

COD content of biomass	1.416	gCOD/g mass
Carbon content of biomass	0.531	gC/g mass
Nitrogen content of biomass	0.124	gN/g mass
Phosphorous content of biomass	0.025	gP/g mass

4.5.4. Polyphosphate (PP) release stoichiometry

The P release from both polyphosphate (PP) breakdown and biomass death takes place in the AD, According to Ikumi and Harding (2020) , the PP release occurs quicker than the AD hydrolysis rate, therefore it is assumed all PP-P from PAO's is released during AD of BEPR WAS. However, the organically bound P (which contributes a smaller quantity of P in BEPR WAS) is released with death of biomass and hydrolysis of its biodegradable particulate organics (Ikumi and Ekama, 2019). This section gives the stoichiometry of the release of PP, where the release of the organically bound P is carried out according to the stoichiometry in Section 4.5.3.

The PAO's have the following generic formula made up from biomass and PP content:

$$PAO\ biomass \cdot q(PP)$$

$$C_xH_yO_zN_aP_b \cdot q(Mg_cK_dCa_ePO_3)$$

The biomass portion is given by $C_xH_yO_zN_aP_b$ and the PP portion is given by $Mg_cK_dCa_ePO_3$. The amount of PP associated with the biomass is linked by "q" which represents the molPP/molPAO biomass. This value along with the PP stoichiometric mass fraction are given in Table 4-6.

The release of PP and its impact on anaerobic weak acid base chemistry was evaluated by Harding *et al* (2011). Steady-state stoichiometric equations were extended to include orthophosphate release from PP in addition to biomass P.

Studies show that during the release of PP in the anaerobic zone of an activated sludge system energy-rich poly3-hydroxybutyrate (PHB) is formed in the presence of readily biodegradable COD (Wentzel *et al.*, 1990). It is thus assumed that in the anaerobic digester the same process occurs using volatile fatty acids. However, in the digester there is no alternating aerobic zone to allow PAO growth and all the PAO's stored products are eventually released. Ikumi & Ekama (2019) further developed the stoichiometry to include acetate uptake for PHB formation and the subsequent PP release that occurs with PA death and hydrolysis.

The stoichiometry for PP release from Ikumi & Ekama (2019) is given as:

$$
\begin{aligned}
Mg_cK_dCa_ePO_3 + H_2O \\
&= \left[(1-f)\cdot q - \left(f_{qPP}\cdot Y_{PP}\cdot q\cdot\frac{9}{4}\right)\right]CO_2 + \left[f_{qPP}\cdot Y_{PP}\cdot q\cdot\frac{1}{4}\right]CH_4 \\
&+ \left[(1-f)\cdot(-q) - \left(f_{qPP}\cdot Y_{PP}\cdot q\cdot 2\right)\right]HCO_3^- + [f\cdot q]H_2PO_4^- \\
&+ [(1-f)\cdot q]HPO_4^- + [q\cdot c]Mg^{2+} + [q\cdot d]K^+ + [q\cdot e]Ca^{2+}
\end{aligned} \tag{4-22}
$$

Where,

- q is the ratio of PP to biomass, molPP/molPAOVSS, given in Table 4-6
- f is the split of phosphorous to the $H_2PO_4^-$ species, with the balance HPO_4^{2-}
- Y_{PP} is the mols of P released per mol of PHB formed, which can vary, and in this study is taken as 0.65 (Smolders *et al.*, 1995)
- f_{qPP} is the fraction of PP released with PHB uptake, and the balance is released by PAO death. For conventional digestion this is given as 0.8 (Smolders *et al.*, 1995) and set to zero for THP digestion (Han *et al.*, 2017).

In THP digestion the polyphosphate chains are all broken down and P released as OP prior to digestion during the THP step. Hence, the f_{qPP} in the above PP release equation was set to zero in the case of THP digestion.

4.5.5. Struvite precipitation stoichiometry inside AD

According to Harding *et al* (2011) digestion of PP rich WAS from BEPR activated sludge releases metals and phosphates, which with the presence of saline ammonia, results in precipitation of struvite.

The stoichiometry for struvite precipitation is given by Harding *et al* (2011) as:

$$
\begin{aligned}
&Mg + NH_4^+ + [f]H_2PO_4^- + [1-f]HPO_4^{2-} + [1+f]HCO_3^- \\
&= MgNH_4PO_4 + [1+f]CO_2
\end{aligned} \tag{4-23}
$$

The extent of struvite precipitation potential can be estimated from the difference between the ionic product (IP) and the solubility product (K_s'). The model assesses the struvite precipitation potential based on the molar products of the reacting species. Either Mg^{2+}, NH_4^+ or PO_4^{3-} could be limiting. If the IP is greater than K_s' then the precipitation is predicted.

In solving the model the amount of struvite precipitated was varied until the solubility product and ionic product were equal.

The solubility product for struvite was corrected for ionic activity is given by Loewenthal *et al* (1994) as follows:

$$
K_s' = f_m \cdot f_d \cdot f_t \cdot K_s \tag{4-24}
$$

Where,

- K_s' is the activity corrected solubility product for struvite
- K_s is the solubility product for struvite at 25°C of $10^{-12.6}$ (or pK_s = 12.6)
- f_m is the monovalent activity coefficient, applicable from NH_4^+

- f_d is the divalent activity coefficient, applicable due to Mg^{2+}
- f_t is the trivalent activity coefficient, applicable due to PO_4^{3-}

The ionic product of struvite in the anaerobic digester is:

$$IP = [Mg^{2+}] \cdot [NH_4^+] \cdot [PO_4^{3-}] \tag{4-25}$$

Where,

- Mg^{2+} is the concentration of magnesium ions, mol/l, largely from the breakdown of polyphosphate
- NH_4^+ is the concentration of ammonium species, mol/l
- PO_4^{3-} is the concentration of phosphate ions, mol/l

The above total dissolved phosphorous and nitrogen are split between various species according to pH and equilibrium constants, corrected for ionic activity. The following speciation formulae are applicable to this study and are used to calculate the concentrations above.

Triprotic orthophosphate

$$[H_3PO_4] = PT \frac{W_p}{(1 + W_p + X_p + X_pY_p)} \tag{4-26}$$

$$[H_2PO_4^-] = PT \frac{1}{(1 + W_p + X_p + X_pY_p)} \tag{4-27}$$

$$[HPO_4^{2-}] = PT \frac{X_p}{(1 + W_p + X_p + X_pY_p)} \tag{4-28}$$

$$[PO_4^{3-}] = PT \frac{X_pY_p}{(1 + W_p + X_p + X_pY_p)} \tag{4-29}$$

$$where\ W_p = 10^{pK'_{p1} - pH}, X_p = 10^{pH - pK'_{p2}}\ and\ Y_p = 10^{pH - pK'_{p3}}$$

Monoprotic free and saline ammonia

$$[NH_3] = N_T \frac{1}{(1 + W_n)} \tag{4-30}$$

$$[NH_4^+] = N_T \frac{W_n}{(1 + W_n)} \tag{4-31}$$

$$Where\ W_n = 10^{pK_n' - pH}$$

Diprotic inorganic carbon

$$[H_2CO_3] = C_T \frac{W_c}{(1 + W_c + X_c)} \tag{4-32}$$

$$[HCO_3^-] = C_T \frac{1}{(1 + W_c + X_c)} \tag{4-33}$$

$$[CO_3^{2-}] = C_T \frac{X_c}{(1 + W_c + X_c)} \tag{4-34}$$

$$where\ W_c = 10^{pK_{c1}' - pH}\ and\ X_c = 10^{pH - pK_{p2}'}$$

4.5.6. Ionic activity

The dissociation constants used in the calculations in this study have been corrected for non-ideality. In solutions ions interact with each other and with surrounding water molecules and at low concentrations these interactions are negligible. The solution can be called an ideal solution. However, as concentration increases the interactions become more apparent and the solution becomes non-ideal. For solutions where ionic strength exceeds 0.2mol/kg (taken as 0.2mol/l assuming AD liquor has density of water) corrections for ion activity become necessary, especially when calculating pH from a model (Tait *et al.*, 2012). This is applicable for anaerobic digester mixed liquor which can range from 0.1-1 mol/kg, and can be seen as non-ideal, especially with the higher concentration of THP digestion.

To cater for the extent of non-ideality a correction is made to an ion's concentration by applying an activity coefficient and in doing creates the ion's activity. It is this activity that is used in calculations instead of concentration (Batstone *et al.*, 2012) . Activity is calculated as follows:

$$a_i = \gamma_i \cdot C_i^{Z\pm} \tag{4-35}$$

Where,

- a_i is the activity of component I, mol/l
- γ_i is the activity coefficient calculated, dimensionless
- $C_i^{Z\pm}$ is the concentration of component I, mol/l

The activity tends to be lower than the actual concentration and is a representation of the quantity of an ion's availability to partake in reactions. The greater the charge of the ion (positive or negative), the greater the effect of non-ideality. The activity in a non-ideal solution is seen to be analogous to molality in an ideal solution.

The correction is also applied to dissociation constants for non-ideal solutions as shown in Loewenthal (1989). The valency of the ions is also considered in the correction.

$$K_i' = f_i \cdot K_i \tag{4-36}$$

Where,

- K_i' is the dissociation constant for I and corrected for activity
- f_i is the activity coefficient, and is represented as f_m for monovalent ions, f_d for divalent ions and f_t for trivalent ions.
- K_i is the activity coefficient without any correction, calculated above.

The activity coefficients of ions can be calculated from the Debye-Huckel theory, and one of the most widely used modifications of it is the Davies equation. The Davies equation is suitable for ionic strengths up to around 0.5mol/kg (Tait *et al.*, 2012). This requires ionic strength to be calculated which requires a complete analysis of the water. However, the activity coefficients calculated from the Davies equation are not very sensitive to ionic strength (Loewenthal, Ekama and Marais, 1989) and thus a calculated approximation will suffice for this study. The Davies equation used in this study is:

$$\log f_i = -A \cdot z_i^2 \cdot \left(\frac{\mu^{0.5}}{1+\mu^{0.5}} - 0.3\mu\right) \quad (4\text{-}37)$$

Where,

- f_i is the activity coefficient, and is represented as f_m for monovalent ions, f_d for divalent ions and f_t for trivalent ions.
- μ is ionic strength, mol/l
- z_i is the charge of ion i
- A is a temperature dependant constant given by $A = 1.825 \times 10^6 \cdot (78.3 \cdot T)^{-1.5}$, where T is temperature in degrees Kelvin

Ionic strength was calculated from the major dissolved species in the water released from the anaerobic digestion of the feed substrate and are mainly HCO_3^-, H_2PO_4-, HPO_4^{2-}, NH_4^- and metal ions released from polyphosphate Mg^{2+}, Ca^{2+} and K^+. Ionic strength can be calculated from Bhuiyan *et al* (2009) by:

$$I = \frac{1}{2}\sum C_i \cdot z_i^2 \quad (4\text{-}38)$$

Where,

- I is ionic strength in mol/l
- C_i is the concentration of ion i, mol/l
- z_i is the charge of ion i

4.5.7. Alkalinity

The total alkalinity of the system will be calculated for the most protonated species, as shown in Lowenthal *et al* (1989). This will be done according to the formula below:

$H_2CO_3/NH_4^+/H_3PO_4\ alk = alk\ H_2CO_3 + alk\ NH_4^+ + alk\ H_3PO_4 + alk\ H_2O$	(4-39)

Where,

- $alk\ H_2CO_3 = [HCO_3^-]+2[CO_3^{2-}]$
- $alk\ NH_4^+ = [NH_3]$
- $alk\ H_3PO_4 = [H_2PO_4^-] + 2[HPO_4^{2-}] + 3[PO_4^{3-}]$
- $alk\ H_2O = [OH\text{-}]\text{-}[H^+]$

Alk as a prefix means water alkalinity is excluded and alk as a suffix means water alkalinity is include. The alkalinity of AD of is impacted by significantly by the release of organic N released from the breakdown of biodegradable organics. However, in AD for NDEBPR WAS the release of PP also plays a role in alkalinity generation.

4.5.8. Weak acid base chemistry and pH

The CO_2 produced from the bioprocess stoichiometry dissolves at the near neutral pH of the digester, creating mostly bicarbonate (HCO_3^-). The ammonia released in the breakdown of the hydrolysable organics picks up a proton from the dissolved CO_2 (from H_2CO_3) to form saline ammonia (NH_4^+) and bicarbonate (HCO_3^-), given by Sotemann (2005) as:

$$NH_3 + H_2CO_3^* = NH_4^+ + HCO_3^- \tag{4-40}$$

The influent VFA concentration to the digester is assumed to be acetate species. The extent to which this is used in the digester is impacted by the extent of its inlet dissociation, which is governed by inlet pH. Undissociated acetate is converted to carbon dioxide which lowers digester pH via (S. W. Sötemann *et al.*, 2005):

$$CH_3COOH = CH_4 + CO_2 \tag{4-41}$$

Dissociated acetate is converted to methane and bicarbonate, increasing alkalinity and pH via:

$$CH_3COO^- = CH_4 + HCO_3^- \tag{4-42}$$

<u>Influent VFA</u>

THP converts complex substances to soluble organics, including VFA's. While this increases VFA concentration, at the same time significant alkalinity is generated through the hydrolysis of substances (Flores-Alsina *et al.*, 2019). Downstream in the digester high solids digestion of THP pre-treated WAS provides alkalinity and THP digesters often run at higher pH 7.5 to 8 than conventional digestion of pH 7 (Barber, 2016). Han *et al* (2017) showed that VFA's in the feed sludge to AD increase from 250mg/l to 4200mg/l due to THP, while alkalinity increased from 670mg/l $CaCO_3$ to 4230mg/l $CaCO_3$ in the THP step. In the downstream AD the digestion of the THP sludge then resulted in alkalinity rising to 17820mg/l as $CaCO_3$ and digester pH was 8.03. Xue *et al* (2015) fed digesters sludge which had been THP pre-treated at various temperatures resulting in digester feed VFAs ranging from 1200-1800gm/l while the AD maintained pH 7.7 to 7.9. In a study done by Wilson and Novak (2009) it is was found that VFA production in THP was not significant in terms of methanogenic inhibition. Considering THP digesters often run at up to over 10 000mg/l as $CaCO_3$ this would imply the AD could tolerate up to around 3000mg/l VFA before the Ripley ratio exceeds the recommended 0.3 (Merwe-Botha, Borland and Visser, 2019) for stable digestion, as discussed in Section 2.4.3.

While VFAs may be high in the digester feed there is strong evidence in Section 2.7.2 of the literature review to show this is not a concern as THP digestion runs at a higher pH and alkalinity than conventional digestion.

For the purposes of this modelling exercise and to limit scope, all inlet biodegradable organics are lumped together as one term, which includes the inlet VFA's, and thus the processing of organics is done via the kinetics and bioprocess stoichiometry discussed in sections 4.5.2 to 4.5.5, negating the need for inlet pH and VFA concentration.

pH

The pH of the digester is defined by the carbon dioxide partial pressure (p_{CO2}) and bicarbonate concentration generated. The carbonate weak acid base system predominantly defines the pH. Equilibrium exists between the dissolved and gaseous inorganic carbon species. This is given by Sotemann *et al* (2005) as:

$$p_{CO_2} = \frac{[HCO_3^-]\left(1 + 10^{pK'_{c1}-pH} + 10^{pH-pK'_{c2}}\right)}{10^{-pK'_{HCO_2}}\left(1 + 10^{pH-pK'_{c1}} + 10^{2pH-pK'_{c1}-pK'_{c2}}\right)} \tag{4-43}$$

Where,

- pK'_{HCO2} = -ve log10 of the apparent Henry's law constant for CO_2
- pK'_{C1}, pK'_{C2},= -ve log10 of 1st and 2nd carbonate system apparent dissociation constants where apparent means corrected for ionic strength effects.

The weak acid dissociation constants and their temperature dependency for the various species in this investigation were taken from Loewenthal *et al* (1989) and are calculated using the following formula:

$$pK_i = \frac{A}{T} - B + C \cdot T \tag{4-44}$$

Where,

- pK_i is the negative log value of dissociation constant K_i for species i
- A, B and C are terms taken from Table 2 in Loewenthal *et al* (1989)
- T is temperature in degrees Kelvin

4.6. MASS BALANCE VERIFICATION

To add a degree to confidence in the model mass balances were carried out over both the kinetic and stoichiometric sections. The kinetic section of the model uses biodegradable COD to predict the extent of degradation based on the kinetics constants. The stoichiometric section then uses the amount of biodegradable organics converted to predict the quantities of reactants and products of anaerobic digestion. The mass balances assume that no water leaves the system and includes the gas flows in addition to the aqueous and solids flows. Mass balances must be 100% to verify the internal consistency of the model.

4.6.1. Mass balance over kinetics

The kinetic section focuses solely on COD and thus a COD balance was appropriate for checking this section. The difference between influent COD (S_{ti}) and the effluent COD (S_{te}) was carried out. The influent COD are essentially all the biodegradable COD (S_{bi}) available for hydrolysis and the unbiodegradable COD (S_{ui}). The products in this section of the model are methane (S_m) , AD biomass (Z_{AD}) and residual unused biodegradable organics (S_b). The unbiodegradable organics in the influent remain unchanged and simply pass through the AD.

$$S_{ti} = S_{te} \qquad \text{gCOD/l (4-45)}$$

$$S_{bi} + S_{ui} = S_m + Z_{AD} + S_b + S_{ui} \qquad \text{gCOD/l (4-46)}$$

4.6.2. Mass balance over stoichiometry

The stoichiometry section uses the extent of biodegradable organics converted to predict the amount of AD reactants and products synthesized.

COD

As in the kinetics section, a COD balance is applicable and is used as second check to make sure the amount of organics hydrolysed in the in the kinetics section matches that processed via the stoichiometry. The left-hand side of each of the following equations represents the gCOD/l produced in the kinetics section and the right-hand side represents the gCOD/l from the stoichiometry.

The use of brackets "[…]" here refers to the molar concentration of a substance in mol/l.

Biodegradable organics hydrolysed:

$$S_{bi} - S_b = \gamma_s \cdot 8 \cdot [C_xH_yO_zN_aP_b] \qquad \text{gCOD/l (4-47)}$$

Methane produced:

$$S_m = \gamma_{CH_4} \cdot [CH_4], where\ \gamma_{CH_4} = 4 \cdot 1 + 4 \cdot 4 \qquad \text{gCOD/l (4-48)}$$

Biomass produced:

$$Z_{AD} = \gamma_b \cdot [C_kH_lO_mN_nP_p] \qquad \text{gCOD/l (4-49)}$$

Then arranging to show COD balances, the COD balance for kinetics is:

$$S_{bi} - S_b = S_m + Z_{AD} \qquad \text{gCOD/l (4-50)}$$

The COD balance for stoichiometry is:

$$\gamma_s \cdot 8 \cdot \left[C_xH_yO_zN_aP_b\right] = \gamma_{CH_4} \cdot [CH_4] + \gamma_b \cdot [C_kH_lO_mN_nP_p] \qquad \text{gCOD/l (4-51)}$$

<u>Carbon</u>

A mass balance check was carried out for total carbon (TC). The difference between the influent concentration and effluent concentration was compared.

