RACE for 2030
RELIABLE
AFFORDABLE
CLEAN
ENERGY

B5 Research Project

Improving biogas production of sewage sludge

Final Report

Australian Government
Department of Industry, Science and Resources
AusIndustry
Cooperative Research Centres Program

RACE for Business

Research theme B5: Anaerobic digestion for electricity, transport and gas

ISBN: 978-1-922746-47-4

Industry Report

Improving Biogas Production of Sewage Sludge

October 2023

Citations
Kaparaju, P., Parthasarathy, R., Eshtiaghi, N., Al Waili, I., Ordeu, N., Rybachuk, M., Das, T., Haris, M. (2023). Improving Biogas Production of Sewage Sludge. Prepared for RACE for 2030 CRC.

Project team

Griffith University

- Prof P. Kaparaju
- N. Ordeu
- M. Rybachuk

RMIT University

- Prof . R. Parthasarathy
- Prof. N. Eshtiaghi
- I. Al Waili
- T. Das
- M. Haris

Project partners

Acknowledgements

We would like to thank the Industry Reference Group participants from the following organisations: City of Gold Coast, Hunter Water, SA Water, Water Corporation WA, Water Service Association Australia (WSAA)

Although the IRG members and partners have provided valuable inputs and feedback throughout the project, the findings and recommendation included in this report do not necessarily reflect the views of each individual member.

Acknowledgement of Country

The authors of this report would like to respectfully acknowledge the Traditional Owners of the ancestral lands throughout Australia and their connection to land, sea and community. We recognise their continuing connection to the land, waters and culture and pay our respects to them, their cultures and to their Elders past, present, and emerging.

What is RACE for 2030?

RACE for 2030 CRC is a 10-year cooperative research centre with AUD350 million of resources to fund research towards a reliable, affordable, and clean energy future. https://www.racefor2030.com.au

Disclaimer

TABLE OF CONTENTS

List of Tables

List of Figures

List of abbreviations

ACCU	Australian Carbon Credit Unit
AD	anaerobic digestion
BAU	business as usual
BioCNG	biomethane compressed natural gas
BioCO$_2$	carbon dioxide from biogas upgrading process
BioRNG	biomethane renewable natural gas
BMP	biochemical methane potential
CapEx	capital expenditure
CER	Clean Energy Regulator
CERT	Corporate Emissions Reduction Transparency
CHP	combined heat and power plant
COD	chemical oxygen demand
CSTR	continuously stirred tank reactor
DS	Digestate sludge i.e., concentrated solids from digestate from AD
EBITDA	earnings before interest, tax, depreciation and amortisation
EPCM	engineering, procurement and construction management
ERF	Emission Reduction Fund
FM	fresh matter
FOFO	food organics and garden organics
FTIR	Fourier Transform Infra-Red
GHG	greenhouse gas
GPV	gravity pressure vessel
HRT	hydraulic retention time
IRG	industry reference group
IRR	internal rate of return
k	flow consistency, Pa.s^n
K	potassium
MW	megawatt
n	flow index (-)
N	nitrogen
Nm^3	cubic metres, normalised to STP
NmL	millilitres, normalised to STP
NGER	National Greenhouse and Energy Reporting
NPV	net present value
O&M	operation and maintenance costs
OLR	organic loading rate
OpEx	operating expenditure
P	Phosphorus
PBP	payback period
PS	Primary Sludge
PSM	process simulation model

RCSTR	rigorous CSTR
rDON	recalcitrant dissolved organic nitrogen
ROI	return on investment
sCOD	soluble COD
SOR	stoichiometric oxygen requirement
STP	standard temperature and pressure (0°C, 1 atm)
tCOD	total COD
TDS	total dissolved solids
TH	thermal hydrolysis
THP	thermal hydrolysis pretreatment
TKN	total Kjeldahl nitrogen
TS	total solids
TSS	total suspended solids
WAO	wet air oxidation
WAS	waste activated sludge
WWTP	wastewater treatment plant
VFA	volatile fatty acids
VS	volatile solids
VSS	volatile suspended solids
τ_y	shear yield stress, Pa

Executive summary

Annually, Sydney Water generates 58 GWh_{el} of electricity and 61 GWh_{th} of heat in cogeneration plants from the biogas (~60% methane) produced from its anaerobic digestion (AD) facilities. The produced biogas can supply 40-60% of a tertiary treatment plant's electricity requirement. The residual organic rich solids (biosolids) and the digestate still contain residual energy (25-30%). Thus, Sydney Water needs to establish if the remaining energy can be economically recovered. This project aims to study two commercially available technologies, namely Thermal Hydrolysis (TH) and Wet Air Oxidation (WAO) that can be retrofitted to increase production of biogas at different stages of the anaerobic digestion process.

The project stages included:

- **Critical Literature Review** - a rapid review on the effect of TH and WAO treatments on methane production from waste activated sludge (WAS) and anaerobic digestate, respectively, was carried out.
    - The review showed that TH of WAS concentrate (at 175°C) can greatly enhance the biodegradability (30-47%) and biogas production (40-60%) and reduces the hydraulic residence time (HRT) of AD process by around 25%.
    - The rheological characteristics of TH pretreated WAS concentrate were also shown to improve, thus improving the dewaterability of TH pretreated WAS. On the other hand, WAO pretreatment was found to be very effective in sludge reduction and conversion of sludge into useful products such as acetic acid. However, both TH and WAO treatments are energy-intensive processes and thus a thorough economic and technical feasibility of the two treatments was essential.
- **Thermal Hydrolysis** – The aim is to assess the energy balance of TH of thickened WAS prior to AD in terms of methane production and net energy generation.
- **WAO** of solid digestate from the anaerobic digestate - The purpose was to hydrolyse the organic solids and making them more biodegradable for further AD.

Later, chemical composition and methane potential of WAS Concentrate with and without TH pretreatment along with WAO of digestate concentrate at two different oxygen loading levels were performed. TH was performed at 165°C and 10-22 bar pressure for 30 min while WAO was performed at 240°C with 10% and 20% oxygen loading for 30 min. WAS Concentrate (8.5% total solids (TS)) produced methane yields of 378.10 NmL/g volatile solids $(VS)_{added}$. TH of WAS concentrate improved solubilisation of organic matter and enhanced the methane yields by 10.1% (416.24 NmL/gVS_{added}). On the other hand, WAO of digestate concentrate at 200°C resulted in a reduction of both TS and VS content but inhibited methane production at both 20% and 10% oxygen loadings. However, WAO of the concentrated digestate 165°C and at 20% oxygen loading resulted in methane yields of 117.32 NmL/gVS_{added} (21 d) with a lag phase of 6 days. The decrease in methane yields after WAO at 20% oxygen loading was 62%. Both these results thus suggests that further optimisation of the treatment conditions for WAO in terms of solids loading, treatment temperature and oxygen concentration needs to be determined.

- **Techno-economic feasibility** - Two Aspen Plus models of AD, models A and B, were used in this project to evaluate the techno-economic feasibility of WAO and TH treatments on biogas production and energy requirements. Model A uses stoichiometric reactions and can provide information on the kinetics of hydrolysis and methanogenesis stages of AD. However, it has limitations in expressing other two stages of AD viz., acidogenesis and acetogenesis and the inhibitory effects of metabolites on the kinetics of methane production. On the other hand, model B is a novel model that considers all four AD stages and has been used by several researchers due to its comprehensive approach and reliability.

An overview of the financial analyses of TH of WAS and WAO of digestate for a plant processing 86,400 tpa (1 kg/s) are provided in the following section. This analysis covers four scenarios based on three process alternatives:

1. The use of TH to pretreat the concentrated WAS.
2. The use of partial (20% oxygen loading) and complete (100% oxygen loading) WAO pretreatment using Gravity Pressure Vessel (GPV) technology to treat the digestate from the AD of the WAS.
3. Recycling partially oxidised WAO sludge liquid as process water to the AD reactor.

Four different scenarios are evaluated:

- Scenario 1 - Business as usual (BAU) - covers AD treatment of a concentrated WAS, and the separation of digestate into solids (concentrate) and liquid (centrate) fractions. Digestate solids (biosolids) are transported (300 km) for final disposal while the liquid fraction is sent to a wastewater treatment plant (WWTP). The biogas produced is used for cogeneration in a combined heat and power (CHP) plant. Renewable electricity and heat were consumed behind-the-meter.
- Scenario 2 - Similar to Scenario 1, but concentrated WAS is subjected to TH before AD.
- Scenario 3 - Similar to Scenario 1, but biosolids from the AD process are subjected to complete WAO at 200°C and 100% oxygen loading in a GPV.
- Scenario 4 – Similar to Scenario 1, but biosolids are subjected to partial WAO (PWAO) treatment at 200°C and 20% oxygen loading in a GPV. Recovered sludge liquid is recycled from the partial WAO process to the inlet of the AD process.