The influent TC (C_i) is made up from the C content of the organics in the primary sludge and WAS. This can be split into the unbiodegradable ($C_{i,UO}$) and biodegradable ($C_{i,BO}$) fractions in both the PS and WAS.

$$C_i = C_{i,UO} + C_{i,BO} \qquad \text{gC/l (4-52)}$$

$$C_i = f_{c,PS} \cdot X_{i,PS} + f_{c,WAS} \cdot X_{i,WAS} \qquad \text{gC/l (4-53)}$$

$$\begin{aligned} C_i = {} & \Big((f_{c,UO,PS} \cdot f_{S'u,PS} + f_{c,BO,PS} \cdot \big(1 - f_{S'u,PS}\big)\Big) \cdot X_{i,PS} \\ & + \Big((f_{c,UO,WAS} \cdot f_{S'u,WAS} + f_{c,BO,WAS} \cdot \big(1 - f_{S'u,WAS}\big)\Big) \\ & \cdot X_{i,WAS} \end{aligned} \qquad \text{gC/l (4-54)}$$

The effluent carbon (C_e) is made up of the unbiodegradable carbon that passed through from the influent ($C_{i,UO}$), the carbon in the biomass (C_{AD}), the carbon in the residual biodegradable organics (C_{BO}), dissolved inorganic carbon, carbon dioxide and methane.

$$C_e = C_{i,UO} + C_{AD} + C_{BO} + C_{HCO_3^-} + C_{CO_2} + C_{CH_4}$$ gC/l (4-55)

$$C_e = C_{i,UO} + f_{c,AD} \cdot X_{AD} + f_{c,BO} \cdot X_{BO} + 12 \cdot ([HCO_3^-] + [CO_2] + [CH_4])$$ gC/l (4-56)

A mass balance over carbon was then carried out by checking:

$$C_i = C_e$$ gC/l (4-57)

Nitrogen

A mass balance was carried out on nitrogen. The inlet nitrogen was made up from the nitrogen contained in the biodegradable organics ($N_{i,BO}$) and unbiodegradable organics ($N_{i,UO}$) and any FSA (N_{ai}) accompanying the liquid stream. Influent nitrogen was calculated by:

$$N_i = N_{i,UO} + N_{i,BO} + N_{ai}$$ gN/l (4-58)

$$N_i = f_{N,PS} \cdot X_{i,PS} + f_{N,WAS} \cdot X_{i,WAS} + N_{ai}$$ gN/l (4-59)

$$N_i = \Big((f_{N,UO,PS} \cdot f_{S'u,PS} + f_{N,BO,PS} \cdot \big(1 - f_{S'u,PS}\big)\Big) \cdot X_{i,PS} + \Big((f_{N,UO,WAS} \cdot f_{S'u,WAS} + f_{N,BO,WAS} \cdot \big(1 - f_{S'u,WAS}\big)\Big) \cdot X_{i,WAS}$$ gN/l (4-60)

The effluent nitrogen (N_e) is made up of the unbiodegradable nitrogen that passed through the digester from the influent ($N_{i,UO}$), the nitrogen in the biomass (N_{AD}), the nitrogen in the residual biodegradable organics (N_{BO}), the FSA released (N_{ae}) and that in the influent, and the nitrogen contained in the struvite precipitated (N_s).

$$N_e = N_{i,UO} + N_{AD} + N_{BO} + N_{ai} + N_a + N_s \qquad \text{gN/l (4-61)}$$

$$N_e = N_{i,UO} + f_{N,AD} \cdot X_{AD} + f_{N,BO} \cdot X_{BO} + 14 \cdot ([NH_4^+] + [MgNH_4PO_4]) \qquad \text{gN/l (4-62)}$$

Phosphorous

A mass balance was carried out on phosphorous (P). The inlet P was made up from that contained in the biodegradable organics ($P_{i,BO}$) and unbiodegradable organics ($P_{i,UO}$) and any OP (P_{ai}) accompanying the liquid stream. The polyphosphate content (PP) of the WAS also contributed to the influent P. The total influent P was calculated by:

$$P_i = P_{i,UO} + P_{i,BO} + P_{ai} + PP \qquad \text{gP/l (4-63)}$$

$$P_i = \ f_{P,PS} \cdot X_{i,PS} + (f_{P,OHO} + \cdot\, f_{XBGPP} \cdot f_{BG,WAS})\, X_{i,WAS} \qquad \text{gP/l (4-64)}$$

$$P_i = \Big((f_{P,UO,PS} \cdot f_{S'u,PS} + f_{P,BO,PS} \cdot \big(1 - f_{S'u,PS}\big)\Big) \cdot X_{i,PS} + \big((f_{P,UO,WAS} \cdot f_{S'u,WAS} + f_{P,BO,WAS} \cdot \big(1 - f_{S'u,WAS}\big) + f_{XBGPP} \cdot f_{BG,WAS}\big) \cdot X_{i,WAS} \qquad \text{gP/l (4-65)}$$

The effluent P (P_e) is made up of the unbiodegradable P that passed through the digester from the influent ($P_{i,UO}$), the P in the biomass (P_{AD}), the P in the residual biodegradable organics (P_{BO}), the OP released (P_a) along with the OP in the influent (P_{ai}) , and the P contained in the struvite precipitated.

$$P_e = P_{i,UO} + P_{AD} + P_{BO} + P_{ai} + P_a + P_s \qquad \text{gP/l (4-66)}$$

$$P_e = P_{i,UO} + f_{P,AD} \cdot X_{AD} + f_{P,BO} \cdot X_{BO} + 31 \cdot ([H_2PO_4^-] + [HPO_4^{2-}] + [MgNH_4PO_4]) \quad \text{gP/l (4-67)}$$

Metals

A metal balance mass balance was carried out. The metals considered in this study were those contained in the polyphosphate compound i.e. Mg, K and Ca. The influent amounts were essentially those contained within the PP in the influent WAS. The effluent concentrations were those released from PP breakdown, while including the struvite precipitated in the case of magnesium.

The magnesium mass balance was done as follows:

$$Mg_i = Mg_e \quad \text{gMg/l (4-68)}$$

$$f_{Mg} \cdot f_{BG,WAS} \cdot X_{i,WAS} = 24 \cdot ([MgNH_4PO_4] + [Mg^{2+}]) \quad \text{gMg/l (4-69)}$$

The potassium mass balance was done as follows:

$$K_i = K_e \quad \text{gK/l (4-70)}$$

$$f_K \cdot f_{BG,WAS} \cdot X_{i,WAS} = 39 \cdot [K^+] \quad \text{gK/l (4-71)}$$

The calcium mass balance was done as follows:

$$Ca_i = Ca_e \quad \text{gCa/l (4-72)}$$

$$f_{Ca} \cdot f_{BG,WAS} \cdot X_{i,WAS} = 40 \cdot [Ca^{2+}] \quad \text{gCa/l (4-73)}$$

5. MODEL APPLICATION RESULTS

5.1. CONVENTIONAL MAD VS THP-MAD

This section discusses the modelled outputs of the digestion performance of each case and notes major differences. The overall process in each case is explained in more detail in Section 3.1. The development of the model for each case is discussed in more detail in Section 4. To add a more plantwide view to each case some of results include the requirements for nutrient treatment, the details of which are discussed in Section 3.3.6. These have not been modelled to the same extent as the AD and AD+THP processes, which are the main focus of this study. Rather, they are briefly discussed in these results to show their relevance as being an important consideration for AD digestion facilities.

The outputs related to operating costs and the quantifying thereof will be presented in section 5.3.

5.1.1. FSA and TKN

The digester effluent nitrogen concentrations are shown in Figure 5-1. The concentration of FSA inside the AD are calculated using equation (4-12). The concentration of FSA after P-treatment is estimated using the stoichiometry in equation (2-1) and the P removal efficiency stated in Section 3.3.6.

As can be seen the THP digestion has a much higher FSA concentration of 2793mgN-NH_4^+/l than that in conventional digestion of 914mgN-NH_4^+/l. These values are within the ranges found in literature for conventional in digestion of 500-1500mgN-NH_4^+/l in Section 2.4 and for THP digestion of 2500-3500mgN-NH_4^+/l found in Section 2.7 of the literature review. This suggest the model is predicting digester FSA with a reasonable amount of certainty.

The higher FSA concentration in the THP digestion case is most likely due to running the THP digester at a higher solids concentration and the corresponding higher FSA release. As discussed in Section 2.7.3 of the literature review a high FSA content could potentially be inhibitory to digester performance, however as long as the free ammonia nitrogen (FAN) fraction of the FSA is below inhibitory levels of 600mgFAN/l. The FAN is calculated for each case by equation (4-30) and is 10mgFAN/l and 72mgFAN/l for conventional and THP digestion respectively. These values are both well below the inhibitory threshold and therefore inhibition due high FSA concentration in the digesters is not expected.

The FSA concentration in each case might have been higher if WAS containing no PP had been digested. The digestion of NDBEPR WAS containing significant levels of PP most likely lowers the FSA concentration of the AD due to struvite precipitation inside the digester, as discussed in Sections 2.4.4 and 4.5.5. Thus, if THP high solids digestion is to be done on MLE WAS the AD should be carefully monitored and controlled by dilution of the fed sludge to maintain a healthy FAN concentration <400mgFAN/l to manage any risk of ammonia inhibition.

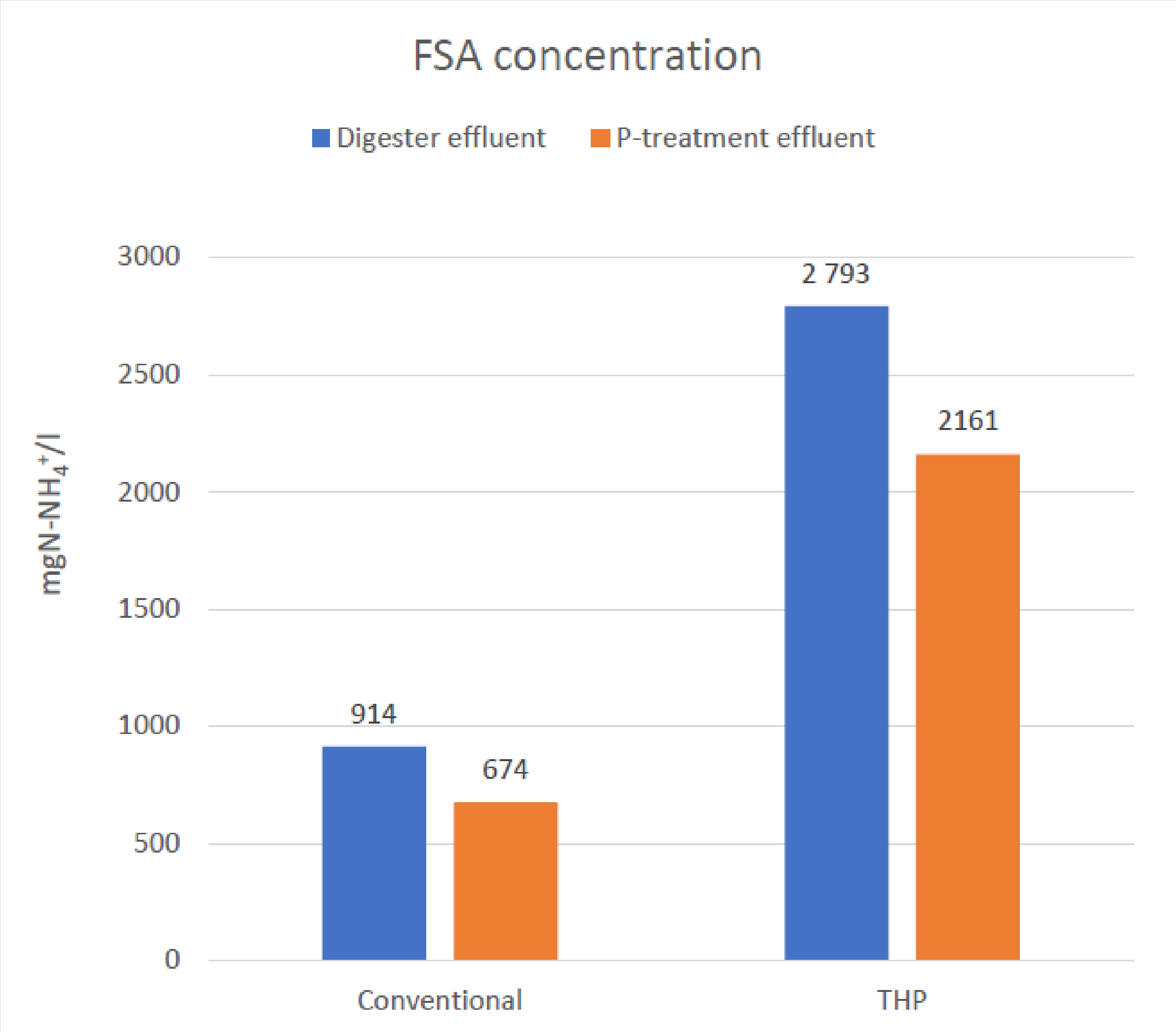

Figure 5-1:FSA concentration

The effluent TKN load requiring downstream treatment by anammox is shown in Figure 5-2. This load is mostly FSA and with some organic TKN from the residual biodegradable organics and AD biomass. The nitrogen in the unbiodegradable organics is not included as this passed through the digester unchanged and will subsequently pass through the downstream nitrogen side-stream treatment step, thus is not considered when defining anammox treatment load. It can be seen that the load in each case is progressively reduced as the digester effluent leaves the digester, moves through P-treatment and then is discharged from final dewatering. The final TKN load for THP digestion requiring N-treatment of 2198kgTKN/d and is approximately 3 times greater than the load from conventional digestion of 724kgTKN/d. This implies the downstream anammox treatment would need to be larger and would incur significantly higher operating costs in THP digestion.

The figure also shows that using struvite precipitation for P-treatment brings a significant reduction in the FSA portion of the TKN load required for downstream treatment. The digestion of NDBEPR WAS results in a high release of OP which subsequently reduces TKN load through struvite precipitation inside the AD and then again during struvite precipitation from side-stream P-treatment. This reduces the downstream TKN load for the anammox plant. Around 15% of the FSA load is reduced from 5525kgN-TKN/d to 4641N-TKN/d by P-treatment struvite precipitation in the THP digestion case. However, if WAS with less P content were to be digested the TKN load to anammox would be higher due to less struvite precipitating and consuming FSA inside the digester, and less struvite would be formed in the downstream P-treatment step, also consuming less FSA. Therefore, if WAS with low P-content is to be digested instead of high-P NDBPER WAS then the downstream load requiring N-treatment will increase.

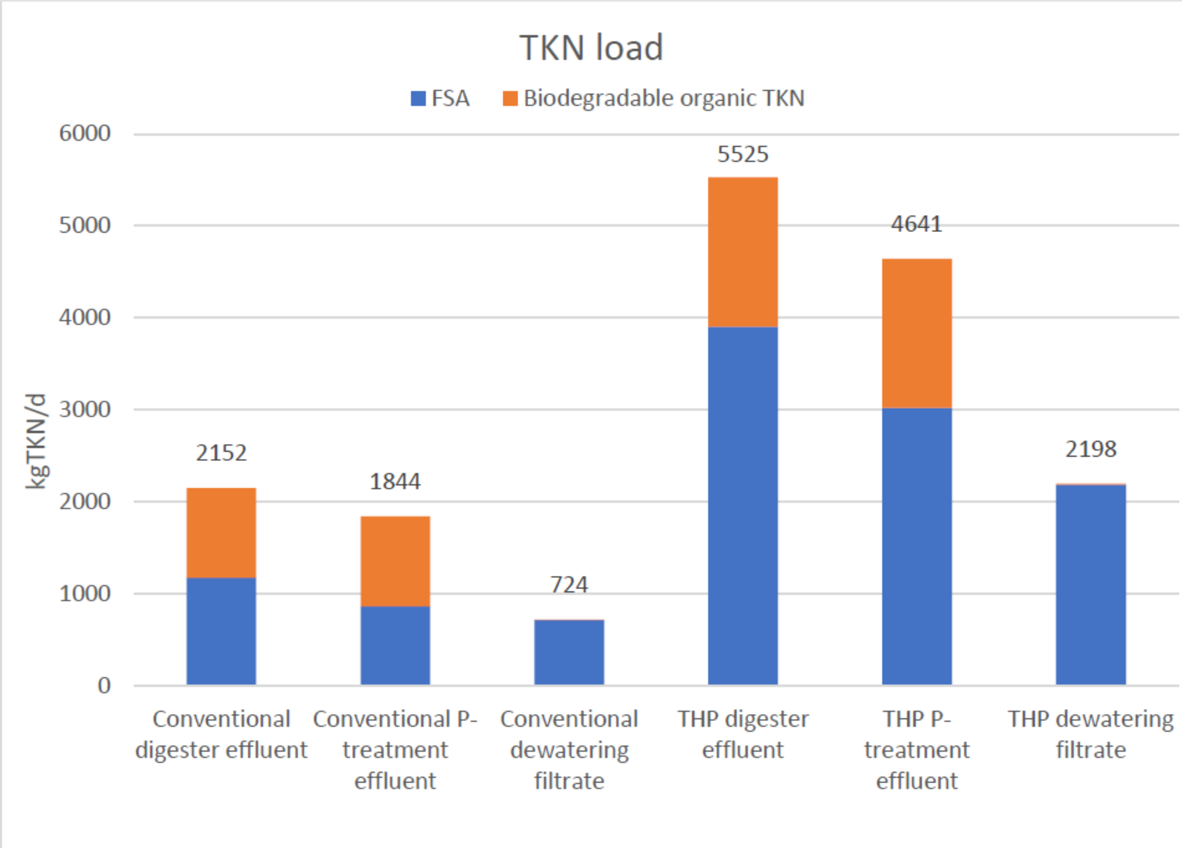

Figure 5-2: TKN load after digestion

5.1.2. Phosphate production

The digester effluent phosphorous is shown in Figure 5-3 below. The digester effluent OP concentrations were calculated from the stoichiometry in equation (4-12) and the P-treatment effluent was determined according to the OP reduction of 85% as discussed in Section 3.3.6.

There is a significantly higher OP concentration experienced in the THP digester of 1646mgP/l compared to that in conventional digestion of 624mg/l. This is expected due to a higher concentration of sludge being processed in THP digestion. However, the amount of sludge processed by THP digestion is 2.5 times greater than the conventional case (see Table 4-2) but the OP concentration is 2.6 times greater. One would expect this to be proportional, however, this small increase in proportionality in OP release is due to the increased rate of digestion as discussed in Section 4.5.2. While struvite precipitation inside the digester consumes some OP it is likely the OP concentration would be significantly lower if WAS from low biological P removal plants was digested. The way material P removal is done through running NDBEPR AS is via WAS sludge, therefore additional P-treatment is necessary if this WAS is to be anaerobically digested.

Both cases have high OP which requires treatment to not load the adjacent WWTW with return liquors (discussed in detail in Section 3.3.6). In Figure 5-4 the TP load requiring treatment leaving the digester is shown as the digester effluent moves through the unit operations downstream of the digester through struvite precipitation and dewatering. This TP load is made up of OP and the remaining biodegradable TP from the residual biodegradable organics and AD biomass. Implementing a P-treatment step such as struvite precipitation with MgO dosing as discussed in Section 3.3.6 has a clear benefit in significantly reducing the TP load from 1073kgTP/d to 391kgTP/d in conventional digestion and from 2733kgTP/d to 776kgTP/d in THP digestion.

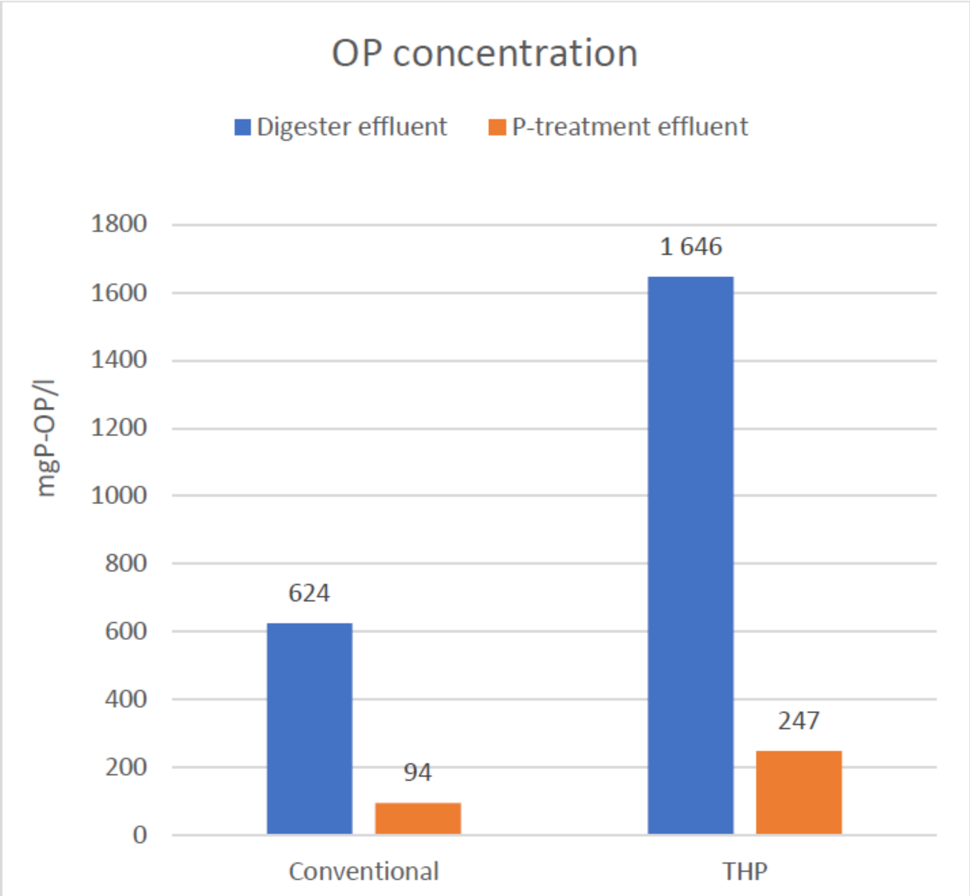

Figure 5-3: OP concentration

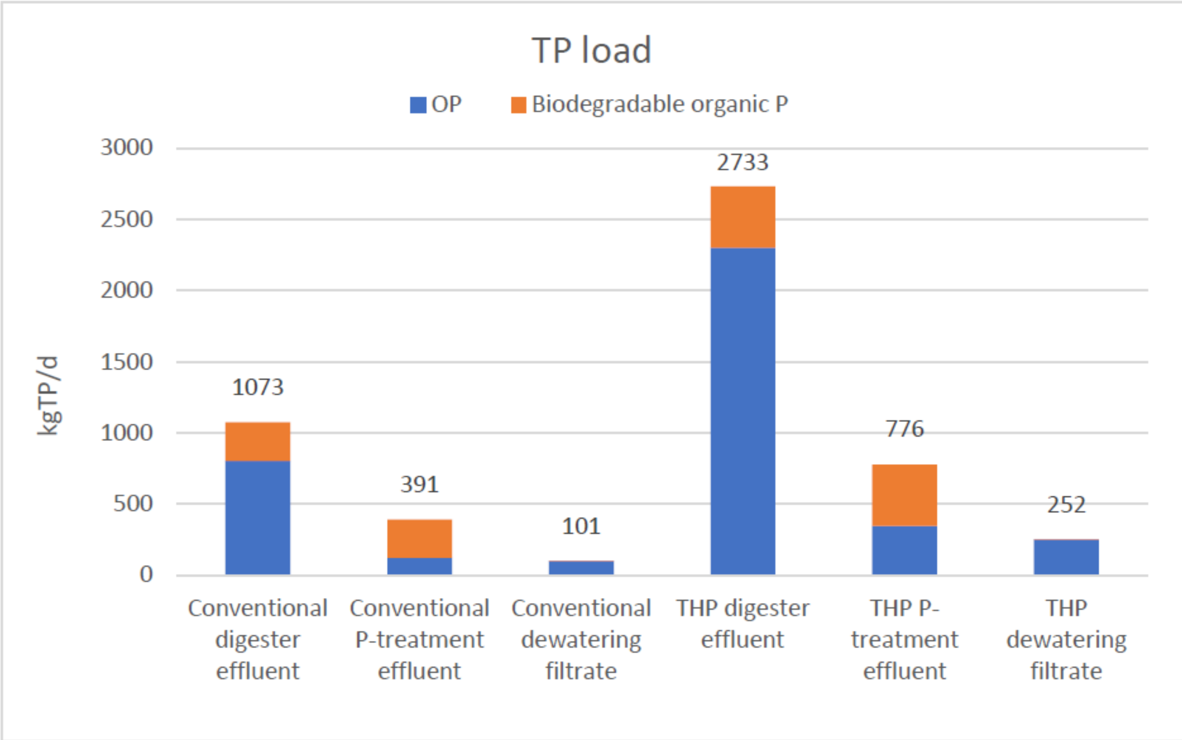

Figure 5-4: TP load requiring treatment

The reduction in TP load via P-treatment shown in Figure 5-4 is due to a reduction in the soluble OP concentration only, and organic P remains unchanged through the struvite treatment step. This is because struvite is formed from OP and not the biological P. It can be seen the OP fraction is significantly reduced by P-treatment using MgO dosing for forced struvite precipitation, and as OP makes up the majority of soluble P requiring treatment, struvite precipitation using MgO dosing is a suitable technology to reduce the TP load back to WWTW. Most of the biological TP other than OP is in the particulate form and can be removed via dewatering. This is shown by the reduction in the orange fraction of each bar between P-treatment effluent and dewatering filtrate in Figure 5-4.

5.1.3. Polyphosphate breakdown

The PP breakdown during anaerobic digestion results in the release of Mg, Ca and K into the aqueous phase. While the Mg and Ca are often found to precipitate to some extent it is assumed that no K precipitates for the operating AD pH range. Assuming no other sources of potassium, the concentration of K found in the digester effluent can be used to infer a measure of how much PP has been broken down. From Table 5-1 it can be seen that the P content of the PP in the PAO’s fed to digestion had 429mgP/l in the case of conventional digestion and 1004mgP/l in the case of THP digestion. This was a result of the WAS feed data presented Section 4.3.3. Using the concentration of potassium in the effluent and the stoichiometric ratios in Table 4-6 the corresponding amount of PP broken down was determined, and the subsequent amount of P released was 429mgP/l in conventional digestion and 1004mgP/l in THP digestion. Dividing the amount of PP released by that which was available in the influent sludge comes to unity. This implies that all the P in the PP fed to digestion was released and that all the PP was broken down into products according to the stoichiometry given in 4.5.4.

Table 5-1: Polyphosphate release

P available form PolyP	**Conventional**	**THP**	
WAS influent concentration	20777	48626	mgVSS/l
Active PAO % in WAS	33%	33%	gPAO,VSS/gVSS
Active PAO's concentration in WAS	6783	15875	mgVSS/l
Mass fraction of P in WAS from PP	6%	6%	gP/gPAO
Influent P in PP (P-PP available)	429	1004	mgP/l
P released from PolyP			
Potassium released	178	417	mgK/l
Potassium released	5	11	mmol/l
K:PP molar ratio	0.33	0.33	molK/molPP
P:PP molar ratio	1.00	1.00	molP/molPP
PP released	14	32	mmol/l
P released form PP	14	32	mmol/l
P released form PP (P-PP released)	429	1004	mgP/l
(P-PP released) / (P-PP available) %	100%	100%	
PP released	1387	3249	mgPP/l
PP released	1784	4544	kgPP/d

5.1.4. Struvite precipitation in digester

The struvite precipitation in the AD is shown in Table 5-2. The values shown below were calculated using the stoichiometry and methodology as discussed in Section 4.5.5.

The quantity of struvite precipitated was modelled by solving for the digester pH at which point the solubility product (pK_{sp}') and ionic product (pIP) were equal. At this point the solution is neither saturated or undersaturated and the solid phase of struvite is in equilibrium with its reactants. The value where the ionic product and solubility product of struvite equated was found to be lower in conventional digestion than in THP digestion, with pK_{sp}' being 11.02 and 10.79 respectively. Both of these values were lower than the standard pK_{sp} of 12.6 (Loewenthal, Kornmuller and van Heerden, 1994). The table below shows that pIP and pK_{sp}' were found to be equal for each case which implies equilibrium was reached for both THP digestion and conventional digester. The digester pH for each of these cases is given in Table 5-3.

More struvite was precipitated per mass of PP released in THP digestion 0.73kg/kg compared to 0.67kg/kg in conventional digestion. This was due to struvite having a greater equilibrium ionic product (and therefore a lower pIP) in THP digestion. This was due to THP digestion operating at a higher pH, as well as THP digestion operating at a higher solids concentration.

Evidence of the effect of the higher solids concentration is demonstrated by THP digestion having a higher ionic strength of 0.41mol/l compared to that of conventional digestion of 0.15mol/l. This is due to the presence of more dissolved substances as more substrate has been converted into AD products. The calculation method discussed in Section 4.5.6 for the corrections due to ionic activity were deemed necessary when ionic strength exceeds 0.2mol/l in the AD (Tait *et al.*, 2012). Thus, the corrections for ionic activity due to the high ionic strength of THP digestion were in fact necessary, where for conventional digestion the ionic strength is lower than the guideline threshold for correction, and so in this study may not be as necessary.

Table 5-2: Struvite precipitation inside digester

	Conventional	**THP**	
Struvite precipitated (in digester)	923	2376	mg/l
Struvite precipitated (in digester)	1187	3324	kgISS/d
Struvite precipitated/PP released	0.67	0.73	kg/kg
Ionic strength	0.15	0.41	mol/l
pK_{sp}'	11.02	10.79	
pIP	11.02	10.79	
Difference: $(pK_{sp} – pIP)/K_{sp}$	0.00%	0.00%	

5.1.5. pH and alkalinity

The pH and alkalinity generated during the digestion process are shown in Table 5-3. These values take into account the release of PP during digestion and the subsequent precipitation of struvite. pH is calculated from equation (4-43) and alkalinity from equation (4-39).

Table 5-3: pH and alkalinity generated

	Conventional	**THP**	Multiple increase	
pH	6.98	7.36	-	
Alk H2CO3	2742	7278	2.7	mg as CaCO3/l
Alk H3PO4	1413	3205	2.3	mg as CaCO3/l
Alk NH4	30	213	7.0	mg as CaCO3/l
Total alkalinity	4186	10698	2.5	mg as CaCO3/l

The THP digestion operates at a pH 7.36 which is higher than that in conventional digestion of pH 6.98. This due to THP digestion operating at a higher solids concentration and subsequently a higher concentration of FSA is released. This FSA which takes a proton from dissolved CO_2 in water forms bicarbonate and increases alkalinity. At a pH≈7 in the digester and with a significant PP breakdown releasing OP the most prevalent species contributing to alkalinity are HCO_3^-, $H_2PO_4^{2-}$, HPO_2^{2-}. The increased production of these substances during operation at a higher loading rate is further reason for the higher alkalinity generated when comparing conventional digestion to that of THP digestion.

However, the pH modelled in THP digestion is lower than that observed in Section 2.7.2 of the literature review of pH 7.5 to 8. This is most likely due to the high PP content of the WAS modelled from NDBEPR activated sludge. A higher PP content releases more OP in the digester. The alk H_3PO_4 increases while the H_2CO_3 decreases, and ultimately the overall alkalinity does not change much. However, the decrease in alk H_2CO_3 results in an increase in carbon dioxide partial pressure, and thus lowers the relative fraction of methane in the biogas as presented in 5.1.7. As digester pH is strongly influenced by alk H_2CO_3 a reduction in this alkalinity causes a reduction of pH (see equation (4-43)).

The breakdown of PP releases metals which allow for metal precipitation, which ultimately lowers pH. In this study struvite precipitation inside the AD was modelled from the magnesium released from PP breakdown i.e. without dosing of Mg, which is often the limiting component for struvite formation in municipal wastes. The stoichiometry of this process given by equation (4-23) consumes HCO_3^- in the formation of struvite, as well as phosphate. The higher the PP content of the WAS then the more magnesium is released in AD, and as a result the greater the amount of struvite precipitated and reduction in alkalinity. The decrease in bicarbonate and phosphate lowers overall alkalinity and results in a lower final digester pH.

5.1.6. Solids reduction and sludge stability

The total solids flux of the influent and effluent are shown for each case in Figure 5-5 and solids data is listed in Table 5-4. For the THP digestion case the influent flux is to the THP+AD combination. Conventional digestion will process 60463kgTSS/d where THP digestion is able to process 153855kgTSS/d in the same digester tank volume. The THP digestion case processes over 2.5 times more sludge and would allow over 90 000kgTSS/d additional sludge to be imported from surrounding WWTW's for beneficiation. Comparing the feed and effluent streams for each case it can be seen how of the biodegradable material was reduced by the AD process, thereby reducing the overall solids flux required for disposal. The change in biodegradable material was converted to biogas and soluble AD products via the stoichiometry given in Section 4.5.3 and 4.5.4. However, some solids were formed in the AD from the precipitation of struvite as discussed in Section 5.1.4.

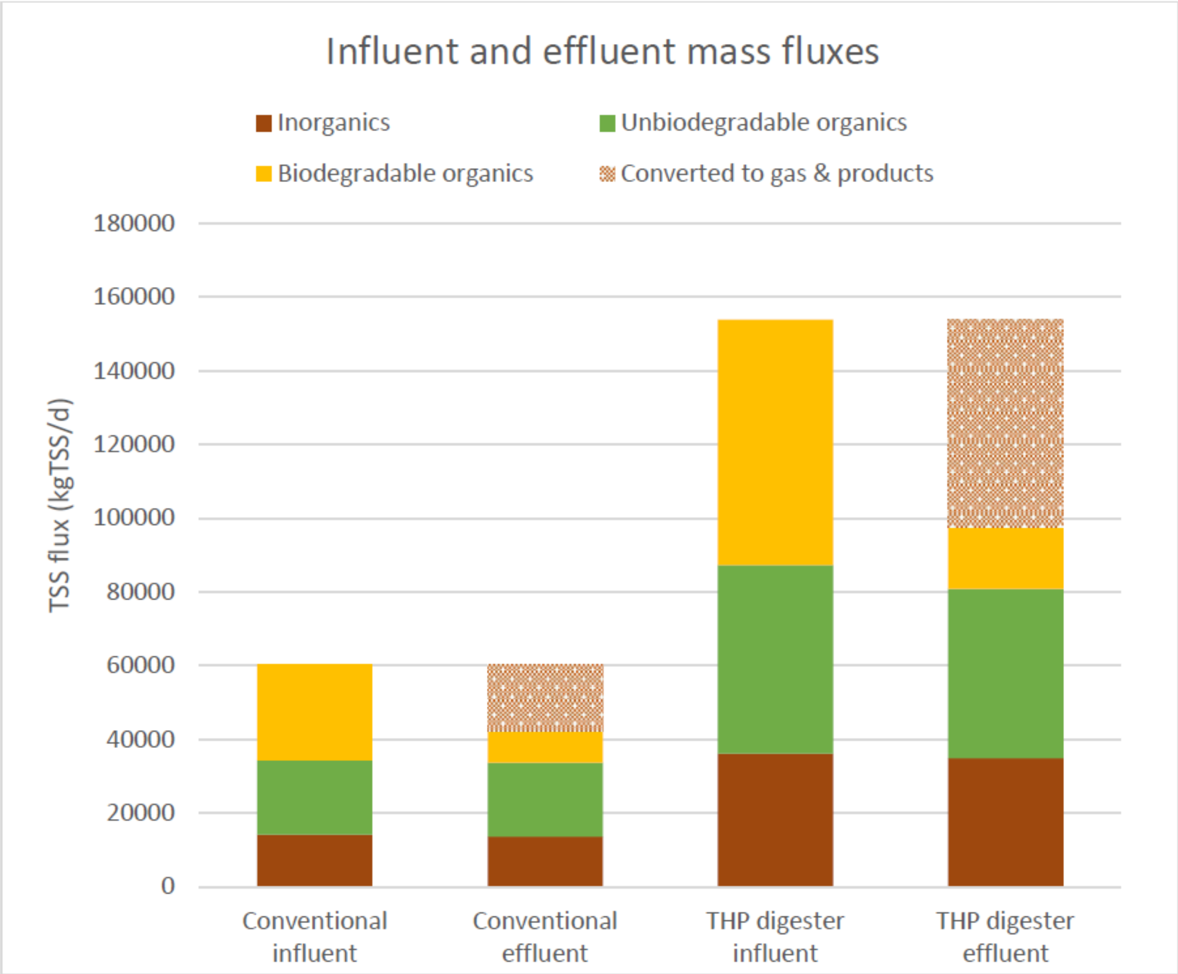

Figure 5-5: Digester influent and effluent mass fluxes

Table 5-4 shows that the reduction in biodegradable organics in THP digestion of 75% was greater than that in conventional digestion of 68%. This was most likely due to the increased rate of digestion, as discussed in Section 4.5.2. The unbiodegradable organics remained unchanged in the case of conventional digestion. However, as discussed in Section 4.3.4 THP increases biodegradability by converting some of the unbiodegradable organics (UO) into biodegradable organic (BO) substances, as well as converting some BO into unbiodegradable soluble organics (USO). Therefore, in THP digestion the change in biodegradable organics of 75% reduction in fact was greater over AD step, as some of the UO was first converted to BO in the THP before entering the digester where the reduction in BO occurred. The inorganic ISS was reduced due to the release of PP (which registers as ISS in solids tests), and even though ISS was generated due to the precipitation of struvite, the overall net impact on ISS was a 4.2% decrease in conventional digestion and 3.4% decrease in THP digestion. The ISS drop in THP digestion was less due to the fact that more struvite was formed per mass of PP released, as discussed in Section 5.1.3.

Table 5-4: Solids breakdown of digester influent and effluent

	Conventional digestion influent	Conventional digestion effluent	% change	THP digester influent	THP digester effluent	% change
	kgTSS/d	kgTSS/d		kgTSS/d	kgTSS/d	
Biodegradable organics	26152	8457	-68%	66545	16574	-75%
Unbiodegradable organics	20108	20108	0%	51166	45909	-10%
Inorganics	14204	13607	-4.2%	36144	34924	-3.4%
Total	60463	42171	-30%	153855	97407	-37%

The guidelines as set by WRC (Herselman and Moodley, 2009) discussed in Section 2.2 state that the treatment of sludge requires a volatile solids reduction of at least 38% on a 90-percentile basis for treated sludge to be classed as a Class 1 stability sludge. This is the highest stability class and allows for more disposal options, as discussed in Section 2.3. Volatile solids reduction is calculated using O'Shaunessy's formula can be used to calculate the volatile solids (VS) reduction in a digester. The below calculation is taken from the WRC guidelines Appendix 2.

$$VS\ reduction\ (\%) = \left(\frac{V_i - V_o}{V_i} - V_i \cdot V_o\right) \cdot 100 \qquad (5\text{-}1)$$

Where,

- V_i = volatile fraction in feed sludge, dimensionless
- V_o = volatile fraction in digested sludge, dimensionless
- The volatile fraction is given by the ratio for each stream $mgVSS.l^{-1}/mgTSS.l^{-1}$

As can be seen in Table 5-5 the volatile solids reduction of both conventional and THP digestion are 36% and 45% respectively. THP digestion comfortably exceeded that required of the WRC guidelines for Class 1 of 38%. However, for conventional digestion the VSR does not comply with Class 1. Further, the VSR for conventional digestion is low compared to that found in Section 2.4.3 of the literature review for mixed sludges of 45%. The VSR in THP digestion is also lower compared to that typically found in literature (see in Section 2.7.3) of 48% to 54%. The lower VSR experienced in this study are most likely due to the influence of ISS in changing the VSS/TSS ratio. In digestion of ND WAS and primary sludge ISS stays relatively unchanged and only VSS is altered by the AD. The digestion of NDBEPR sludge releases PP which reduces ISS, while at the same time precipitating struvite, which again increases ISS. As discussed in Section 5.1.3 the reduction is ISS due to PP release is greater than the subsequent increase due to struvite precipitation. Thus, the net reduction in ISS creates a higher VSS/TSS ratio in the final digested sludge than if ISS had remained unchanged.

The biodegradable fraction of the sludge is the biodegradable organics (XB) divided by the total solids (XT). The biodegradable fraction of the digester effluent has been reduced from 43% to 20% in the case of conventional digestion and 43% to 17% in the case of THP digestion. Note in THP digestion the sludge biodegradability was first increased due to the THP process (discussed in Section 4.3.4) before the AD step reduced the biodegradable organics. Therefore, as discussed previously the reduction in AD biodegradable organics was possibly even better than that stated here. This can be supported by the improved overall COD reduction of THP digestion of 47% compared to that of conventional digestion of 38%. THP also showed a better overall solids reduction of 37%, mostly due to a reduction in organics shown in Table 5-4, where conventional digestion achieves a 30%. However, for the purposes of sludge stabilisation the final sludge mass's biodegradable fraction is the focus and THP digestion still provided a lower final sludge biodegradability. The lower the biodegradable fraction then the more suited the sludge is for application to land, and thus the final sludge from THP digestion is of a more stable quality than that from conventional digestion.

Table 5-5: Solids treatment and stabilisation

	Conventional	**THP**	
Solids flux			
Influent solids flux	60463	153855	kgTSS/d
Effluent solids flux	42171	97407	kgTSS/d
Influent VSS/TSS (V_i)	0.77	0.77	
Effluent VSS/TSS (V_o)	0.68	0.64	
Volatile solids reduction (O'Shaunessy's formula)	36%	45%	
Total solids reduction	30%	37%	
Biodegradability			
Influent biodegradable fraction to digester	43%	43%	XBi/XTe
Effluent biodegradable fraction from digester	20%	17%	XBe/XTe
COD			
Influent COD load	68	174	tonCOD/d
COD reduction	38%	47%	

5.1.7. Gas production

The results of the gas production focus on the amount of methane produced per mass unit of volatile solids, and not total solids. This is because the volatile solids are the focus of sludge treatment via anaerobic digestion, to achieve a reduction of organic solids and stabilisation of the remaining solids.

Figure 5-6 shows the total biogas production for each case. The quantity of gas produced is significantly greater in THP digestion 49666Nm^3/d compared to 15374Nm^3/d in conventional digestion. While this difference is partially due to the THP increasing biodegradability of the WAS (and thus gas production) most of the increase is due to the higher digester loading rate allowing more sludge to be processed, and therefore more biogas is produced.

The methane fraction of gas is similar in both cases, 64% in conventional digestion and 62% in THP digestion. As the stoichiometry of the biological processes in conventional and THP digestion follow a similar development as discussed in Section 4.5 one can expect that the biogas make-up of each case will be similar. However, the percentage methane should not be taken as an indication of the amount of methane produced. It is shown in Figure 5-7 more methane is produced per specific VSS mass in THP digestion than conventional digestion. Rather, the % methane in the biogas is largely impacted by the carbon dioxide partial pressure, which is a function of carbonate alkalinity and pH as shown by equation (4-43). In the case of THP digestion mineral precipitation decreases pH and increases the CO_2 partial pressure of the system, as discussed in 5.1.5. Therefore, more CO_2 is made per ton of VSS in THP digestion to the extent that the fraction of methane in the biogas is reduced, when compared to conventional digestion.

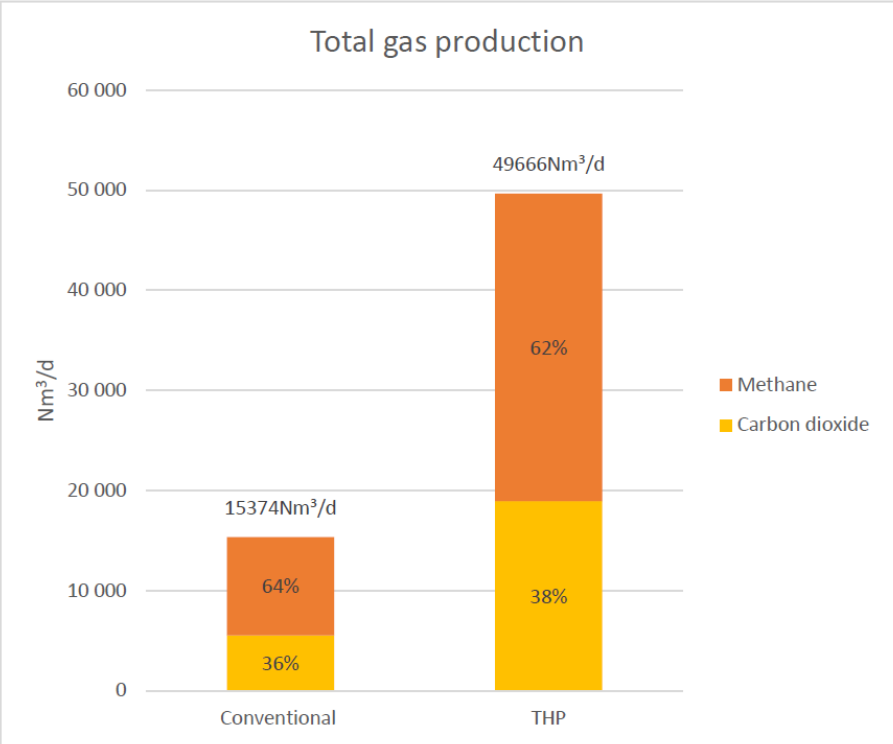

Figure 5-6: Relative gas production

Figure 5-7 shows that per unit mass of volatile solids fed to digestion THP digestion makes more useful methane at 275Nm^{3}CH$_4$/tonVSS$_{ied}$ compared to that of conventional digestion of 213Nm^{3}CH$_4$/tonVSS$_{fed.}$ This implies that in the case of THP digestion there has been an improved COD conversion from organics to methane. This is expected from Section 4.3.4 and 4.5.2 where the effects of THP have been applied in increasing the rate of digestion and increasing biodegradability or organics (within the digester residence time.) This is supported by the modelled output of a greater percentage reduction in COD and greater reduction volatile solids as discussed in Section 5.1.6.

The specific methane production for conventional digestion 213Nm^3/kgVSS$_{fed}$ is slightly lower than that mentioned in the literature review of 230Nm^3/kgVSS$_{fed}$. This is most likely because the sludge mixture in this study is 40/60 PS/WAS where that the literature value contained a greater fraction of PS in a 50/50 PS/WAS mixture. As shown in Table 4-8 PS tends to have a greater biodegradable fraction of 0.7 than WAS at 0.47 which explains why when increasing the fraction of WAS in the AD feed less methane is produced.