Using the process simulation models developed in Aspen Plus, mass and energy balance simulations were constructed in Microsoft® Excel spreadsheet to evaluate the techno-economic analyses. The techno-economic spreadsheet tool allowed to optimise the process parameters such as loading rate, temperature and, residence time and evaluate its impact on economic viability.

The summary of the total capital investment (TCI), operating cost net present value (NPV) and internal rate of return (IRR) for all four scenarios is shown in the table below. The TCI values for Scenarios 1, 2, 3 and 4 are $49.1, $86.1, $127.3 and $139.2 million, respectively. The TCI value for Scenario 1 (BAU case) is the lowest. The TCI value for scenario 4 (AD plus 20% Partial WAO of biosolids) is the highest due to additional equipment like gravity pressure vessel (GPV), heaters, pumps and compressors. The annual operating costs for Scenarios 1, 2, 3 and 4 are $10.02, $23.3, $18.96, and $21.69 million, respectively. The annual operating cost for Scenario 1 is the lowest than the other three scenarios. The high annual operating costs noticed for other 3 scenarios are due to the extra utility consumption by additional process equipment such as TH and WAO units, GPV, additional pumping equipment, heat exchangers and phase separators. The annual operating cost for Scenario 2 (TH plus AD) is the highest due to the relatively higher amount of heating involved in the TH process. It should be noted that these experiments did not test the kinetics of a GPV process, and as such, the cost analysis does not exactly correlate with the estimated benefits.

Two other essential profitability indicators from the cash flow analysis are net present value (NPV) and internal rate of returns (IRR). NPV is the sum of future cash flows in today's dollars and a higher NPV value for a project will make the project attractive for investment. IRR on the other hand is the discount rate at which the project breaks even. NPV is a measure of cash profit, while IRR is a measure of the efficiency of capital utilisation.

As indicated in the table below, the NPV values for Scenarios 1, 2, 3 and 4 are -$78.5, -$164.5, -$169.1 and -$191 million, respectively. All NPV values are negative due to very low revenue generated in all four scenarios compared to the higher capital investment and operating costs. An electricity price of $0.085/kWh and a heating value of 0.038 GJ/m^3 CH_4, corresponding to an electrical conversion efficiency of 35%, respectively, were used to estimate the income. The financial attractiveness of the project can be found by determining the scenario with the lowest negative NPV, which is scenario 1 (-$78.5 million). Among the other three scenarios that include sludge treatments, Scenario 2, which includes TH followed by AD, has the lowest negative NPV, ascribed to the highest electricity sales due to the highest methane volume generation. The IRR values for Scenarios 2, 3 and 4 are 93%, 85%, and 88%, respectively. The IRR value for Scenario 1 is not included because it was higher than 100%, which was considered unrealistic. Among the other three scenarios, Scenario 2, which includes TH followed by AD, has the highest IRR, ascribed to the highest electricity sales for this scenario. Thus, comparing all three scenarios using the NPV and IRR values, it can be concluded that Scenario 2 (TH followed by AD) will lead to the least negative NPV and highest IRR values, due to the highest amount of electricity produced.

Short summary of the economic findings for a plant processing 86,400 tpa (1 kg/s)

	Unit	Scenario 1 (BAU of WAS)	Scenario 2 (TH of WAS followed by AD)	Scenario 3 (AD WAS followed by 100% WAO)	Scenario 4 (AD of WAS followed by 20% WAO)
CapEx	$ million	26.9	47.1	69.6	76.2
Total CapEx including EPCM[1]	$ million	31.5	55.1	81.5	89.2
TCI including contingency. (22% inherent risk and 28% contingency risk)	$ million	49.1	86.1	127.3	139.2
Total OpEx	$ million /year	10.02	23.30	18.96	21.69
Total revenue[2]	$ million /year	0.27	0.35	0.27	0.4
IRR	%	Very high	93	85	88
NPV	$ million	-78.5	-164.5	-169.1	-191

Note – 1 - engineering, procurement, and construction management. 2 – savings in transport costs of approximately $100/t were not included.

The techno-economic analyses showed that an AD plant with TH would be a better option for than an AD plant with WAO treatment. However, both technologies are 1.87 to 2.7 times more expensive than the existing base case scenario.

Both experimental and simulation studies thus suggests that further optimisation of the treatment conditions for WAO in terms of solids loading, treatment temperature and oxygen concentration needs to be determined. A pilot-scale study is needed to examine the effects of above process parameters before scaling up scenarios can be evaluated. As only a few studies on GPV were performed and the data is limited or seldom reported in literature, we recommend that Sydney Water further investigates optimising the process parameters for GPV technology.

1. Introduction

1.1 Project background

Annually, Sydney Water generates 58 GWh_{el} of electricity and 61 GWh_{th} of thermal in cogeneration plants from the biogas (~60% methane) produced from its anaerobic digestion (AD) facilities. The produced biogas can supply 40-60% of a tertiary treatment plant's electricity requirement. The residual organic rich solids (biosolids) and the digestate still contain residual energy (25-30%). Sydney Water needs to establish if the remaining energy can be economically recovered.

Wastewater treatment plants (WWTPs) in general consume a significant amount of energy, representing about 1% of the grid electricity across Australia. In 2019/20, Sydney Water's WWTPs consumed 188 GWh of grid electricity costing $23M. Surplus electricity produced through AD facilities will reduce the energy cost for the WWTP, avoid carbon emissions and generate additional heat that can be used for other purposes. There is potential for WWTPs to become net energy generators rather than significant energy consumers.

Sydney Water treats 500 GL of sewage and produces 180,000 tonnes of biosolids from all its wastewater treatment facilities annually, and the generated biosolids are trucked inland and reused as compost or soil conditioner. However, due to long trucking distances of around 300 km, transport costs of biosolids are high (approx. $100/t) and also lead to significant fossil fuel greenhouse gas (GHG) emissions. Reduction in the volume of biosolids produced will have a direct impact on trucking costs and the associated GHG emissions.

Sydney Water is constructing a thermal hydrolysis (TH) pretreatment plant to reduce the quantity and improve the quality of biosolids. However, current thermal pretreatment technologies, such as the TH process, are energy intensive and only result in small positive net energy generation. Therefore, this project aims to assess the energy balance of TH of waste activated sludge (WAS), the excess biomass from an activated sludge process removed during the secondary treatment, pretreatment (TH-WAS) and wet air oxidation (WAO) treatment of digestate obtained from AD of primary sludge (WAS Digestate). Both treatments will be compared with their business-as-usual (BAU) scenarios in terms of net energy balance, the reduction in biosolids amounts, the associated costs and GHG emissions from disposal of the residual biosolids to landfill.

1.2 Aim and objectives

1. To evaluate the technical feasibility and economic viability of TH pretreatment of concentrated WAS and WAO treatment of concentrated biosolids from AD (DS) on biogas production
2. To carry out mass and energy balances for the two treatment processes
3. To analyse the chemical composition and biomethane potential of concentrated and treated WAS and AD digestate.

2. Literature review

2.1 Introduction

Sewage sludge contains large amounts of water, microorganisms and mineral components, which make sludge treatment and disposal difficult. The microbial-originated extracellular polymeric substances (EPS) generated during sludge treatment process are a complex mixture of biopolymers consisting mainly proteins, nucleic acids, uronic acids, humic substances, polysaccharides, and lipids. EPS is relatively recalcitrant to anaerobic and aerobic digestion by nature. Therefore, pretreatment of sludge is essential and has shown to account for 50% of operational costs of WWTP (Kroiss, 2004; Neyens and Baeyens, 2003; Spinosa and Vesilind, 2001).

Hydrolysis is considered as the rate-limiting step during the AD of sewage sludge (Appels et al., 2008). TH pretreatment at high temperatures and pressures has been shown to be effective in breaking down the complex molecular compounds and cellular content of sewage sludge, thereby increasing the hydrolysis of the EPS in the pretreated sewage sludge. As a result, the biodegradability and biogas production of the TH pretreated sewage sludge is improved. This is particularly effective when treating WAS that contains bacterial cells, which are not easily biodegradable (Chen et al., 2012). In addition, dewaterability of WAS is also improved (Neyens and Baeyens, 2003). At high temperatures and pressures, WAO treatment oxidises organic and inorganic substances in an aqueous solution using air or oxygen to break down the complex molecules into simpler components or convert them into water and carbon dioxide with complete oxidation. The technology is comparable with incineration but with no harmful emissions such as nitrous oxides (NOx) and can also produce useful intermediate products such as acetic acid (Strong et al., 2011).