The specific methane production of the THP digestion case of 275Nm^3/tonVSS$_{fed}$ was lower than that found in Section 2.7.3 of the literature review of 290-340Nm^{3}CH$_4$/tonVSS$_{fed}$. This is because the digester in this exercise operated at a shorter SRT of 13.5 days as shown in Table 4-2 than those in the literature references of 20-28days shown in Table 2-3. Although a shorter SRT was applied it was still within that found in literature of full scale plants of 10-15days (Barber, 2016). This means the AD biomass had less time to process the biodegradable COD into AD products and thus less methane was made. While the VSS loading for THP digestion is the same as that in literature the digestion of NDBEPR sludge with high PP content will result in a shorter SRT due to the presence of increased ISS. This decreases the VSS/TSS ratio of the sludge and to get the same VSS throughput a higher TSS loading rate must be applied. To do this a greater feed flow is required. For a flow-through type digester the hydraulic retention time (HRT) is the same as the SRT, and increasing flow decreases HRT and therefore SRT is reduced.

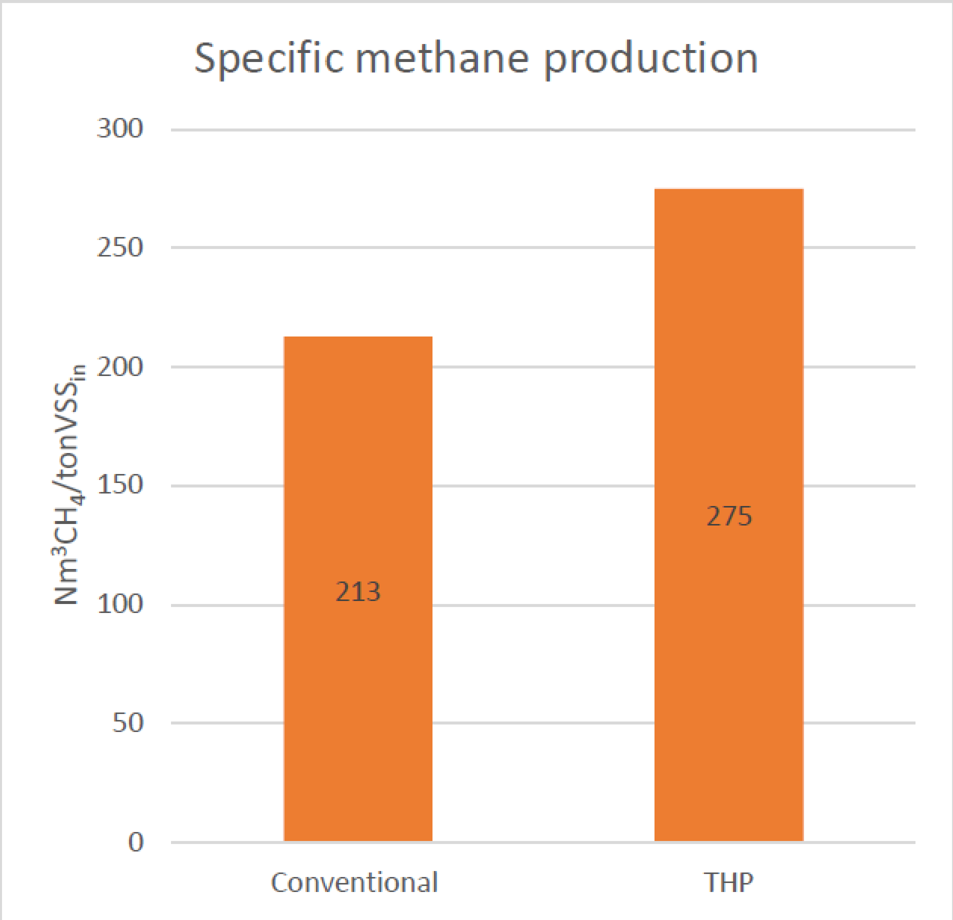

Figure 5-7: Specific methane production

Both PP release and struvite production release CO_2 as shown in stoichiometry in sections 4.5.4 and 4.5.5 respectively. In both cases the release all the PP contained in their feed WAS occurred, as discussed in Section 5.1.3. However, in Section 5.1.4 it was found THP digestion produces more struvite per mass of PP released, compared to conventional digestion, and thus would produces more CO_2 per digested mass. This is illustrated in Figure 5-8 where conventional digestion produces $120Nm^3/tonVSS_{fed}$ and THP digestion produces $170Nm^3/tonVSS_{fed}$.

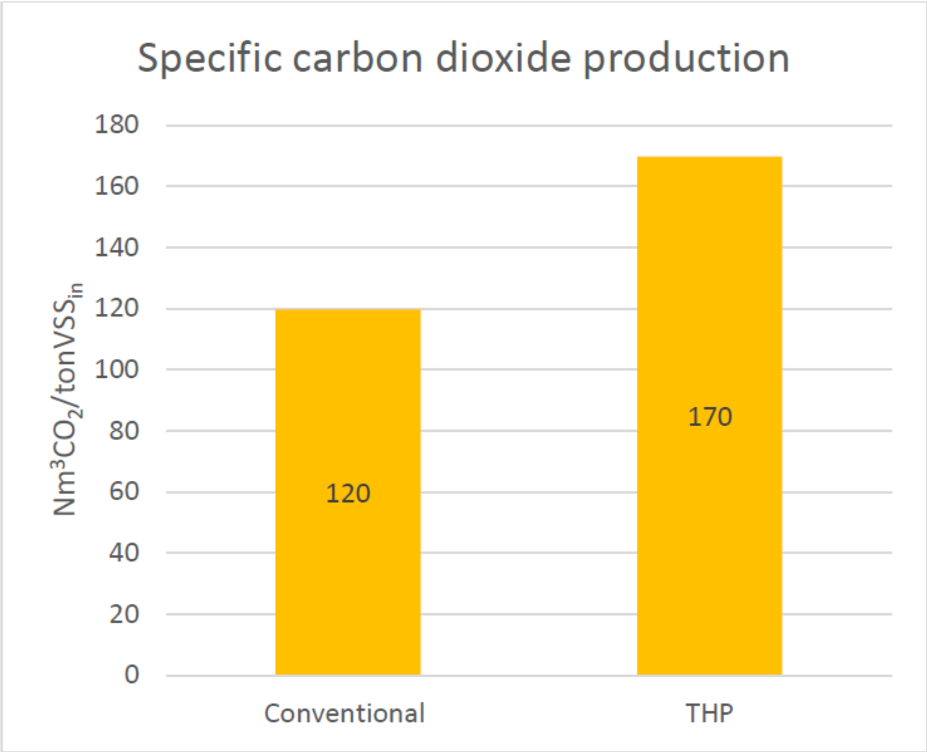

Figure 5-8: Specific carbon dioxide production

5.2. MASS BALANCE OUTPUTS

Mass flux balances were carried out over the modelled process (boundaries as shown in Section 4.1) for COD, TOC, TN and TP. These were carried out according to the calculation discussed for each component in Section 4.6. The concentrations were multiplied by the digester flow rate to generate fluxes. As shown in Table 5-6 it can be seen that mass balances were achieved for each component, thereby serving as a check that all material was accounted for over the modelled process.

The first row in Table 5-6 shows filtered organic concentrations. These represent soluble organic substances in the digester effluent. The soluble COD in THP digestion of 796mg/l is significantly higher than that of conventional digestion of 47mg/l, even though it was shown in Table 5-4 and Table 5-5 that THP digestion achieves a better organics reduction and overall removal of COD than conventional digestion. The soluble COD shown in Table 5-6 is the unbiodegradable soluble COD generated as a by-product of the THP process, as determined from Section 4.5.1. This COD will simply pass through the AD and subsequent side-stream treatment steps to join the wastewater flow in the adjoining WWTW.

Table 5-6: Effluent concentrations and mass flux balances

	Conventional	THP	Conventional	THP	Conventional	THP	Conventional	THP
Effluent concentrations	COD	COD	TOC	TOC	TN	TN	TP	TP
Filtered organic (mg/l)	47	796	16	271	1	52	0	15
Particulate organic (mg/l)	32413	65289	11440	23001	1437	2608	604	1179
Total organic (mg/l)	32460	66085	11456	23272	1438	2660	604	1194
Filtered inorganic (mg/l)	0	0	663	1751	914	2793	624	1646
Particulate inorganic (mg/l)	0	0	0	0	53	136	117	301
Gas (mg/l)	20243	58457	5965	17724	0	0	0	0
Total (mg/l)	52703	124543	18084	42747	2405	5589	1344	3141
Flux balances								
In (kg/d)	67800	174196	23264	59790	3094	7817	1729	4393
Out gas (kg/d)	26042	81763	0	0	0	0	0	0
Out effluent (kg/d)	41758	92432	23264	59790	3094	7817	1729	4393
Balance (Out/In)	100%	100%	100%	100%	100%	100%	100%	100%

5.3. EVALUATION OF OPERATING COSTS

This section compares the operating costs using the methods discussed in Section 3.3.

5.3.1. Heating energy

The heating required for each case was calculated from the equations and methodology discussed in Section 3.3.4 and is presented in Table 5-7. In the case of conventional digestion 1323kW of heating is required. This heat can be supplied from hot water through a heat exchanger circulating from CHP and will require a flowrate of 37 968kg/h (≈38m^3/h). For THP unit operation 4874kg/h of steam is required from the CHP waste heat boiler to supply 3615kW of heat.

For both cases CHP is able to provide the necessary heat required by the process. It must be pointed out that in the case of THP production heat energy is required as steam. This requires a heating temperature in excess of the of LGH temperature, which means only a portion of the LGH would be utilised e.g. to pre-heat boiler feed water. However, the HGH required for THP digestion can be recovered from the CHP's HGH recovery heat exchanger. The THP HGH heating requirement of 2794kW can be satisfied by the 2967kW available as HGH from CHP. The balance of low-grade heat required can be supplied with residual HGH from the CHP and the LGH from CHP. In conventional digestion the 1323kW of heating required is only LGH and can be satisfied from the combination of 950kW HGH and 950kW LGH, as both of these temperatures are above that required for digester heating. As shown below no external fuel sources are required to supply heat and the process in each case is self-sustaining in terms of energy.

Table 5-7: Heating requirements

	Conventional	**THP digestion**	
Heat required			
High grade heat (HGH) required	-	2794	kW
Low grade heat (LGH) required	1323	821	kW
Total heat required	1323	3615	kW
Heating medium	Water (85^{o}C supply; 55^{o}C return)	Steam (600kPag)	
Flow of heating medium	37968	4874	kg/h
Heat recovered from CHP			
HGH	950	2967	kW
LGH	950	2967	kW
Heat available	1901	5934	kW
Heat deficit	0	0	kW
Heat required form fuel	0	0	kW

5.3.2. Polyelectrolyte consumption

The values presented in this section were determined using the methods in Section 3.3.7.

The polyelectrolyte consumed during operation in each case are presented in Figure 5-9. As can be seen the THP digestion case uses significantly more polyelectrolyte than conventional digestion. In the final dewatering step for each case THP digestion uses 1743kg/d where conventional digestion uses 390kg/d. This can be expected due to THP digestion having a greater solids flux requiring dewatering as shown in Section 5.1.6 from both a higher throughput of feed sludge processed (leaving a greater residual amount of solids), and as a result a greater amount of struvite precipitated, both inside the digester and during P-treatment. A further contributor to the greater polyelectrolyte consumption in THP digestion is due to the fact that the dewatered sludge cake in THP digestion is dewatered to dryer quality of 30%DS compared to 22%DS in conventional digestion, which requires more polyelectrolyte. THP digestion also has an additional pre-dewatering step where sludge is thickened to a 17%DS cake prior to feeding to THP. This pre-dewatering step requires a polyelectrolyte consumption of 212kg/d of polyelectrolyte, extending the difference between the consumption of the two cases.

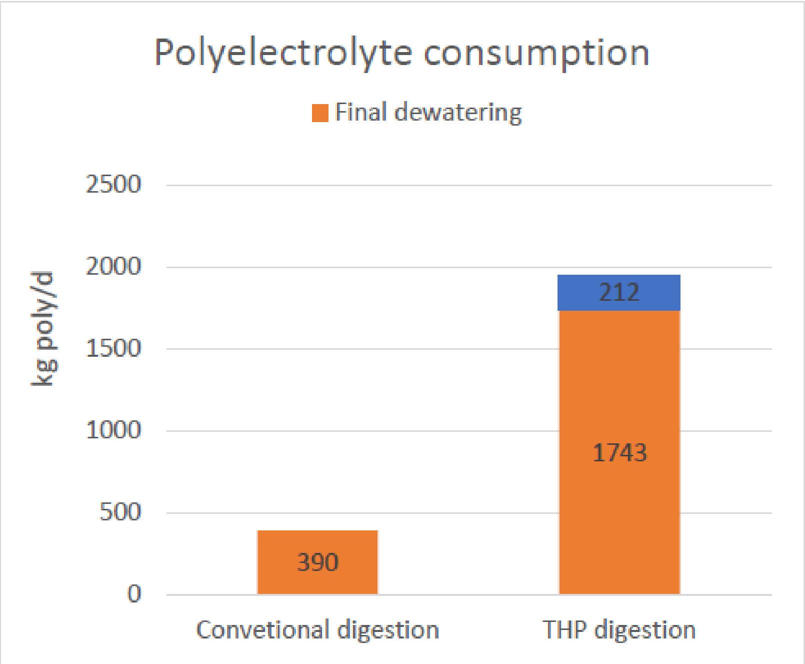

Figure 5-9: Polyelectrolyte consumption

Table 5-8 shows the polyelectrolyte consumed and the corresponding dewatering performance. This then uses the polyelectrolyte consumption and specific cost to determine the total annual operating expense. For conventional digestion the annual polyelectrolyte expense is R5 719 698. For THP digestion the pre-dewatering annual polyelectrolyte expense is R3 103 572, which is added to the THP digestion final dewatering cost of R25 561 921 to total R28 665 494. These costs are illustrated in Figure 5-10.

Table 5-8:Polyelectrolyte consumption for sludge dewatering

	Final dewatering conventional digestion	**Pre-dewatering THP digestion**	**Final dewatering THP digestion**	
Solids flux to dewatering	48751	60463	116198	kgTSS/d
Capture rate to cake	99.50%	99.50%	99.50%	
Cake dryness %	22%	17%	30%	
Dewatering liquor TSS concentration	229	324	573	mgTSS/l
Specific polyelectrolyte usage	8	3.5	15	kg poly powder/tonTSS
Polyelectrolyte consumed	390	212	1743	kg dry powder/d
Specific polyelectrolyte cost	R 40.18	R 40.18	R 40.18	R/kg poly
Polyelectrolyte cost	R 5 719 698	R3 103 572	R 25 561 921	R/annum

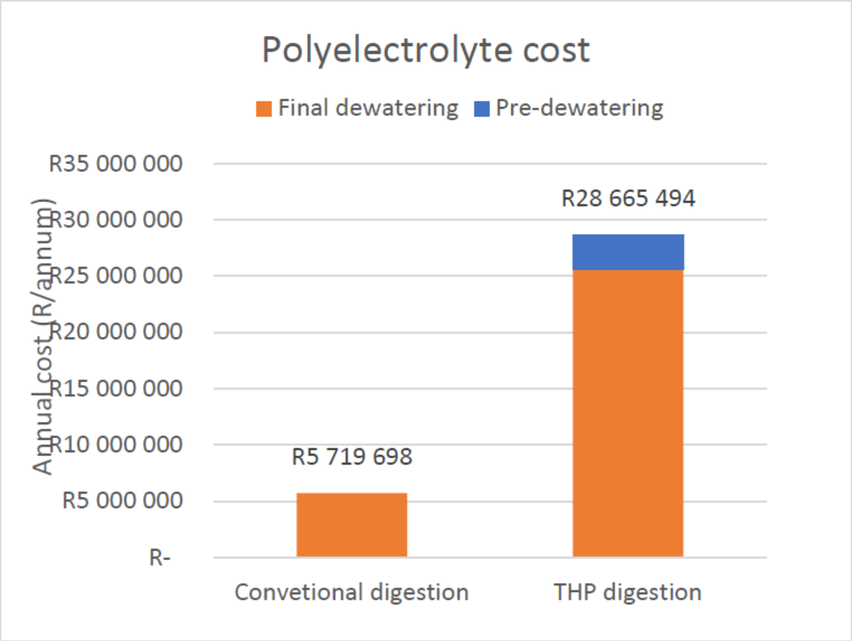

Figure 5-10: Polyelectrolyte costs

5.3.3. Electrical energy recovery

The results for electrical energy consumption and production are discussed in this section. The electrical energy for the plant operation (for all unit process except N-treatment) was determined from Section 3.3.2 along with the power generated from CHP. The N-treatment power requirement was calculated on its own Section in 3.3.7.

The energy consumed by each process is compared in Figure 5-11. This is split into the energy required for the digestion processes, including all ancillary unit processes in each case, and into N-treatment energy. The combined total is then shown against the electrical energy generated from CHP. The CHP electrical output is shown in red. The net energy surplus (or deficit) electrical energy is also shown alongside in blue. The net energy is output from CHP less the total power required to run all processes. It can be seen that both processes are self-sufficient with regards to electrical energy, and produce a net surplus. This surplus can be used in the Cape Flats WWTW (e.g. to supplement the aeration blowers and reactor pumps and mixers) or sold to external users.

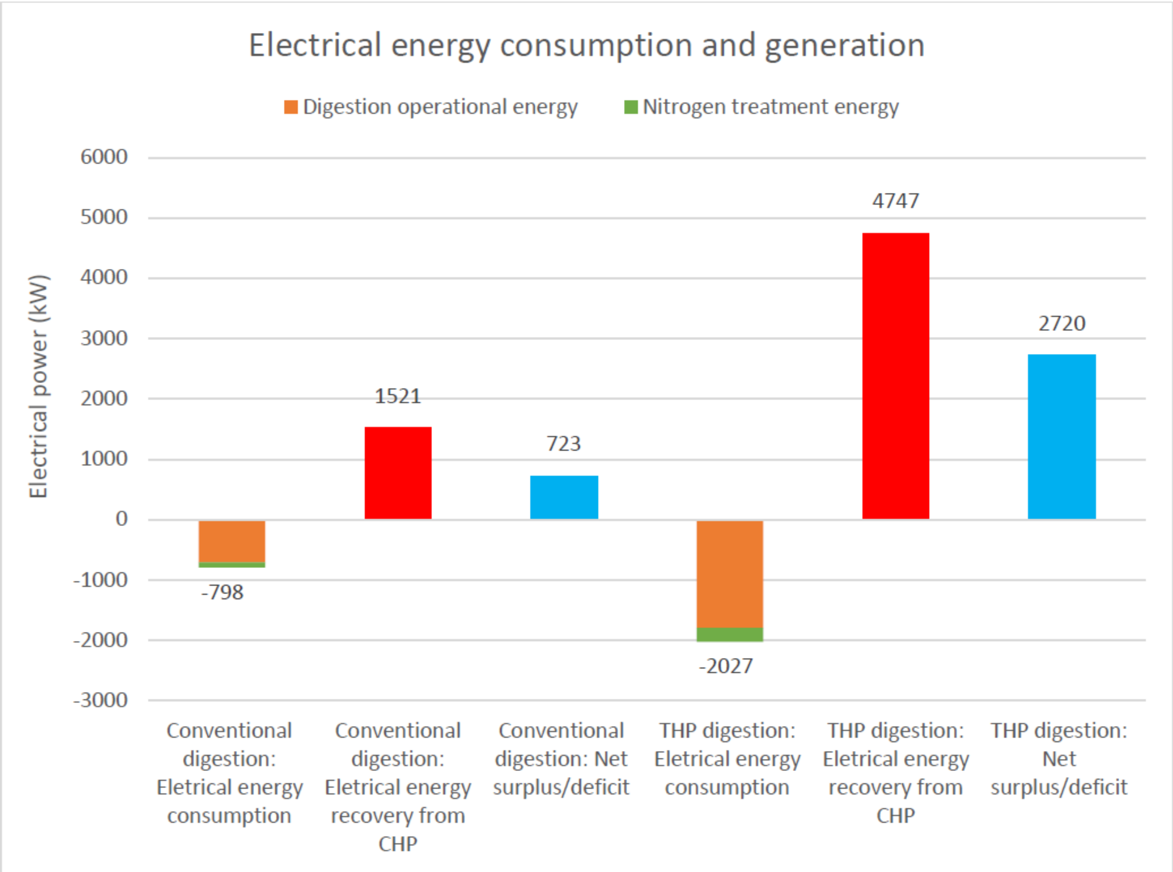

Figure 5-11: Electrical energy consumption and generation

In Table 5-9 the contribution for each electrical energy consumer is shown. Using an average energy cost of R1.02/kWh the total saving or expense for electrical power is then calculated. It can be seen in both cases there is a surplus which results in an operating saving of R6 459 739 for conventional digestion and R24 3030 920 for THP digestion.

Table 5-9: Electrical energy consumption and generation

	Conventional	**THP digestion**	Increase	
Operating energy	705	1795	154%	kW
Nitrogen treatment energy	92	232	152%	kW
Total electrical energy required	798	2027	154%	kW
Electrical energy from CHP	1521	4747		kW
Net surplus energy	723	2720	276%	kW
Parasitic load	52%	43%		
Cost	R 1.02	R 1.02		R/kWh
Saving	R 6 459 739	R 24 303 920		R/annum

While operating power increased 154% from conventional digestion (705kW) to THP digestion (1795kW) the net surplus electrical energy generated increase by 276%. This shows a clear energy recovery improvement in the case of THP digestion. This increase is due to the THP digestion's case having a lower parasitic energy fraction drawn to run the process.

5.3.4. Phosphorous treatment

This section shows the results generated from the review in Section 3.3.6 for struvite precipitation as P-treatment to treat the high load of OP coming from digestion. While struvite precipitates on its own inside the digester (as discussed in Section 4.5.5 and 5.1.4) this section looked at the forced struvite precipitation done by dosing a magnesium source to the digester effluent downstream of AD to reduce the OP load returned to the WWTW. This form of P-treatment generates struvite precipitation under controlled conditions and has an operational benefit in ensuring precipitation does not occur spontaneously in the unit processes downstream of the AD, which has been an issue at the Cape Flats AD facility. Therefore, in this study the struvite precipitation process is done upstream of final dewatering rather than on the dewatering liquor, and thus reduces the risk of blocked pipes and damage to dewatering equipment.

The MgO dosage required for P-treatment is presented in Table 5-10 below. To remove 85% of the OP from the digester effluent and achieve the effluent OP concentrations after P-treatment (as presented here and in Section 5.1.2) THP digestion required a 287% greater daily usage of MgO over conventional digestion. For conventional digestion an MgO dosage of 1187kg/d was required in conventional digestion and 3404kg/d for THP digestion. From this dose and the price of MgO of R11.57/kg at a 96% purity the total cost for treatment was calculated. For conventional digestion would cost R5 010 814 per annum and for THP digestion this would cost R14 374 789 per annum.

The relative MgO dosages are illustrated in Figure 5-12. It can be seen that the THP digestion case requires significantly more MgO dosing to achieve the same 85% reduction in OP. This is due to THP digestion having a much higher P-treatment requirement as shown in Figure 5-4. As a result the struvite production in THP digestion is significantly greater than that in conventional digestion. This can be seen in Figure 5-13.

Table 5-10: Phosphorous treatment costs and quantities

	Conventional digestion	**THP digestion**	
Digester effluent OP concentration	624	1646	mgP/l
Stoichiometric dosing ratio	1.1	1.1	molMg:molP
Purity of MgO	96%	96%	
Dose of MgO required	922	2434	mgMgO/l
Dose of MgO required	1187	3404	kg/d
OP removal efficiency	85%	85%	
P-treatment effluent OP concentration	94	247	mgP/l
OP removed	682	1957	kgP/d
Struvite mass flux before P-treatment	1187	3324	kg/TSS/d
Struvite mass flux after P-treatment	6579	18792	kg/TSS/d
MgO cost	R 11.57	R 11.57	R/kg
Cost of P-treatment	R 5 010 814	R14 374 789	R/annum

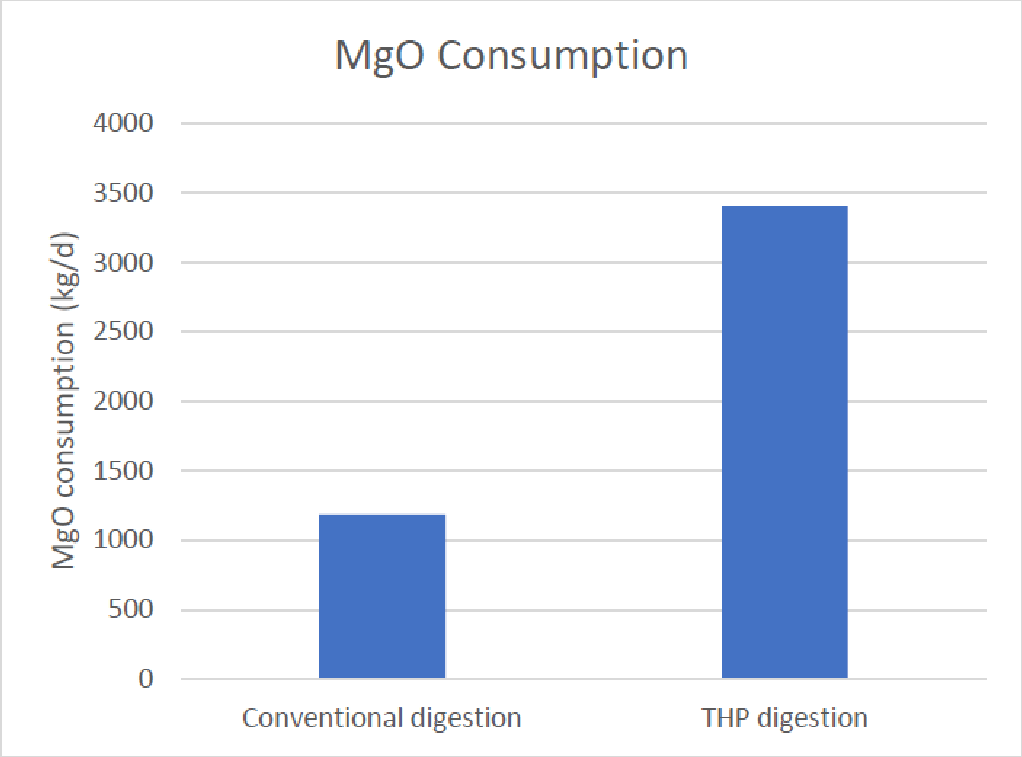

Figure 5-12: Magnesium hydroxide (MgO) consumption for P-treatment

Figure 5-13 shows the struvite precipitation in the digester effluent flow as it moves through the plant. Struvite is initially precipitated in the digester due to PP breakdown making magnesium available to combine with the FSA and OP inside the digester, as discussed in Section 5.1.4. In P-treatment of the digester effluent significantly more struvite is then precipitated in the P-treatment step, adding to the struvite mass flux headed to final dewatering.

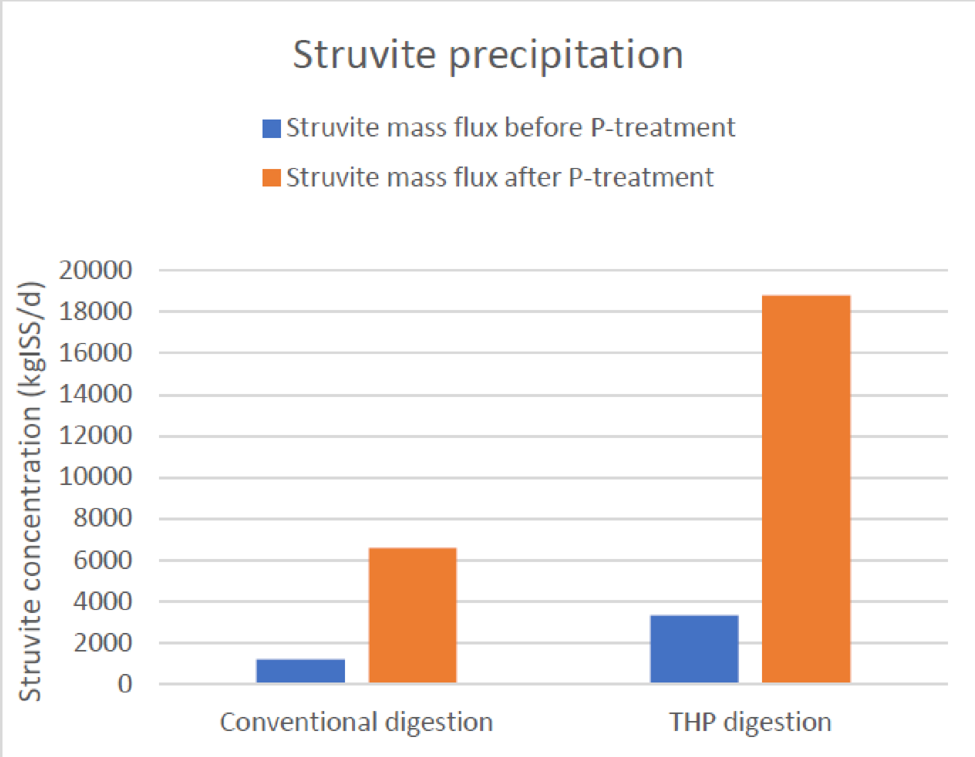

Figure 5-13: Struvite precipitation

The use of struvite precipitation in P-treatment has the added benefit of lowering the final dewatered sludge VSS/TSS ratio due to the production of struvite ISS. Following the discussion in Section 5.1.6 this further improves the stability class of the of the final sludge by increasing the VSR. This is shown in Table 5-11. For conventional digestion the P-treatment step improves the VSR from 36% to 57%. This is significant as a VSR of 38% is required for a Class 1 stability sludge, as discussed in Section 2.3.1. Therefore the P-treatment step would improve the stability class for conventional digestion. For THP digestion the VSR improves from 45% to 64%.

Table 5-11: Solids fluxes and VSR after P-treatment

	Conventional digestion influent	Conventional digestion effluent	Conventional digestion P-treatment effluent	THP digester influent	THP digester effluent	THP digester P-treatment effluent
Solids flux						
Biodegradable organics (kgTSS/d)	26152	8457	8457	66545	16574	16574
Unbiodegradable organics (kgTSS/d)	20108	20108	20108	51166	45909	45909
Inorganics (kgTSS/d)	14204	13607	20186	36144	34924	53715
Total (kgTSS/d)	60463	42171	48751	153855	97407	116198
Solids fractions						
%dry solids	4.7%	3.3%	3.8%	11%	7.0%	8.3%
VSS/TSS	0.77	0.68	0.59	0.77	0.64	0.54
VSR	-	36%	57%	-	45%	64%

5.3.5. Sludge disposal

This section shows the sludge generated for disposal as determined from Section 3.3.5.

The sludge disposal quantities after final dewatering are shown in Figure 5-14. These values are given in tonnes of dewatered sludge cake per day, which include both the TSS content and moisture content of the sludge for disposal. Assuming the bulk density of the sludge cake is the same as water then these numbers can also be taken as the sludge volume in m^3 requiring disposal every day. These include struvite precipitated from inside the digester and the struvite precipitated in the P-treatment step, shown in green as “struvite”.

The combination of the grey and greens areas of Figure 5-14 in each case represents the sludge leaving the Cape Flats digestion facility, which is a mixture of digested TSS and struvite. It can be seen that THP digestion results in more final digested sludge and struvite for disposal. However, in the case of conventional digestion significantly more sludge requires disposal if one considers the untreated PS and WAS from surrounding WWTW’s is not able to be treated. In the THP digestion case this PS and WAS would be able to processed due to the increased capacity of THP digestion. This means that in conventional digestion 756 tonnes of dewatered sludge will require disposal each day (made up of digested sludge from Cape Flats and the untreated PS and WAS from surrounding WWTWS). Of this value the majority is made up of untreated primary sludge and WAS. In THP digestion 385 tonnes per day require disposal and this is made up of only sludge from Cape Flats. This is a combination of stabilised digested sludge and struvite. The result is the THP digestion case requires roughly approximately half the volume of sludge to be disposed of. A major benefit is the final sludge from THP digestion can be disposed of in many more ways than that from conventional digestion, due to improved stabilisation and pathogen elimination. Ultimately, the difference between the two cases is 371 tonnes per day of more sludge requiring disposal in the conventional digestion case.

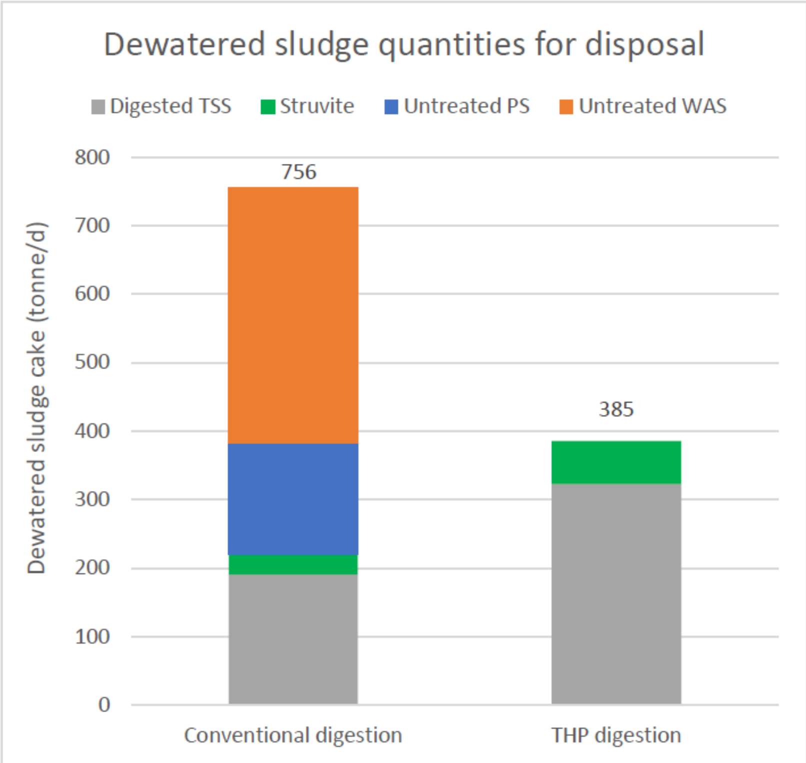

Figure 5-14: Dewatered sludge quantities for disposal

The sludge TSS fluxes to disposal are shown for each type of sludge in Table 5-12. Along with these are each sludge's dewatered cake quantities at the corresponding percentage dryness, given as wet tonnes per day (wet ton/d). This value is then used with the rate for transport and disposal to calculate the disposal cost for each sludge type. These costs are then presented graphically in Figure 5-15. The total sludge disposal cost in conventional digestion is the sum of the disposal cost of the untreated (i.e. undigested) PS of R51 293 660 and WAS of R18 366 278 and the digested sludge (incl. struvite) of R10 840 102. This gives a total sludge disposal cost of R80 500 400 for conventional digestion. This is significantly greater than the R7 487 354 for THP digestion (which includes only digested sludge and struvite). The cost in THP digestion could possibly be lowered even further if the final sludge were to be sold as biosolids for agriculture. However, the market value is not yet known and so to be conservative sale of sludge was excluded from this study. Please refer to Table 9-4 and Table 9-5 in the appendix for the characteristics of the final sludge for disposal.

Table 5-12: Sludge disposal quantities and costs

-	**Conventional digestion**			**THP digestion**
Sludge type to disposal	Primary	WAS	Conventionally digested	THP digested
Dryness	23%	15%	22%	30%
Amount of sludge (tonDS/d)	37	56	49	116
Amount of sludge (wet ton/d)	162	374	220	385
Destination for disposal	Landfill	Agriculture (restricted)	Agriculture (restricted)	Agriculture (unrestricted)
Disposal+transport cost	R 865	R 135	R 135	R 53
Cost (R/annum)	R 51 293 660	R 18 366 278	R 10 840 102	R 7 487 354

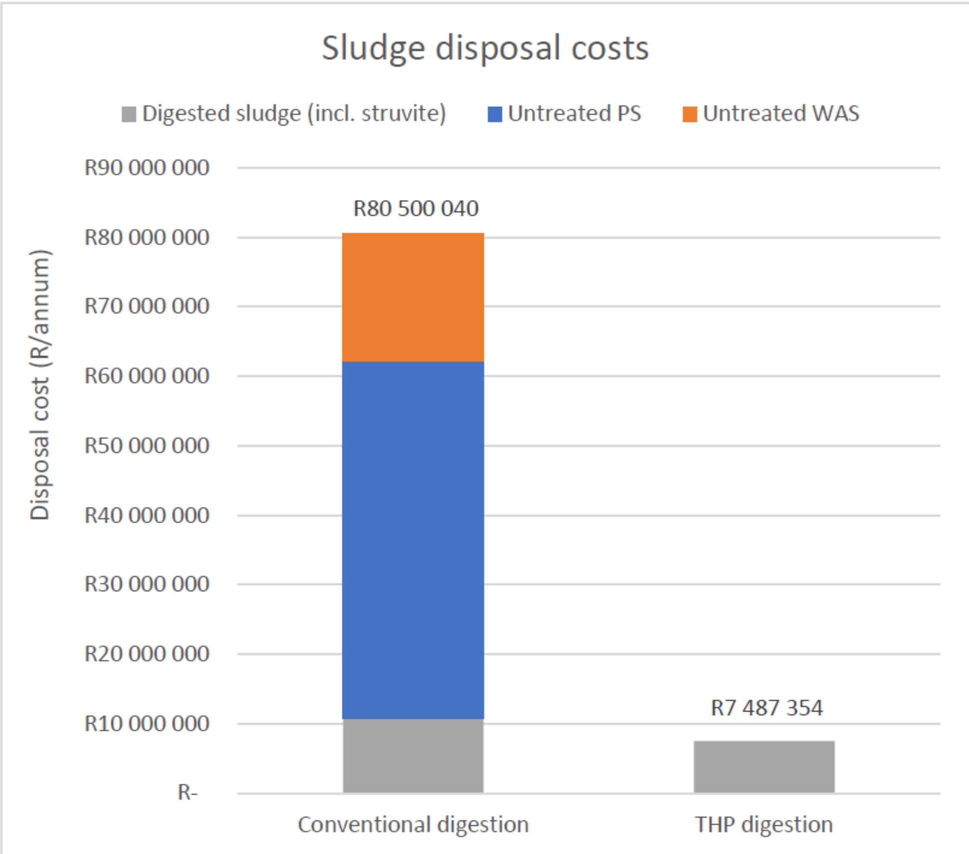

Figure 5-15: Sludge disposal costs

5.3.6. Carbon emissions credits

The results for carbon emissions were generated from the methodology listed in Section 3.3.8. In addition to the economic benefits presented here this section also seeks to provide a degree of environmental significance between the two cases.

The carbon emissions savings for each case is presented in Figure 5-16. Positive values represent revenue from the sale of carbon credits and negative values are emissions are from the combustion of methane in the generation of renewable energy from CHP. It can be seen from the emissions savings sludge diverted from landfill is the most significant source of carbon credits for both conventional digestion and THP digestion. Renewable energy also creates an emissions saving and is an additional source of carbon credits. The renewable energy recovered through CHP refers to electrical power only and excludes the heat energy recovered (heat energy is discussed in Section 5.3.1). Both cases generate emissions from the combustion of methane in CHP, but this is small relative to the emissions savings. The result is a net production of carbon credits. In comparing the two cases it is shown that seen that THP digestion results in significantly more emissions reductions and thus brings with it an environmental benefit over conventional digestion. For THP digestion the net emissions reduction is over 3 times higher than that of conventional digestion.

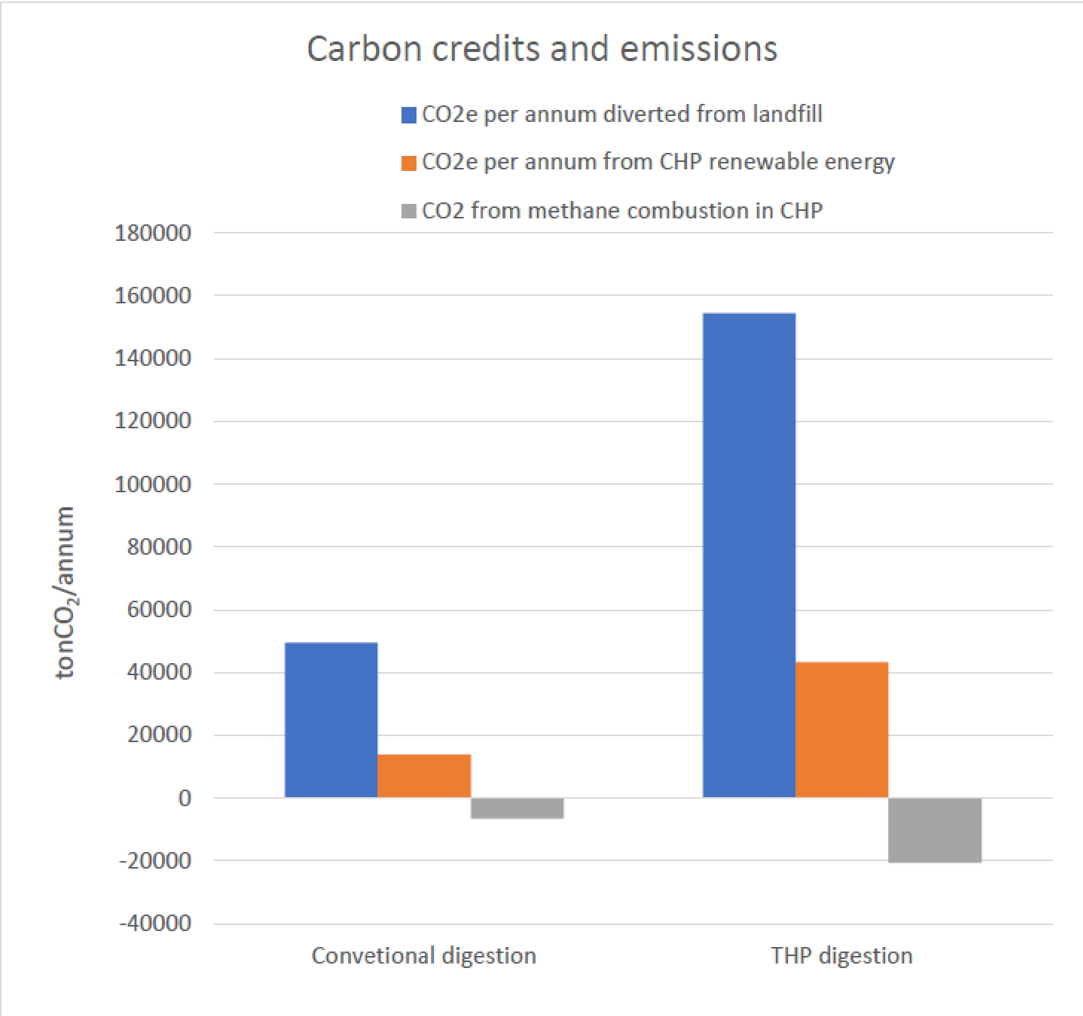

Figure 5-16: Carbon credits and emissions

Table 5-13 shows the contribution to the net carbon credits for each case. For conventional digestion a net production of 56750tonCO_2e/annum is generated and for THP digestion a net production of 177171tonCO_2e/annum is gernated. Using a market related sale rate of R102/tonCO_2e by selling 15% below the tax rate of R120/tonCO_2e (as discussed in section 3.3.8) a total revenue of R5 788 545/annum can be generated for conventional digestion and

R18 071 413/annum can be generated in THP digestion. This income streams are used to offset other operational costs in each case.