Among different sludge treatment technologies, hydrothermal technologies are the most efficient and economically viable. The hydrothermal technologies are classified into two main groups, namely oxidative and non-oxidative. In oxidative pretreatment technologies, WAO is considered the most popular and is usually carried out at high temperatures (>200°C) in the presence of an oxidant such as oxygen gas or hydrogen peroxide (Singh & Garg, 2022). The early WAO technology, named the Zimmerman process or Zimpro, used air at high pressure to combust organic compounds suspended or dissolved in the water. Almost all (95%) of the organic matter was oxidised at temperatures up to 300°C and pressures up to 175 bar. The main products were carbon dioxide, nitrogen, ammonia, ash and small amounts of acetic acid. A modified version of this process which operated at lower temperatures (<200°C) was later used to treat municipal sewage sludge. This process, called low pressure oxidation, involved very little oxidation and is in fact more similar to the TH processes (Camacho et al., 2008). However, many plants faced issues with corrosion and high energy costs, and eventually closed (Odegaard, 2004; Debellefontaine and Foussard, 2000).

TH is a non-oxidative pretreatment technology and is the most commercially proven technology carried out at a lower temperature range (100–200°C) without the addition of oxidants (Hu et al., 2020). TH is generally used as a pre-treatment step prior to AD, whereas WAO is used as a polishing step for the digestate from AD, particularly the concentrated biosolids (DS). A TH concept was found during the 1960s in the form of the Porteous process, which involved applying heat treatment to sludge to improve its dewaterability before incineration (Camacho et al., 2008). The Porteous process operated at higher temperatures (>200°C) than today's TH processes and resulted in an end product which could be dewatered to 40–60% solids content without the aid of chemicals (Hecht and Duvall, 1975). However, technical problems, issues with odour and economic factors led to most plants shutting down (Kepp et al., 2000). The more recent TH processes developed by Cambi were a result of research work which showed advantages of operating at lower temperatures (150–200°C). The resulting processes were designed around an optimum temperature of 170°C which gave the best compromise between improved dewaterability at higher temperatures and better digestibility at lower temperatures. Currently, Cambi is the most widely used TH process in the world (Maugans and Ellis, 2002; Camacho et al., 2008).

The main difference between WAO and TH is that oxidation reactions are desired in WAO processes, whereas they are not in TH processes; oxidation is achieved by the addition of an appropriate oxidant such as oxygen gas or hydrogen peroxide. The TH process is largely used as a pre-treatment prior to AD in sludge treatment due to its ability to alter sludge properties. As oxidation is an ultimate method for organic waste destruction, it is typically used as one of the final processes in sludge treatment.

Previously, the effect of WAO and TH processes in the treatment of aqueous wastes, industrial waste (Debellefontaine and Foussard, 2000) and sludge (Hii et al., 2014) were reviewed. Appels et al. (2008) and Carrere et al. (2010) also reviewed pretreatment processes used to improve AD and presented TH as one of the methods. This review presents an overview of the TH pretreatment of WWTP sludge entering the AD process and WAO treatment of digestate from the AD process, in particular, the process conditions, differences in process mechanism and their effects on chemical composition and methane yields.

2.2 Mechanisms of thermal hydrolysis (TH) and wet air oxidation (WAO) treatments

2.2.1 Mechanism of TH

Figure 1 presents the mechanism of TH pretreatment. TH was originally used for conditioning the sludge and improving its dewaterability. It was found to destroy the structural integrity of microbes in the sludge and facilitate the lysis of cell walls, which released cell contents. Higher temperatures for longer times were found to destroy more cell walls and insoluble proteins could also be broken down into more soluble amino acids. When used as a pretreatment before AD, TH can enhance biogas production and remove odour. Under high temperatures (130–200°C), the hydrolysis reaction breaks down complex molecules in sludge into simpler compounds. This results in higher bioavailability of sludge contents for AD (Hii, Baroutian, Parthasarathy, Gapes, & Eshtiaghi, 2014).