Table 5-13: Carbon credits revenue

	Conventional	THP	
CO2e per annum diverted from landfill	49470	154441	tonCO2e/annum
CO2e per annum from CHP renewable energy	13853	43248	tonCO2e/annum
CO2 from methane combustion in CHP	-6572	-20517	tonCO2/annum
Net carbon credits	56750	177171	tonCO2e/annum
Tax rate	R 120	R 120	R/tonCO2e
Discount to sale	15%	15%	
Sale rate	R 102	R 102	R/tonCO2e
Saving	R5 788 545	R 18 071 413	R/annum

5.3.7. Comparison of operating costs and savings

Figure 5-17 shows a comparison of the major operating expenses and savings for each case. This section gives stakeholders considering THP digestion insight to what the major operating expenses and saving are and quantifies these for the case study in this research.

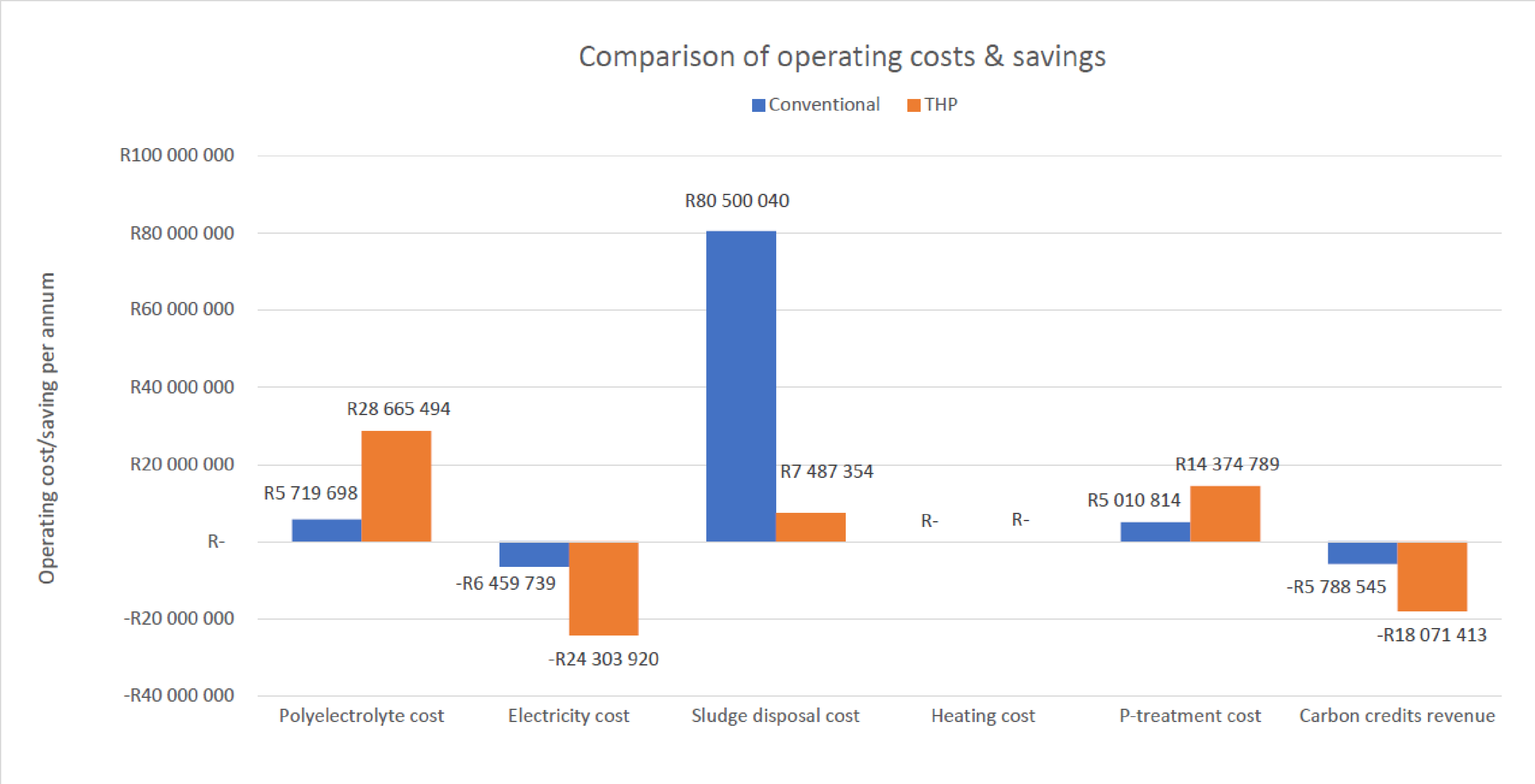

Figure 5-17: Comparison of operating costs & savings

A summary of each operating cost is compared for the two cases in Table 5-14. The negative operating cost for electricity is a net surplus of electrical power for use within the adjoining WWTW or export, and for carbon credits the negative value represents a revenue stream. The difference between the totals shows an annual operating cost difference of R70 829 964. This difference is the annual reduction in operating costs that an investment in a THP digestion facility would bring over simply maintaining and reinstating the existing conventional digestion facility.

Table 5-14: Operating cost comparison

	Conventional digestion	**THP digestion**	
Polyelectrolyte cost	R 5 719 698	R 28 665 494	R/annum
Electricity cost	-R 6 459 739	-R 24 303 920	R/annum
Sludge disposal cost	R 80 500 040	R 7 487 354	R/annum
Heating cost	R -	R -	R/annum
P-treatment cost	R 5 010 814	R 14 374 789	R/annum
Carbon credits revenue	-R 5 788 545	-R 18 071 413	R/annum
Total (net)	R 78 982 267	R 8 152 303	R/annum
Difference	R -	-R 70 829 964	R/annum

The discussion of the overall cost analysis in evaluating THP digestion over conventional digestion follows.

- *Polyelectrolyte for dewatering:* Polyelectrolyte for dewatering proved to be a major cost, especially in the case of THP digestion. This can be expected in that THP has more extensive dewatering requirements, both upstream of the THP and in final dewatering. Upstream the THP requires feed sludge to be dewatered to a sludge cake. In final dewatering the THP digestion case the process runs at a higher throughput and therefore has a greater residual solids load requiring final dewatering.
- *Electrical power generation:* In each case a net electricity saving was achieved, which implies that each facility can generate sufficient power from CHP engines to run all of its equipment. The net electrical surplus generated the greatest savings in this study, followed by carbon credits.
- *Heating:* The heating cost for each case is zero. This is because sufficient heat is recovered from CHP to satisfy the process heating requirements in each case. This zero cost implies no excess fuel is required for combustion in a supplementary boiler.
- *Sludge disposal:* The most significant cost is that of sludge disposal. The sludge disposal cost for conventional digestion of R80 500 040 is significantly higher than that of THP digestion R7 487 354. This can be expected as the conventional digestion case is not able to treat as much sludge as THP digestion, and therefore more untreated sludge requires disposal. Further, the sludge remaining after THP digestion is of a

higher quality than that after conventional digestion and thus allows for less costly disposal options.

- *Nutrient treatment:* Phosphorous treatment in the case of THP digestion requires a significantly higher orthophosphate load to be treated. This in turn requires a greater consumption of treatment chemicals and therefore results in a higher operating cost. This can be expected as the higher throughput of THP digestion over conventional digestion results in the breakdown of more organics and PP and the subsequent release of nutrients. A similar observation can be made in Section 5.1.1 for the nitrogen load requiring treatment which is higher in THP digestion and therefore resulting in a higher operating cost. Both major nutrients, N and P, therefore show the requirement of more treatment in THP digestion over conventional digestion and therefore result in a higher operating cost. However, in both cases good nutrient treatment can be achieved so that no additional strain is placed on the adjoining Cape Flats WWTW.
- *Caron credits:* In both cases a saving from the sale of carbon credits could be realised. As shown in Section 5.3.6 the bulk of the saving is from the diversion of sludge from landfill. In the case of THP digestion more sludge is treated and therefore it can be expected more carbon credits will be created for sale. Further, as the THP digestion case generates more methane and subsequently more power, this further increases the carbon credits sourced from the production of renewable energy. Emissions from the combustion of methane in CHP were small in each case, resulting in a net production of carbon credits for sale. For THP digestion the net emissions reduction is over 3 times higher than that of conventional digestion, showing a significant environmental benefit due from THP digestion.

5.4. CAPITAL COST AND PAYBACK

The capital cost estimated for each scenario is shown in Table 5-15. These were calculated using the methods and inputs described in Section3.4. The convetional digestion case requires R551 860 732 to maintain and reinstate where the THP digestion case requires R916 864 662 to implement. The difference between the two cases is R365 003 938 and represents the additional capital investment required to upgrade from conventional digestion to THP digestion. Considering the annual saving of R70 829 964 shown in Table 5-14 that THP digestion brings over conventional digestion the additional capital investment would take approximately 5 to 6 years to pay back. Thereafter the reduction in operating cost would continue to be realised as a further saving each year.

Table 5-15: Capital cost and payback

	Conventional digestion	**THP digestion**
Capacity (kgTSS/d)	60463	153855
Capital cost	R 551 860 723	R 916 864 662
Capital difference	R -	R 365 003 938
Saving generated (R/annum)	R-	-R 70 829 964
Estimated payback	-	5 to 6 years

6. CONCLUSIONS

This research aimed to provide local authorities with a useful tool for considering the applicability of THP technology as a means to improving their sludge management practices. It also aimed to provide insight into the impact of importing and digesting large quantities of NDBEPR WAS and the changes THP high solids AD brings in comparison to conventional AD. Finally, it sought to quantify major operating costs and sought to find if THP digestion can be economically and environmentally beneficial when retrofitted to an existing conventional digestion installation at Cape Flats WWTW.

6.1. THP AND MODELLING STEADY-STATE AD

Through a literature review, the fundamental changes that THP generates in sewage sludge were established (as discussed in Section 2.7). This insight was then able to be used to determine what impact these have on the steady-state modelling of the downstream AD in Section 4.3 and 4.5 . In summary, THP reduces viscosity to allow higher sludge loading rates without compromising digestion efficiency. THP also increases sludge biodegradability within the digester's SRT, thereby producing more gas. THP increases digestion rate and sludge biodegradability and improving the rate limiting hydrolysis step allows shorter sludge ages to achieve the same performance as conventional digestion. Modelling has shown that important parameters, such as free ammonia concentration, pH, alkalinity and methane production are within the correct range for stable digester operation, as presented in Sections 5.1.1, 5.1.5 and 5.1.7 respectively. THP provides improved VSR and sludge stabilisation over conventional digestion as shown in Section 5.1.6.

6.2. HIGH SOLIDS DIGESTION OF NDBEPR WAS

6.2.1. VSR

Digesting large amounts of NDBEPR WAS reduces VSR possible, even if a good organics removal is achieved. The reason for this is a reduction in the release of PP from NDBEPR WAS reduces the ISS content of the sludge, thereby increasing the digested sludge VSS/TSS ratio. However, organics conversion is still good as shown in Section 5.1.6.

6.2.2. Gas production

Further, digesting of significant amounts of NDBEPR WAS reduces specific methane per unit mass in the feed. The reason for this is if the AD is operating at a fixed VSS loading rate and feed %DS increasing the PP content of NDBEPR WAS fed to digestion increases TSS loading, which decreases digester SRT, and with it specific methane production per ton of VSS decreases. This would appear to be the case regardless if THP digestion or conventional digestion were used.

6.2.3. Digester effluent P-treatment required

The THP digestion allows for greater throughput. This allows greater quantities of sludge to be imported from some surrounding WWTW's. While this is beneficial to increasing sludge treatment capacity this brings with it some challenges. Conventional digestion of NDBEPR sludge with increasingly higher P-content directly increases the extent of P-treatment required downstream of the AD before the AD liquors return to the adjacent WWTW. This effect is exacerbated in the THP digestion case due to importing of large quantities of NDBEPR sludge, and therefore large quantities of nutrients. The digester effluent when dewatered will return large quantities of P to the adjacent WWTW, which being an NDBEPR process, can only remove P via WAS (and through AD the P is subsequently released back to the WWTW, thereby creating a loop of P which ultimately will deteriorate the Cape Flats WWTW treated effluent quality). Therefore, P treatment is required in the AD effluent, and even more so when increasing the of amount sludge digested in creating a regional sludge facility. Treatment of

OP in digester effluent can be done by MgO dosing to create struvite precipitation. This reduces OP by 85% and has the added benefit of improving the stability of the final sludge (which is now a mix of sludge+struvite) by increasing VSR due to struvite as ISS lowering the sludge's VSS/TSS ratio, as discussed in Section 5.3.4.

6.2.4. Digester effluent N-treatment required

The THP digester may operate at high solids and thus high FSA concentrations. The amount of WAS in the AD feed and the type of WAS fed must be carefully considered so as to not exceed FAN inhibition concentrations. It was found in this study that this would not be the case and FAN was within the limits for stable digestion, as discussed in Section 5.1.1.

The digestion of NDBEPR sludge may benefit downstream N-treatment. This is due to increased struvite precipitation potential inside the digester when PP is broken down to release magnesium, which combines with both N and P to form struvite. Downstream of the digestion process, P-treatment is required, and when selecting a P-treatment technology such as struvite precipitation from MgO dosing the TKN load will reduce as NH_4^+ is consumed in struvite formation. The TKN load requiring downstream treatment will therefore be less. This is only possible because high amounts of OP are present in the digester effluent from digesting high PP content NDBEPR WAS. The extent of TKN removal in the P-treatment step will be less when digesting low P content WAS e.g. WAS from an MLE system, and therefore increase the TKN load requiring treatment. Therefore, digesting NDBEPR WAS (with the inclusion of struvite precipitation) can reduce on required TKN treatment from the digester effluent, but the cost of P-treatment will most likely rise.

6.2.5. Economics of nutrient treatment not prohibitive

While both THP digestion and conventional digestion cases may require the inclusion of treatment for dewatering liquor, before it is returned to the adjacent WWTW, the extent of nutrient treatment required is significantly more in the case of THP digestion, as shown in Section 5.1.1, 5.1.2 and 5.3.4. A measurement of the impact of P and N treatment was evaluated in this study by applying high level basic operational costing of popular treatment technologies, using struvite precipitation with MgO dosing for P-treatment and anammox for N-treatment. A brief literature review was done on these technologies in Section 3.3.6 and followed with the evaluative modelling of their impact on operating cost. The findings in Section 5.3.3 showed the power required for N-treatment increased by 152% from conventional digestion to THP digestion, and Section 5.3.4 shows that the chemical dosing required for P-treatment increase by 287%. However, the overall impact on the total operating costs (given in 5.3.7) shows THP digestion still brings significant net savings over conventional digestion, and therefore suggested that these additional requirements for nutrient removal will not negatively jeopardise the economic benefit of implementing THP digestion over conventional digestion. Any drawbacks from additional nutrient treatment required in the case of THP digestion are overshadowed by other benefits THP brings (reduced sludge disposal, increased power recovery, generation of carbon credits).

6.3. INCREASED THROUGHPUT

When retrofitting THP to an existing digestion installation the sludge throughput can be increased by over 2.5 times while maintaining optimum digester operation, as discussed in Section 2.7 and 5.1.6. This would allow sludge from surrounding WWTW's to be imported and thereby enabling Cape Flats to be a regional sludge processing centre. THP digestion would allow the imported sludge to be stabilised and its volume reduced, which is a major benefit for the imported sludge that otherwise would have gone to landfill, as there are no other operational digesters in the city. As shown in Section 5.3.4 the THP digestion case and increased imports result in approximately half the volume of sludge requiring disposal that is required in the conventional digestion case. A further benefit is the final sludge from THP digestion has increased disposal options than that from conventional digestion, due to improved stabilisation and pathogen elimination resulting in less restrictions for beneficial sludge use.

6.4. DRIVERS FOR THP

This comparison allows one to appreciate some of the major costs and savings in an AD plant's operational budget. For example, it is clear from Figure 5-17 that a reduction in sludge disposal costs is a major driver. Further, including CHP in an AD facility brings with it a significant cost saving due to energy recovery and thus should be included as part of such installations. With regards to major costs, nutrient treatment and polyelectrolyte costs cannot be ignored as these bring major operational expenses in THP digestion. Finally, generation of carbon credits has the possibility to create significant revenue streams during sludge treatment via AD.

The following drivers support the case in investment in THP over conventional digestion.

- The need to reduce operational cost. This would come in the form of savings from diverting sludge from landfill and generating increased amounts of power on site. This is shown in Section 5.3.3.
- The generation of power on site also makes the WWTW more resilient by increasing independence from external utility power.
- The diversion of sludge from landfill and generation of renewable energy from biogas reduces GHG emissions as shown in section 5.3.6.
- The sale of carbon credits from emissions reductions brings a potential saving in the form of revenue.
- Another driver would be complying with environmental legislation by diverting sludge away from landfill as well as saving landfill space, especially as landfills reach the end of their usable life.
- Another driver for installing THP is needing to increase digestion throughput without building new digesters. This might be due to increased sludge production when expanding the capacity of a WWTW, or the need to create a regional facility and import sludge from other WWTW's, all while having space constraints.
- Benefits would still be achieved using THP in scenarios other than retrofitting to increase capacity e.g. building a greenfields site with smaller digester volume. In addition to smaller digester volume many of the other THP benefits would apply in this scenario, such as making more biogas per ton VSS and thus more renewable energy and allowing for less restrictive final sludge usage options.

With these drivers in mind some challenges also exist. The retrofit of THP requires capital investment and once built increases plant complexity. Further, the increased plant throughput requires increased P and N treatment of AD effluent, however, these challenges are not prohibitive to the THP digestion case.

6.5. ECONOMIC AND ENVIRONMENTAL BENEFITS

The findings of this research indicate that it is more economically and environmentally sustainable to retrofit thermal hydrolysis process (THP) to an existing sewage sludge anaerobic digestion facility rather than simply maintaining conventional anaerobic digestion. The retrofit of THP to the existing AD at Cape Flats increases capacity and brings with it an environmental benefit over conventional digestion by reducing landfill disposal volumes and lowering net carbon emissions, as shown in Sections 5.3.4 and 5.3.6 respectively. Economically, due to net lower operational costs as shown in Section 5.3.7, the investment in additional capital to retrofit THP at Cape Flats WWTW has shown to result in a 5 to 6 year pay-back, after which these savings will continue to be realised, thus proving to show economic benefit over conventional digestion. All hypotheses in this research have thus been proven true.

7. RECOMMENDATIONS FOR FURTHER RESEARCH

The following further investigation is suggested:

1. Modifying the model to include the digestion of particulate and soluble organics separately, using their own kinetic parameters. This would be especially relevant in the case of THP digestion, considering the degree of solubilisation.

2. Future work can be done to establish the change in biodegradability THP creates for the feed organics. This would be done to investigate to what extent slowly biodegradable organics and unbiodegradable organics are changed to readily biodegradable organics. Tests can be carried out where raw sludge and THP treated sludge are separately batch digested at long sludges ages >100day SRT to determine if the same amount of biogas is ultimately produced. This will then determine if THP has in fact changed the biodegradability of substances that were previously unbiodegradable, within a digester residence time.

3. Establish extent of inlet and effluent VFA concentration dissociated acetate species, and link this to digester pH.

4. Perform the same comparison study, but instead of using the same digester volume for each case rather the same throughput is used. This will be applicable for the scenario where new digesters are to be constructed to process the same amount of sludge. In this way, the volume required for THP digestion is expected to be significantly lower than that of conventional digestion and the economic impacts could be investigated.

5. Perform laboratory scale or pilot scale tests thermally pre-treating sludge followed by anaerobic digestion, with the focus to generate the data required as inputs to the steady-state model. Kinetic data for the digestion of both the soluble and particulate fractions of the hydrolysed sludge will be particularly useful. Detailed analysis of the influent and effluent through the THP unit can then allow modelling of the THP unit process operation in isolation and predict the change in characterisation of a sludge from its raw state in the THP influent to its THP treated state in the THP effluent (which will be fed as input to the AD). This will allow for full characterisation of the sludge fed to AD and improve modelling of the AD of THP pre-treated sludge.

6. Calibrate the kinetic inputs of the model using plant specific data e.g. from the Cape Flats WWTW anaerobic digesters operating at full loading rate. This would then impact the methane production, biomass growth and COD removal.

7. Investigate to effects of THP digestion on sulphate reduction and the corresponding formation of hydrogen sulphide (H_2S). According to log species diagrams shown in Poinapen *et al* (2009), and operating at a higher pH as observed in THP digestion, may reduce the production of hydrogen sulphide and improve biogas quality (i.e. less H_2S treatment required for biogas usage in CHP engines).

8. Investigate struvite recovery from THP sludge prior to AD. Han *et al* (2017) observed most of the OP created through the THP+MAD of NDBEPR WAS occurred during the THP unit step. Further, significant FSA is also released during THP, as shown in Table 2-3 of the literature review, providing Both N and P required for struvite formation.

9. Quantify other potentially significant operating savings and costs. Significant savings include the revenue that could be generated from the sale of sludge, especially in the THP digestion case where improved sludge quality no longer requires restricted disposal. Other operating expenses include the cost of maintenance and operator expenses.
10. The assessment of the anammox should be further investigated. This should include a capacity analysis of the plant, assessment of influent biodegradability and the required quality of the final treated side stream.
11. Further evaluate the effect of the THP pulper/reactor high pressure may have on inorganic carbon and how this influences the buffer capacity of the system.
12. Balance operational pH and Mg dosage (and type of compound) when precipitating struvite. This would require detailed water chemistry modelling of the phosphorous treatment struvite precipitation process (downstream of the AD).

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9. APPENDICES

In the below calculation sections please note:

	Yellow cells are input values
	Blue cells are calculated values

9.1. SLUDGE CHARACTERISTICS – INFLUENT AND EFFLUENT

This section shows the sludge characterisation for each sludge type fed to AD.

9.1.1. Raw sludge fed to conventional digestion

Table 9-1: Characteristics of PS and WAS fed to conventional digestion

PS raw						
	BSO	USO	BPO	UPO	Inorganic	Total
COD - mgCOD/l	165	50	39306	17189	0	56709
TKN - mgTKN/l	4	1	867	693	40	1605
TOC - mg TOC/l	57	17	12386	5941	5	18406
TP - mgP/l	1	0	320	614	7.3	942
VSS - mg VSS/l	0	0	26569	11387	0	37955
ISS - mgISS/l	0	0	0	0	9044.96	9045
TSS - mgTSS/l	0	0	26569	11387	9044.96	47000
WAS raw						
	BSO	USO	BPO	UPO	Inorganic	Total
COD - mgCOD/l	0	45	23200	26786	0	50032
TKN - mgTKN/l	0	1	2229	668	40	2938
TOC - mg TOC/l	0	15	8306	9544	5	17870
TP - mgP/l	0	0	644	247	7.3	898
VSS - mg VSS/l	0	0	16168	18459	0	34628
ISS - mgISS/l	0	0	0	0	12372.2	12372
TSS - mgTSS/l	0	0	16168	18459	12372.2	47000
Mixed PS and WAS to conventional AD						
	BSO	USO	BPO	UPO	Inorganic	Total
COD - mgCOD/l	66	47	29643	22947	0	52703
TKN - mgTKN/l	2	1	1684	678	40	2405
TOC - mg TOC/l	23	16	9938	8103	5	18084
TP - mgP/l	0	0	514	393	7.3	915
VSS - mg VSS/l	0	0	20328	15630	0	35959
ISS - mgISS/l	0	0	0	0	11041.3	11041
TSS - mgTSS/l	0	0	20328	15630	11041.3	47000

9.1.2. Sludge fed to THP digestion (with mass balance over THP)

The sludge characteristics in Table 9-2 are for PS and WAS before and after THP.

Table 9-3 following is simply a flux-weighted mixture of the two sludges. Note: these tables do not include the processing and release of OP (from PolyP) which is modelled in another part of the model. FSA release due to THP has not been modelled in this study.

Table 9-2: Characteristics of each sludge fed to THP and post-THP fed to AD

PS raw to THP								**THP PS**								
	BSO	USO	BPO	UPO	Inorganic	Total		BSO	USO	BPO	UPO	Inorganic	Total		*Mass balances*	
COD - mgCOD/l	1550	30	91992	40229	0	133801		1550	796	91226	40229	0	133801		*100%*	*COD*
TKN - mgTKN/l	36	1	2029	1622	40	3728		36	18	2012	1622	40	3728		*100%*	*TKN*
TOC - mg TOC/l	534	10	28989	13904	10	43447		534	251	28748	13904	10	43447		*100%*	*TOC*
TP - mgP/l	9	0	749	1437	7.3	2202		9	6	743	1437	7.3	2202		*100%*	*TP*
VSS - mg VSS/l	0	0	62182	26649	0	88831		0	0	61664	26649	0	88313			
ISS - mgISS/l	0	0	0	0	21169	21169		0	0	0	0	21169	21169			
TSS - mgTSS/l	0	0	62182	26649	21169	110000		0	0	61664	26649	21169	109482			
WAS raw to THP								**THP WAS**								
	BSO	USO	BPO	UPO	Inorganic	Total		BSO	USO	BPO	UPO	Inorganic	Total		*Mass balances*	
COD - mgCOD/l	1350	30	54299	62692	0	118371		10440	796	53533	53601	0	118371		*100%*	*COD*
TKN - mgTKN/l	8	1	5216	1564	40	6829		235	74	5142	1337	40	6829		*100%*	*TKN*
TOC - mg TOC/l	495	10	19439	22337	0	42281		3734	284	19165	19098	0	42281		*100%*	*TOC*
TP - mgP/l	2	0	1507	577	7.3	2093		86	21	1485	493	7.3	2093		*100%*	*TP*
VSS - mg VSS/l	0	0	37841	43203	0	81044		0	0	37307	36939	0	81044			
ISS - mgISS/l	0	0	0	0	28956	28956		0	0	0	0	28956	28956			
TSS - mgTSS/l	0	0	37841	43203	28956	110000		0	0	37307	36939	28956	110000			

Table 9-3: Characteristics of the mixed sludge fed to THP and post-THP fed to AD

	Mixed PS and WAS to THP							**THP PS+WAS mixed**								
	BSO	USO	BPO	UPO	Inorganic	Total		BSO	USO	BPO	UPO	Inorganic	Total		*Mass balances*	
COD - mgCOD/l	1430	30	69376	53707	0	124543		6884	796	68610	48252	0	124543		*100%*	*COD*
TKN - mgTKN/l	19	1	3941	1587	40	5589		156	52	3890	1451	40	5589		*100%*	*TKN*
TOC - mg TOC/l	511	10	23259	18964	4	42747		2454	271	22998	17020	4	42747		*100%*	*TOC*
TP - mgP/l	5	0	1203	921	7.3	2136		55	15	1188	871	7.3	2136		*100%*	*TP*
VSS - mg VSS/l	0	0	47577	36581	0	84159		0	0	47050	32823	0	79873			
ISS - mgISS/l	0	0	0	0	25841	25841		0	0	0	0	25841	25841			
TSS - mgTSS/l	0	0	47577	36581	25841	110000		0	0	47050	32823	25841	105714			

9.1.3. Characteristics of final sludge for disposal

The sludge characteristics shown in Table 9-4 and Table 9-5 are for the sludge quantities for disposal listed in Table 5-12. For information, the sludge characteristics are shown with and without struvite P-treatment. This shows the range of struvite content likely in the sludge.

Table 9-4: Characteristics of sludge for disposal without undergoing struvite precipitation P-treatment

	Conventional digestion							THP digestion						
	COD	**TOC**	**TKN**	**TP**	**VSS**	**ISS**	**TSS**	**COD**	**TOC**	**TKN**	**TP**	**VSS**	**ISS**	**TSS**
Organics														
Filtered organic (mg/l)	47	16	1	0	0	0	0	796	271	52	15	0	0	0
Particulate organic (mg/l)	217526	76777	9644	4050	149015	0	149015	281250	99084	11235	5077	192439	0	192439
Total organic (mg/l)	217573	76792	9645	4050	149015	0	149015	282046	99355	11287	5092	192439	0	192439
Inorganics														
Filtered inorganic (mg/l)	0	663	914	624	0	0	0	0	1751	2793	1646	0	0	0
Struvite ISS (mg/l)	0	0	354	784	0	6195	6195	0	0	585	1295	0	10237	10237
Other ISS (mg/l)	0	0	0	0	0	64791	64791	0	0	0	0	0	97324	97324
Total (mg/l)	217573	77455	10913	5458	149015	70985	220000	282046	101106	14665	8034	192439	107561	300000
VSS/TSS					0.68							0.64		

Table 9-5: Characteristics of sludge for disposal including struvite precipitation after P-treatment

	Conventional digestion							THP digestion						
	COD	**TOC**	**TKN**	**TP**	**VSS**	**ISS**	**TSS**	**COD**	**TOC**	**TKN**	**TP**	**VSS**	**ISS**	**TSS**
Organics														
Filtered organic (mg/l)	47	16	1	0	0	0	0	796	271	52	15	0	0	0
Particulate organic (mg/l)	188170	66415	8342	3504	128904	0	128904	235767	83061	9418	4256	161318	0	161318
Total organic (mg/l)	188217	66431	8344	3504	128904	0	128904	236563	83332	9470	4271	161318	0	161318
Inorganics														
Filtered inorganic (mg/l)	0	663	674	94	0	0	0	0	1751	2161	247	0	0	0
Struvite ISS (mg/l)	0	0	1961	4343	0	34322	34322	0	0	3307	7323	0	57875	57875
Other ISS (mg/l)	0	0	0	0	0	56773	56773	0	0	0	0	0	80806	80806
Total (mg/l)	188217	67094	9018	3597	128904	91096	220000	236563	85082	11631	4518	161318	138682	300000
VSS/TSS					0.59							0.54		

9.2. UNBIODEGRADABLE SOLUBLE COD CREATED FROM THP

Unbiodegradable COD increase			
Specific USO production		11	kgUCOD/tDS
Digester effluent TSS flux		97407	kgTSS/d
Digester effluent TSS concentration		69642	mgTSS/l
USO flux generated		1071	kgCOD/d
USO concentration generated		766	mgCOD/l
WWTW flow		130	Ml/d
Increase in WWTW effluent COD		8	mgCOD/l

9.3. FEED PARAMETERS

		Conventional digestion				THP digestion			
Feed sludge characterisitics		PS	WAS	Combined		PS	WAS	Combined	
Make-up % of feed DS		40%	60%	100%		40%	60%	100%	
% active PAO's in WAS	fBG,WAS		32.6%				32.65%		
Poly in PAO's	q		0.23				0.230000		mol PP/molPAObiomass
fp,VSS of active PAO's (orgP+PP)			10.30%				10.31%		gP/gPAO,VSS
P content of WAS			6.05%				6.05%		gP/gVSS,WAS
UPO,VSS/VSS,total		0.300	0.533	0.4398		0.300	0.533	0.4398	
Unbiodegradable particulate organi	fS'up	0.30	0.54	0.435		0.30	0.45	0.387	THP could possibly lower these values
Unbiodegradable soluble organics	fS'us	0.0009	0.0009	0.0009		0.00595	0.00673	0.006391859	THP should increase these values more
Boiodegradability	fS'b	0.696	0.464	0.564		0.693	0.540	0.606171882	
X_{BPO}/X_T		0.565	0.344	0.43		0.563	0.339	0.43	
X_{BPO}/X_V		0.7000	0.47	0.56		0.698	0.460	0.56	
VSS/TSS calc PP change			0.737						
Difference			**0.00%**						
VSSin/TSSin		0.81	0.737	0.77		0.81	0.737	0.77	
Feed %dry solids TSS		4.70%	4.70%	4.70%	proportional to sludge	11.00%	11.00%	11.00%	proportional to sludge age
TSS flux to digestion		24185	36278	60463	kgTSS/d	61542	92313	153855	kgTSS/d
VSS flux to digestion		19531	26728	46259	kgVSS/d	49698	68013	117711	kgVSS/d
Volumetric flow to digestion		515	772	1286	m3/d	559	839	1399	m3/d
Total carbon to nitrogen ratio	C/N			7.5				7.6	
BO carbon to nitrogen ratio	C/N			5.9				6.3	

9.4. DIGESTER OPERATING PARAMETERS

Digester details		Conventional	THP	AD (after THP)	
Sludge age	Rs	14.65	13.48		days
Height	h	20.00	20.00		m
Diamter	d	20.00	20.00		m
Volume per digester		6283	6283		m
No. of identical digesters		3	3		
Digester volume	VD	18850	18850		m3
Inlet flowrate to digester		1.29	1.40		Ml/d
Inlet flowrate to digester		1286	1399		m3/d
Loading rate, TSS		3.21	8.16		kgTSS/m3/d
Loading rate, VSS		2.45	6.24	5.93	kgVSS/m3/d
TSS feed rate		60463	153855		kgTSS/d

9.5. MODEL INPUTS FOR SATURATION KINETICS

Model Inputs - Saturation kinetics		Conventional		THP	THP
		PS	WAS	PS	WAS
Temperature in digester		37	37	37	37
Increase in Km		-	-	0%	100%
maximum specific hydrolysis rate	Km	5.270	1.951	5.270	3.902
half saturation cocentration	Ks	7.98	9.109	7.98	9.109
Acidogen biomass Yield	YAD	0.113	0.113	0.113	0.113
endogenous respiration rate	bAD	0.041	0.041	0.041	0.041
biomass unbiodeg fraction (Yad)	fAD	0	0	0	0

9.6. COD KINETIC CONVERSION CALCULATIONS

Effluent concentrations - Saturation kinetics		Conventional			THP				
		PS	WAS	Combined	PS	WAS	Combined		
Biodegradable organics									
Particulate biodegrable COD	Sbpe	4.608	9.988	7.836	11.759	12.561	12.240	gCOD/l	Sbpi /[1 + (YAD*Km-(1/R+bAD))*(1+bAD*R*(1-YAD)) / (YAD*Ks*(1/R+bAD))]
Particulate biodegrable VSS	VSS,bp,e	3.116	6.961	5.423	7.967	8.792	8.462	gVSS/l	
Particulate biodegrable C	C,bp,e	1.453	3.576	2.727	3.711	4.496	4.182	mgC/l	
Particulate biodegrable N	N,bp,e	0.102	0.960	0.616	0.260	1.056	0.737	mgN/l	
Particulate biodegrable P	P,bp,e	0.037	0.277	0.181	0.095	0.308	0.223	mgP/l	
AD biomass									
Acidogen biomass	ZAD	2.570	1.002	1.629	6.144	3.899	4.797	gCOD/l	ZAD = YAD (Sbpi - Sbpe) / [1+bAD Rs{1-YAD(1-fAD)}]
Acidogen endogneous residue	ZED	0.000	0.000	0.000	0.000	0.000	0.000	gCOD/l	ZED = fAD bAD Rs ZAD
Hydrolysis rate									
Hydrolysis rate	rh	2.485	1.021		6.264	3.975			rh=Km*(Sbp/ZAD)/ [Ks+Sbp/ZAD]
Methane									
COD of methane flow	Sm	32.293	12.210	20.243	74.873	47.514	58.457	gCOD/l	Sm = (1-YAD) rh Rs
Methane flow		12.160	4.598		28.194	17.892		Nm3/l	Sm/64x24.1
Methane flow		15644	6431		39435	25025		Nm3/d	Sm(24.1/64)Qi
Effluent concentraiton	Ste	24.416	37.822	32.460	58.928	70.857	66.085	gCOD/l	Supi + ZAD + ZED + Sbpe +Susi
Conversion COD									
Fraction COD converted to sludge	E	0.07372	0.07583		0.07583	0.07583			E = (ZAD+ZED)/(Sbpi-Sbpe)
Fraction COD converted to sludge	E	0.07372	0.07583		0.07583	0.07583			YAD / (1+bAD*Rs*(1-YAD))
Fraction COD converted to methane		0.92628	0.92417		0.92417	0.92417			Sm / (Sbi-Sbpe)
Conversion COD BPO		88.3%	56.9%	74%	87.3%	80.4%	84%		
Conversion COD total		0.56945	0.24405		0.55958	0.40140			(Sti-Ste)/Sti
Total COD out		56.709	50.032	52.703	133.801	118.371	124.543	gCOD/l	Ste+Sm
total COD in		56.709	50.032	52.703	133.801	118.371	124.543	gCOD/l	Sti
COD balance		100%	100%	100%	100%	100%	100%		

9.7. ELEMENTAL CALCULATION FOR BIODEGRADABLE ORGANICS AND AD BIOMASS

Biomass digestion stoichiometry

Sludge BO (influent)	Conventional digestion		THP digestion		
	PS	WAS	PS	WAS	
fcv	1.47851	1.43493	1.47591	1.42866	gCOD/gVSS
fc	0.46610	0.51370	0.46582	0.51138	gC/gVSS
fn	0.03262	0.13784	0.03259	0.12009	gN/gVSS
fp	0.01202	0.03981	0.01195	0.03508	gP/gVSS
y	7	7	7	7	
z	2.080	1.738	2.087	1.911	y(1- 1/8 fcv - 8/12 fc - 17/14 f
x	3.198	4.828	3.202	4.802	fc/12 (y+16z)/(1 - fc - fn - fp)
a	0.192	1.110	0.192	0.967	fn/14 (y+16z)/(1 - fc - fn - fp)
b	0.032	0.145	0.032	0.128	fp/31 (y+16z)/(1 - fc - fn - fp)
f	0.579	0.579	0.788	0.788	
c	0.000	0.000	0.000	0.000	
ch	0.000	0.000	0.000	0.000	
Molar mass	82	113	82	113	g/mol
Electron donating capacity of sludge BO, γs	15.215	20.228	15.217	20.125	e-/mol of BPO
Molar COD of influent BPO	121.720	161.821	121.733	161.000	gCOD/mol BPO
Moles of BPO converted	0.286419673	0.081646	0.66553	0.3193331	mol/l

Biomass	Conventional di	THP digestion		
Biomass				
fcv	1.4159	1.4159	gCOD/gVSS	
fc	0.5310	0.5310	gC/gVSS	
fn	0.1239	0.1239	gN/gVSS	
fp	0.0250	0.0250	gP/gVSS	
l	7	7		hydrogen
m	2.083	2.083		oxygen
k	5.574	5.574		carbon
n	1.115	1.115		nitrogen
p	0.102	0.102		posphorous
				p split
s	0.000	0.000		sulphur
ch	0.000	0.000		charge
Molar mass	126	126	g/mol	
Electron donating capacity of biomass BO, γb	22.294	22.294	e-/mol of biomass	
COD of biomass	178.353	178.353	gCOD/mol biomass	gamma*8

9.8. STOICHIOMETRY OUTPUTS FOR AD PRODUCTS, POLYP RELEASE AND STRUVITE PRECIPITATION (IN THE AD)

Conventional Digestion					
	PS	WAS	PP	Struvite	Combined
	mol/l of PS	mol/l of WAS	mol/l	mol/l	mol/l
Reactants					
CxHyOzNaPb	0.286	0.082			0.164
H2O	0.155	0.327			0.258
$Mg_cK_dCa_ePO3$			0.014		0.014
Products					
$SO4^{2-}$	0.000	0.000			0.000
CO2	0.303	0.104	-0.010	0.006	0.179
CH4	0.505	0.191	0.002		0.318
CkHlOmNnPpSs	0.014	0.006			0.009
Nh4+	0.039	0.084		-0.004	0.062
H2PO4-	0.004	0.007	0.008	-0.002	0.012
$HPO4^{2-}$	0.003	0.005	0.006	-0.002	0.008
HCO3-	0.028	0.068	0.009	-0.006	0.055
MgNH4PO4				0.004	0.004
Mg2+			0.004	-0.004	0.000
K+			0.005		0.005
Ca2+			0.000		0.000
Ca:Mg ratio			1.10		12

THP Digestion					
	PS	WAS	PP	Struvite	Combined
	mol/l	mol/l	mol/l	mol/l	mol/l
Reactants					
CxHyOzNaPb	0.666	0.319			0.458
H2O	0.360	1.166			0.843
$Mg_cK_dCa_ePO3$			0.032		0.032
Products					
$SO4^{2-}$	0.000	0.000			0.000
CO2	0.701	0.432	0.007	0.017	0.564
CH4	1.170	0.742	0.0000		0.913
CkHlOmNnPpSs	0.034	0.022			0.027
Nh4+	0.089	0.284		-0.010	0.197
H2PO4-	0.014	0.030	0.026	-0.008	0.042
$HPO4^{2-}$	0.004	0.008	0.007	-0.002	0.011
HCO3-	0.068	0.238	-0.007	-0.017	0.146
MgNH4PO4				0.010	0.010
Mg2+			0.010	-0.010	0.000
K+			0.011		0.011
Ca2+			0.001		0.001
			1.10		526

9.9. MASS BALANCES TO ENSURE CONSISTENCY BETWEEN KINETICS AND STOICHIOMETRY

COD balances - Kinetics vs stoichiometry						
	Conventional digestion			**THP digestion**		
	PS	**WAS**		**PS**	**WAS**	
BPO hydrolyzed kinetics	34.863	13.212		81.017	51.413	gCDO/l
BPO hydrolyzed stoichiometry	34.863	13.212		81.017	51.413	gCDO/l
Difference	100%	100%		100%	100%	gCDO/l
CH4 produced kinetics	32.293	12.210		74.873	47.514	gCDO/l
CH4 produced stoichiometry	32.293	12.210		74.873	47.514	gCDO/l
Difference	100%	100%		100%	100%	gCDO/l
Biomass grown kinetics	2.570	1.002		6.144	3.899	gCDO/l
Biomass grown stoichiometry	2.570	1.002		6.144	3.899	gCDO/l
Difference	100%	100%		100%	100%	gCDO/l

9.10. PP CONTENT OF PAO'S AND DISTRIBUTION IN WAS VSS

Polyphosphate release from WAS					
		Conventional digestion	**THP digestion**		
Organically bound P in Pao's	fP	4.0%	4.0%	gP/gPAO,VSS	
Polyphosphate P in PAO's	fXBGPP	6.3%	6.3%	gP/gPAO,VSS	
(fP+fXBGPP)	fXBGP	10.3%	10.3%	gP/gPAO,VSS	fP+fXBGPP
P content of WAS		6.0%	6.0%	gP/gVSS,WAS	
P content of WAS		4.5%	4.5%	gP/gTSS,WAS	
PAO biomass (excl. PP)					
% active PAO's in WAS	fBG,WAS	33%	33%		
WAS flow		26728	68013	kgVSS/d	
Active PAO's in WAS	MXBG	8726	22204	kgVSS/d	
PAO biomass concentration	XBG	6783	15875	mgVSS/l	MWGB/Qtotal
PAO biomass concentration		0.060	0.141	mol/l	n=m/M
Polyphosphate					
P in PolyP		552	1405	kgP/d	
P in PolyP		429	1004	mgP/l	
PolyP concentration		0.014	0.032	molPP/l	
Poly in PAO's	q	0.230000	0.230000	mol PP/molPAObiomass	

9.11. PP ELEMENTAL CHARACTERISATION

Polyphosphate mass fractions				
P in PP		1	1	molP/molPP
P in PP		0.014	0.032	molP/l
P in PP		429	1004	mgP/l
Mg in PP	c	0.30	0.30	molMg/molPP
Mg in PP		0.00	0.01	molMg/l
Mg in PP		100	233	mgMg/l
K in PP	d	0.33	0.33	molK/molPP
K in PP		0.00	0.01	molK/l
K in PP		178	417	mgK/l
Ca in PP	e	0.03	0.03	molCa/molPP
Ca in PP		0.00	0.00	molCa/l
Ca in PP		17	39	mgCa/l
O in PP		3.00	3.00	molO/molPP
O in PP		0.04	0.10	molO/l
O in PP		664	1555	mgO/l
mass PP		1387	3249	mgPP/l
Molar mass PP		100.27	100.27	g/mol
Fraction of PP released with PH	fqPP	0.80	0.00	
	YPP	0.65	0.65	mol P released / mol PHB formed