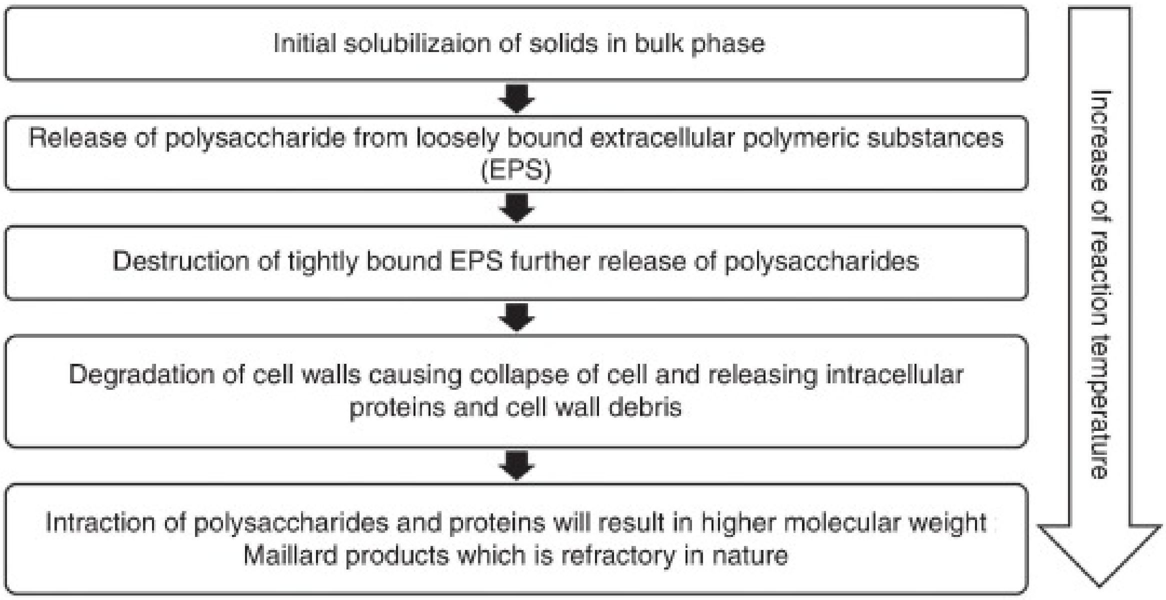


Figure 1. Mechanism of TH pre-treatment of sewage sludge (Singh & Garg, 2022).

2.2.2 Mechanism of WAO

Figure 2 presents the mechanism of WAO treatment. WAO at temperatures between 125-320°C and pressures between 0.5-30 MPa for 15–120 min is generally applied to treat aqueous waste and sludges and to produce acetic acid from lignocellulosic material (Castro & Agblevor, 2020). The type of application is usually determined by the range of temperatures used. Low temperature oxidation (100–200°C) is used for municipal and paper industry sludge, whereas medium temperature (200–260°C) is used for the treatment of ethylene spent-caustics and other industrial wastes, as well as for the regeneration of powdered activated carbon used in wastewater treatment. Higher temperatures (260–320°C) are used for sludge destruction and treatment of industrial wastewaters including organic industrial wastes such as pharmaceutical wastes and solvents. At the higher end of this temperature range, complete destruction of municipal, pulp and paper and other organic sludge is expected (Hii et al., 2014). WAO is particularly attractive because it yields a carbon-rich effluent which can be used in the production of economic products like acetic acid (Hii et al., 2014). In addition, it is considered environmentally benign as it involves no hazardous emissions and produces inert wastes (Tarpani, Alfonsin, Hospido, & Azapagic, 2020).

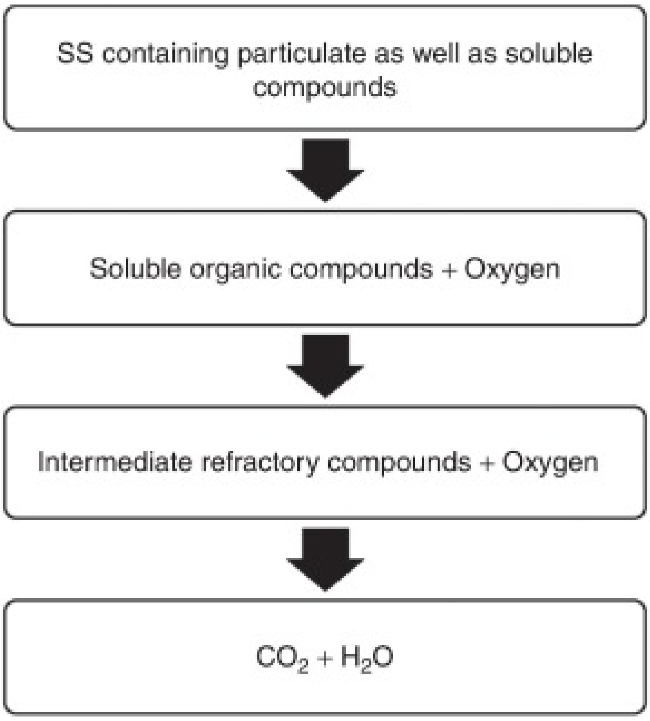


Figure 2. WAO reaction pathways (Singh & Garg, 2022)

2.2.3 Effects of TH pre-treatment on biomass composition and methane yields

Table 1 presents the literature studies showing the effect of TH on the chemical composition and methane yields of sludge along with the pretreatment conditions. Pretreatment residence time and operating temperature are shown to have a profound influence on the performance of the TH process. The best pretreatment temperature for TH of WAS has been shown to be between 160-180°C to achieve increased methane yield in subsequent AD and any temperature higher than this range showed a decrease in biodegradability of sludge and adversely impacted methane yields in the subsequent AD process (Bougrier, Albasi, Delgenès, & Carrère, 2006).

The effect of TH pretreatment temperature (62-175°C) and treatment time (15-60 min) showed that solubilisation of organic matter increased with increases in temperature and the best anaerobic degradability and methane production from TH of WAS was obtained when the temperature was between 150-175°C and the residence time was 30-60 minutes (Li and Noike, 1992). Moreover, the degree of solubilisation was dependent on the type of organic compounds present in the WAS. The solubilisation of carbohydrate and protein were found to increase as the treatment temperature increased from 120-175°C. At any given temperature, carbohydrate had the greatest solubilisation, followed by protein and lipid. Furthermore, the COD removal efficiency in the subsequent AD process using TH pretreated WAS increased with temperatures between 120-170°C. At temperatures >170°C, the COD removal efficiency decreased, indicating an optimal temperature of 170°C. Similarly, Carrere et al. (2010) also reported that most studies reported an optimum TH temperature between 160-180°C with treatment times of 30–60 min. Treatment temperatures above 170–190°C were shown to decrease the biodegradability and methane yields from the pretreated WAS. Previously, Donoso-Bravo et al. (2011) studied the effect of a TH reaction time between 0 and 30 min and a treatment temperature of <170°C on sewage sludge composition and AD performance. Results showed that the studied residence time was too short to notice any improvements in methane yields.

There are several issues associated with TH pretreatment including the formation of brown and ultraviolet-quenching compounds that contain recalcitrant dissolved organic nitrogen (rDON). For instance, melanoidins are produced by the Maillard reaction between the reducing sugar and amino groups. Factors influencing the formation of rDON in the TH pretreatment process include reactant availability, pH, heating temperature, time, and the presence of metallic ions (D. Zhang, Feng, Huang, Khunjar, & Wang, 2020). Higgins et al. (2017) investigated TH of a blend of dewatered primary and secondary sludge at five different temperatures (130, 140, 150, 160, 170°C) before feeding the digesters with 10.5% TS and sludge retention times (SRT) of 15 days. Higher temperatures improved the solids reduction and methane yields by about 5–6%. Increases in temperature reduced solid viscosity and increased the solids content after dewatering, but higher temperatures also increased the dissolved organic nitrogen and ultraviolet (UV) absorbance.

Decreasing the operating temperature from 165°C to 140°C resulted in a decrease in TH effluent colour with no significant impact on anaerobic biodegradability and chemical oxygen demand (COD) biodegradation of the sludge (Dwyer, Starrenburg, et al. (2008). Wilson and Novak (2009) investigated whether hydrolysis temperature can impact the previously observed drawbacks of TH such as high total ammonia nitrogen (TAN) and the production of highly coloured recalcitrant organics. The above authors found that solids loading rather than TH temperature appeared to be a more dominant factor for ammonia levels in downstream AD. Ammonification of protein by TH is a methanogenic inhibitor and the production of ammonia could be only marginally reduced by lowering the temperature to 150°C. At temperatures below 220°C, caramelisation did not appear to contribute to the UV254 (UV absorbance (UVA) per cm (cm^{-1}) or UV transmittance (UVT) %) of sludge hydrolysate, a test which provides a quick measurement of the organic matter in water). Dwyer, Kavanagh, and Lant (2008) investigated the simultaneous degradation of dissolved organic nitrogen (DON) and colour, present as a result of these compounds, using an advanced oxidation process (AOP). UV irradiation of hydrogen peroxide was used to produce hydroxyl radicals for oxidation. The oxidation process provided colour removal (99%), dissolved organic carbon (DOC) removal (50%) and DON removal (25%) at the optimal dosage of hydrogen peroxide (3,300 mg/L). Colour was caused by organic molecules with molecular weight >10 kDa. Oxidation caused a 41–15% reduction in DON, 29–14% reduction in DOC associated with molecular weight fraction >10 kDa and almost complete colour removal (87–93% of the colour). The degraded DON was consumed in the formation of ammonia and small molecular weight compounds. The same group (Dwyer, Griffiths, & Lant, 2009) focused on removal of the refractory coloured, organic nitrogen compound melanoidin in the Wetalla sewage treatment plant (Toowoomba, Australia) using hydrated aluminium sulphate as the alum source for coagulation of melanoidin. The authors in the above study have reported significant removal of colour (75%), DON (42%) and DOC (30%).