9.12. EFFLUENT ORGANICS MASS FRACTIONS

Effluent organics characterisation														
	Conventional	**THP**	**Conventional**	**THP**	**Conventional**	**THP**	**Conventional**	**THP**	**Conventional**	**THP**	**Conventional**	**THP**	**Conventional**	**THP**
	Unfiltered	Unfiltered	Soluble	Soluble	Particulate	Particulate	Particulate	Particulate	Particulate	Particulate	Particulate	Particulate	Particulate	Particulate
			USO	USO	UPO+ZAD+ZED+Sbpe	UPO+ZAD+ZED+Sbpe	UPO	UPO	Biomass, ZAD	Biomass, ZAD	ZED	ZED	BPO, Sbpe	BPO, Sbpe
gCOD/l	32.460	66.085	0.047	0.796	32.413	65.289	22.947	48.252	1.629	4.797	0.000	0.000	7.836	12.240
COD/VSS, fcv			1.493	1.454	1.460	1.462	1.468	1.470	1.416	1.416	1.416	1.416	1.445	1.446
gVSS/l (or mass mg/l)	22.236	45.220	0.031	0.547	22.204	44.673	15.630	32.823	1.151	3.388	0.000	0.000	5.423	8.462
OrgC/VSS, fc			0.498	0.495	0.515	0.515	0.518	0.519	0.531	0.531	0.531	0.531	0.503	0.494
OrgC/l	11.456	23.272	0.016	0.271	11.440	23.001	8.103	17.020	0.611	1.799	0.000	0.000	2.727	4.182
OrgN/VSS, fn			0.036	0.094	0.065	0.058	0.043	0.044	0.124	0.124	0.124	0.124	0.114	0.087
gOrgN/l	1.438	2.660	0.001	0.052	1.437	2.608	0.678	1.451	0.143	0.420	0.000	0.000	0.616	0.737
orgP/VSS, fp			0.000	0.028	0.420	0.452	0.025	0.027	0.025	0.025	0.025	0.025	0.033	0.026
gOrgP/l	0.604	1.194	0.000	0.015	0.604	1.179	0.393	0.871	0.029	0.085	0.000	0.000	0.181	0.223

9.13. BIOGAS COMPOSITION

Biogas production				
Molar volume of gas @ 20degC, 1atm	0.0240	0.0240	Nm3/mol	V/n = R*T/p
Methane				
Methane gas flow	9838	30714	Nm3/d	mol*24l/mol*Qps
Methane gas fraction	64%	62%	Nm3/d	
Methane produced/VSS fed	213	275	Nm3CH4/tonVSSfed	
Methane produced/BCOD fed	194	237	Nm3CH4/tonBCODfed	
Methane produced/COD fed	135	164	Nm3CH4/tonCODfed	
Lower heating value methane	8.026E+08	8.026E+08	J/kmol	Table 2-179, Perrys 8th Ed
Lower heating value methane	33.4	33.4	MJ/Nm3	(Molar value/p/R/T/1000)/1000000
LHV biogas	21.4	20.6	MJ/Nm3	
Carbon dioxide				
Partial pressure of CO2	0.3601	0.3816	atm	molCO2 / (molCO2+molCH4)
CO2 gas fraction	36%	38%		
CO2 gas flow	5536	18952	Nm3/d	
Carbon dioxide/VSS fed	120	170	Nm3CO2/tonVSSfed	
Total biogas				
Total gas flow / l influent	11.95	35.51	l gas / l sludge	
Total gas flow	15374	49666	Nm3/d	
Total gas flow	641	2069	Nm3/h	
Biogas produced/VSS fed	332	445	Nm3/tonVSfed	
Biogas produced/TSS fed	254	323	Nm3/tonDSfed	
Biogas produced/BCOD fed	303	383	Nm3/tonBCODfed	
Biogas produced/COD fed	211	265	Nm3/tonCODfed	

9.14. STRUVITE PRECIPITATION INSIDE AD

Struvite precipitation potential		Conventional digestion	THP digestion	Mutiple increase		
Concentration		0.923	2.376		g/l	
Molar mass		245.0000	245.0000		g/mol	MgNH4PO3.6H2O
Struvite precipitated	R	0.004	0.010		mol/l	solver to equate Iionic product and solubility product
P total soluble	PT	0.020	0.053		mol/l	
N total soluble	NT	0.065	0.200		mol/l	
Magnesium concentration	[Mg2+]	0.0004	0.0000	0.05	mol/l	
Magnesium concentration	[Mg2+]	0.382738	0.020331		mmol/l	
Ammonium concentration	[NH4+]	0.0647	0.1952	3.02	mol/l	
Free ammonia	[NH3]	0.0006	0.0043		mol/l	
Free ammonia	[NH3]	10	72		mg/l	above 600mg/l free ammonia is inhibtion
Phosphate concentration	[PO43-]	3.90E-07	4.05E-06	10.40	mol/l	
Struvite solublity coefficient @25degC	pKsp	12.6	12.6			
Solublity coefficient @T	pKsp	12.60	12.60			
Ionic product		9.6433E-12	1.6078E-11			
Solublity product corrected		9.6433E-12	1.6078E-11			
Struvite solubility coefficient corrected for ionic strength	pKsp'	11.0157721	10.793764		pKsp' = pKsp+logfm+logfd+logft	
Ioninc product	pKIP	11.0157721	10.793764		pKIP = - (log[NH4+]+log[PO4-3]+log[Mg2+])	
diff		0	0			
Precipitation or soluble? i.e. when IP > Ksp		**Soluble**	**Soluble**			
Mass struvite formed in AD		1187	3324		kg/d	

9.15. DIGESTER PH AND PHOSPHOROUS F-VALUE SPECIATION

Digester pH		**Conventional**	**THP**	
Bicarbonate concentration	[HC03-]	0.0548	0.1456	mol/l
Dissociation constant 1 for cabonate system	pK'c1	6.191144	6.175286	
Dissociation constant 2 for cabonate system	pK'c2	9.900003	9.852430	
Hnery's law constant for CO2	pK'HCO2	1.6073	1.6073	
pH		6.981	7.364	solver to equate pCO2's
Partial pressure CO2 - current	pCO2	0.360	0.382	atm
Partial pressure CO2 - calc form eqn 31	pCO2 calc.	0.360	0.382	
Diffreence pCO2		0.0000%	0.0000%	
Phosphorous speciation		**Conventional**	**THP**	
pH from P system		6.98	7.36	pH = pK'p2 - log([HPO42-] / [H2PO4-])
pH from C system		6.98	7.36	
difference pH		0.0000%	0.0000%	
f		0.58	0.79	solver to equate pH's

9.16. EQUILIBRIUM CONSTANTS AT TEMPERATURE AND SPECIATION CONSTANTS

Convetional digestion						pKT = A/T-B+C*T+D*logT		
		pK @25degC	A	B	C	pK @T	pK'@T	
Henry's constant	pKCO2	1.47	-1760	-9.619	-0.00753	1.6073		
Carbonate	pKc1	6.352	3404.7	14.8435	0.03279	6.3043	6.19114	pKc1+logfm
Carbonate	pKc2	10.33	2902.4	6.498	0.02379	10.2395	9.900	pKc2-logfm+logfd
Phosphate	pKp1	2.149	799.3	4.5535	0.01349	2.2068	2.0936	pKp1+logfm
Phosphate	pKp2	7.2	1979.5	5.3541	0.01984	7.1818	6.8423	pKp2-logfm+logfd
Phosphate	pKp3	12.023	0	0	0	12.0230	11.4572	pKp3-logfd+logft
Ammonium	pKn	9.246	2835.8	0.6322	0.00123	8.8968	9.0100	pKn-logfm

Speciation constants

P system			N system
Wp = 10^(pKp1'-pH)	Xp=10^(pH-pKp2')	Yp=10^(pH-pKp3')	Wn=10^(pKn'-pH)
Wp	Xp	Yp	Wn
0.000013	1.37685	0.000033	106.858

THP digestion						pKT = A/T-B+C*T+D*logT		
		pK @25degC	A	B	C	pK @T	pK'@T	
Henry's constant	pKCO2	1.47	-1760	-9.619	-0.00753	1.6073		
Carbonate	pKc1	6.352	3404.7	14.8435	0.03279	6.3043	6.17529	pKc1+logfm
Carbonate	pKc2	10.33	2902.4	6.498	0.02379	10.2395	9.852	pKc2-logfm+logfd
Phosphate	pKp1	2.149	799.3	4.5535	0.01349	2.2068	2.0778	pKp1+logfm
Phosphate	pKp2	7.2	1979.5	5.3541	0.01984	7.1818	6.7947	pKp2-logfm+logfd
Phosphate	pKp3	12.023	0	0	0	12.0230	11.3779	pKp3-logfd+logft
Ammonium	pKn	9.246	2835.8	0.6322	0.00123	8.8968	9.0259	pKn-logfm

Speciation constants

P system			N system
Wp = 10^(pKp1'-pH)	Xp=10^(pH-pKp2')	Yp=10^(pH-pKp3')	Wn=10^(pKn'-pH)
Wp	Xp	Yp	Wn
0.000005	3.70900	0.000097	45.906

9.17. IONIC STRENGTH CALCULATION

Ionic strength calculation

Conventional digestion

ST	0.0	mgHs2-S/l	0.0000	mol/l
AT	0.0	mgHAc/l	0.0000	mol/l
PT	623.9	mgP/l	0.0201	mol/l
NT	913.9	mgN/l	0.0653	mol/l
Mg	0.0	mgMg/l	0.0000	mol/l
CT	663.2	mgC/l	0.0553	mol/l
K+	0	mgK/l	0.0	mol/l
Ca2+	0	mgCa/l	0.0	mol/l
Ionic strength				
Ionic strength (μ)	0.1508	mol/L	μ = sigma(Ci*Zi^2)	
Total dissolved solids, TDS	6053	mg/l	rearranged μ = 2,5*10^-5*(TDS-20)	
Conductivity	898	mS/m	EC = μ/(1,68*10^-4)	
A = 1,825*10^6/(78,3*T)^1,5	0.4826			
log(fi) = -AZi^2x(mu^0.5/(1+mu^0.5)-0.3mu)	log(fi)		fi = 10^log(fi)	
monovalent	-0.11316	fm	0.77062	
divalent	-0.45264	fd	0.35267	
trivalent	-1.01843	ft	0.09584	

THP digestion

ST	0.0	mgHs2-S/l	0.0000	mol/l
AT	0.0	mgHAc/l	0.0000	mol/l
PT	1646	mgP/l	0.0531	mol/l
NT	2793	mgN/l	0.1995	mol/l
Mg	0.0	mgMg/l	0.0000	mol/l
CT	1751	mgC/l	0.1459	mol/l
K+	0	mgK/l	0.0000	mol/l
Ca2+	0	mgCa/l	0.0000	mol/l
Ionic strength				
Ionic strength (μ)	0.4117	mol/L	μ = sigma(Ci*Zi^2)	
Total dissolved solids, TDS	16486	mg/l	rearranged μ = 2,5*10^-5*(TDS-20)	
Conductivity	2450	mS/m	EC = μ/(1,68*10^-4)	
A = 1,825*10^6/(78,3*T)^1,5	0.4826			
log(fi) = -AZi^2x(mu^0.5/(1+mu^0.5)-0.3mu)	log(fi)		fi = 10^log(fi)	
monovalent	-0.12902	fm	0.74299	
divalent	-0.51607	fd	0.30474	
trivalent	-1.16115	ft	0.06900	

9.18. OPERATIONAL ENERGY AND HEATING ENERGY

Operational energy				
Specific usage factor			0.28	kWh/kgTSS
Plant throughput		60463	153855	kgTSS/d
Energy requirements		6179337	15723973	kWh/annum
Power consumption		705	1795	kW
HEnet (heating energy)				
Heating energy for conventional		**Conventiona**	**THP**	
Influent temperature		20	20	degC
AD temperature		37	37	degC
Boiler efficiency		80%	80%	
Heat to influent	HE,convent	1323	1439	kW
	P,HGH	950		kW
Hot water supply temperatre from CHP		85	85	
Hot water return temperature to CHP		55	55	
Hot water flow		38	41	m3/h
Steam consumption for THP		**Conventiona**	**THP**	
Temperatureof steam flash vessel			102	degC
Temperature of steam in THP reactor			165	degC
Internal temperature differnce	delT		63	degC
System efficiency	epsilon,steam		85%	
Dry solids entering THP	DS		0.17	
Steam	S		760	kg steam/tonne dry solids
Throughput	m,solids		6.41	tonDS/h
Steam required	S x m,solids		4874	kg steam/h
Heat capacity of steam 6barg	Hvap		2064	kJ/kg
Latent heat			2794	kW
Specific heat capacity water	Cp		4.182	kJ/kg
Inlet water temperature to boiler	Ti		20	
Steam temperature	T,THP		165	
Specific heat required			821	kW
Total energy	HE,THP		3615	kW
	P,HGH		2967	kW

9.19. ENERGY RECOVERY FROM METHANE PRODUCTION

MP (methane production energy)				
		Conventional	**THP**	
Electrical efficiency	ε,elec	0.4	0.4	kW recovered /kW in gas feed
Thermal efficiency to high grade heat (+/- 400^{o}C)	ε,HGH	0.25	0.25	kW recovered /kW in gas feed
Thermal efficiency low grade heat (+/-85^{o}C)	ε,LGH	0.25	0.25	kW recovered /kW in gas feed
Lower heating value methane	LHV	8.03E+08	8.03E+08	J/kmol
Lower heating value methane	LHV	33.4	33.4	Mj/Nm3
Methane flow	Qch4	9838	30714	Nm3/d
Energy in biogas		3801	11868	kW
Energy in biogas (per CH4+CO2 total vol)		5.93	5.73	kWh/Nm3
Energy recovered				
Electrical power	Pelec	1521	4747	kW
Electrical power consumed	MP,elec	13320033	41584165	kWh/annum
High grade heat	Phgh	950	2967	kW
Low grade heat	Plgh	950	2967	kW
	MP	3421	10681	kW
	MP,heat	1901	5934	kW

9.20. N-TREATMENT ENERGY CONSUMPTION

Nitrogen treatment (NE)				
effluent flowrate	Qe	1286	1399	m3/d
Residual biodegradable TKN	X,TKN,BO	759	1157	mgN/l
FSA concentration	X,FSA	674	2161	mgN/l
Total TKN fo treatment		1433	3318	mgN/l
Total TKN fo treatment	Qe	1844	4641	kgN/d
Power consumption anammox	P,AN	1.2	1.2	kWh/kgN
Running power		92	232	kW
Power consumed	NE	807618	2032819	kWh/d

9.21. OP TREATMENT VIA STRUVITE PRECIPITATION

Orthophophate treatment (Mg)		Conventional	THP	
effluent phosphate concentration	P	624	1646	mg/l
effluent flowrate	Qe	1286	1399	m3/d
Molar mass phosphorous	MMp	31	31	g/mol
Molar mass magnesium oxide	MM,Mgo	40	40	g/mol
Stochiometric dosing ratio	MR	1.1	1.1	molMg:molP
Purity of Mgo	epsilon,Mg(	96%	96%	kgMgo/kg,material
MgO dosing required	MG	1187	3404	kg/d
MgO dosing required		30	85	kmol/d
Mgo cost		11.57	11.57	R/kg
treatment cost		R 13 728	R 39 383	R/d
treatment cost		R 5 010 814	R 14 374 789	R/annum
OP				
OP removal		85%	85%	
OP concentration before MgO		624	1646	mgP/l
OP concentration after MgO		94	247	mgP/l
OP removed		682	1957	kgP/d
OP removed		22	63	kmolP/d
FSA				
FSA removed		22	63	kmolN/d
FSA removed		308	884	kgN/d
FSA removed		239	632	mgN/l
FSA before MgO		914	2793	mgN/l
Free ammonia		13	92	mgFAN/l
FSA after MgO		674	2161	mgN/l
Mg				
Mg before Mgo		0.0004	0.0000	mol/l
Mg before Mgo		0.00049	0.00003	kmolMg/d
Mg after MgO dosing (before struvite precipitates)		30	85	kmolMg/d
Mg taken from struvite		22	63	kmolMG/d
% Mg Used		74%	74%	
Mg after struvite		8	22	kmolMG/d
Mg after struvite		184	527	kgMg/d
Mg after struvite		143	377	mgMg/l
MgNH4PO4				
Struvite before MgO		0.004	0.010	mol/l
Struvite before MgO		923	2376	mg/l
Struvite before MgO		1187	3324	kg/d
Struvite before MgO		4.847	13.567	kmol/d
Struvite formed MgO dosing		22.007	63.133	kmol/d
Struvite formed MgO dosing		5392	15468	kg/d
Struvite after MgO dosing		26.8538	76.7001	kmol/d
Struvite after MgO dosing		0.0209	0.0548	mol/l
Struvite after MgO dosing		6579	18792	kg/d
Struvite after MgO dosing		5114	13435	mg/l

9.22. CARBON EMISSIONS AND CARBON CREDITS

Carbon credits				
Diverted from landfill		**Conventional**	**THP**	
Methane flow	Qch4	9838	30714	Nm3/d
Density methane	rho,ch4	0.67	0.67	kg/Nm3
Methane flow		6548	20441	kg/d
Global warming potential	GWP,ch4	23	23	kgCH4/kgCO2
Oxidation factor	OX	0.1	0.1	
CO2e per annum diverted from landfill		**49470**	**154441**	tonCO2e/annum
From renewable energy				
Electrical power	Pelec	1521	4747	kW
Power usage			113929	kWh/d
Power usage		13320033	41584165	kWh/annum
Emissions factor	EF	1.04	1.04	kgCO2e/kWh
CO2e per annum from CHP renewable energy		**13853**	**43248**	tonCO2e/annum
Ch4 methane combustion		409	1278	kmol/d
CO2 from methane combustion		409	1278	kmol/d
CO2 from methane combustion		18006	56212	kg/d
CO2 from methane combustion in CHP		**-6572**	**-20517**	tonCO2/annum
Net CO2e per annum from renewable energy		**7281**	**22730**	tonCO2e/annum
TOTAL net CO2e		**56750**	**177171**	tonCO2e/annum
Rate		R 120	R 120	R/tonCO2e
Discount		15%	15%	
Saving		R 5 788 545	R 18 071 413	

9.23. DEWATERING PERFORMANCE AND POLYELECTROLYTE USAGE

Polyelectrolyte cost (Dewatering)		**Conventional**	**THP pre-dewatering**	**THP final**	
%DS in dewatering cake	DS	**22%**	**17%**	**30%**	
Capture rate		99.50%	99.50%	99.50%	centrifuges can achieve typically if %DS fed is >2%
Digester effluent concentration	TSS,e	37895	47000	83077	mg/l
Digester effluent flowrate	Qe	1286	1286	1399	m3/d
Sludge disposal		49	60	116	tonDS/d
Sludge disposal	SP	222	356	387	ton/d
Specific polyelectrolyte usage	pw,i	8	3.5	15	kg poly / ton TSS
Polyelectrolyte consumed	PY	390	212	1743	kg poly/d
Specific cost		R 40.18	R 40.18	R 40.18	R/kg poly
Polyelectrolyte cost		R 15 670	R 8 503	R 70 033	R/d
Polyelectrolyte cost		R 5 719 698	R 3 103 572	R 25 561 921	R/annum

	Conventional				**THP feed**				**THP final**		
	Sludge	Cake	Filtrate		Sludge	Cake	Filtrate		Sludge	Cake	Filtrate
Flow	1286	220	1066	m3/d	1286	354	933		1399	385	1013
Solids flow	48751	48507	244	kg/d	60463	60161	302		116198	115617	581
TSS	37895	220000	229	mg/l	47000	170000	324		83077	300000	573
%DS	3.8%	22.0%	0.02%		4.7%	17.0%	0.0%		8.3%	30.0%	0.1%
VSS/TSS	0.59	0.59	0.59						0.54	0.54	0.54
BPO/VSS	0.30								0.27		
TKN in sludge											
TKN,BPO / VSS	0.03	0.03	0.03	gN,BO/gVSS					0.03	0.03	0.03
TKN,BPO concentration	759	4406	5	mgN,BO/l				mgN,BO/l	1157	4178	8
FSA (after MgO)	674	674	674	mgN-NH4/l				mgN-NH4/l	2161	2161	2161
TKN flux	1844	1120	724	kgN/d				kgN/d	4641	2443	2198
OP &BioOrgP in sludge											
TP,BPO / VSS	0.009	0.009	0.009	gP,BO/gVSS				gP,BO/gVSS	0.01	0.01	0.01
TP,BPO concentration	210	1219	1.3	mgP,BO/l				mgP,BO/l	308	1112	2
OP (after MgO)	94	94	94	mgP-NH4/l				mgP-NH4/l	247	247	247
BioOrgP flux	391	289	101	kgP/d				kgP/d	776	524	252

		Conventional digester effluent	Conventional P-treatment effluent	Conventional dewatering filtrate	THP digester effluent	THP P-treatment effluent	THP dewatering filtrate	
FSA		1176	868	719	3907	3023	2190	kgN/d
Biodegradable organic TKN		976	976	5	1618	1618	8	kgN/d
Total		2152	1844	724	5525	4641	2198	kgN/d

		Conventional digester effluent	Conventional P-treatment effluent	Conventional dewatering filtrate	THP digester effluent	THP P-treatment effluent	THP dewatering filtrate	
OP		803	120	100	2303	345	250	kgP/d
Biodegradable organic P		270	270	1.4	2733 → 431	431	2	kgP/d
Total		1073	391	101	2733	776	252	kgP/d

9.24. CAPITAL COST

	Conventional		THP		
CapacityA		60000		143000	kgDS/d
Cost A	R	539 000 000	R	861 000 000	
Ii		114.9		114.9	
It		117.1		117.1	
n		0.60		0.60	
CapacityB		60463		153855	kgDS/d
CostB	**R**	**551 860 723**	**R**	**916 864 662**	Rmil/(tonDS/d)
Specific cost A	R	9.13	R	5.96	Rmil/(tonDS/d)