2.2.4 Effects of WAO treatment on biomass composition and methane yields

Table 2 presents studies from literature that demonstrate the effect of WAO on chemical composition and methane yields of sludge along with the pretreatment conditions. The WAO processes usually take place at temperatures between 150–320°C and pressures 20–150 bar. Maugans and Ellis (2002) divided the typical range of WAO temperatures into low (100– 200°C), medium (200–260°C) and high (260–320°C). The high temperature range is commonly used for sludge destruction and industrial wastewater treatment. The range of low temperatures is used for sludge-conditioning purposes. However, oxidation is unlikely to occur at low temperatures and these processes are, in fact, TH processes. The WAO process typically becomes energetically self-sufficient at medium and high temperature ranges.

The effect of WAO treatment on the AD digestate before dewatering and recycling the filtrate back to AD was investigated by Y. Zhang, Li, and Li (2021). Results showed that carbohydrate and protein were broken into acetic acid and other volatile fatty acids (VFAs) while humic acids remained as refractory species. The above authors also reported that incorporation of WAO in an AD process showed peak removal rate of 83.3% for sludge organic matter at 240°C with a 42.7% conversion rate to methane although ammonia was removed prior to AD. The corresponding values at 180°C were 71% organic removal rate and 53.3% conversion rate to methane compared with the no-WAO treatment AD system. Similarly, Thomsen, Anastasakis, and Biller (2022) investigated the application of non-catalytic WAO in the aqueous phase from hydrothermal liquefaction of sewage sludge with the purpose of reduction of organic matter and heat production at different temperatures (200–350°C), residence times (2–180 min) and excess oxygen. Results showed that 97.6% reduction of COD and 96.1% reduction of the total organic carbon (TOC) was achieved at the highest temperature and retention time. The minimum energy requirement achieved was 9.6 kWh/kg COD removed at 200°C for 180 min, and the exothermic reaction of the process covered 28.3% of the required heat.

The effects of WAO and alkaline WAO (AWAO) at 170°C and 0.4 MPa air for 30 minutes on the structure and biomethanation kinetics of water hyacinth was reported by (Castro & Agblevor, 2020). The study showed that the methane production rate [NmL CH_4/g feed.day] during the biomethanation of water hyacinth increased by 63% after WAO and 117% after AWAO. AWAO enhanced the biomethane potential [N mL CH_4/g feed] of water hyacinth by 24%. Similarly, Bhoite and Vaidya (2018b) experimented with iron based catalysts for WAO applied to biomethanated spent wash with high COD (40,000 mg/L) in a sugar factory to increase biogas recovery and the efficacy of wash water treatment at 175°C and 0.69 MPa. Approximately 1.2 Nm^3 biogas (72% methane by volume) was produced per cubic meter of the wastewater, whereas without catalytic WAO pre-treatment, the yield of biogas (11% methane) was 1 Nm^3 per cubic meter of wastewater. The same authors, Bhoite and Vaidya (2018a), in a similar experiment applied catalytic WAO using the same temperature and oxygen pressure on biomethanated distillery wastewater and achieved a 73% reduction in COD and 1.1 Nm^3 of biogas (69% methane) per cubic meter of wastewater. Finally, Pham et al. (2021) studied the effect of WAO with the addition of potassium carbonate as a catalyst on sewage sludge and reported a methane yield of 202 mL/g VS_{fed} from WAO pretreated sludge without catalyst and 277 mL/g VS_{fed} from WAO pretreated sludge with 10 w/w% of potassium carbonate added at 180°C with 30 min of residence time.

Table 1. Summary of TH pretreatment conditions and its impact on solubilisation and methane yields of WAS (Singh & Garg, 2022).

Substrate	Treatment conditions	Main findings	Reference
Primary sludge and WAS	Temperature: 120°C-220°C; Time: 45 min	From primary sludge, volatile suspended solids (VSS) reduction = 35% at 220°C and 20% enhancement in biomethane potential (BMP) increase at 160°C For WAS, VSS reduction = 55% at 220°C and maximum BMP = 1.5 times of that with 'control' at 160°C	Pinneekamp (1988)
WAS	Temperature: 60°C-175°C; Time: 15 min-120 min	COD, carbohydrate, protein, and lipids solubilisation of 50%, 45%, 45% and 15%, respectively, at 175°C for 60 min and 20% enhancement in BMP increase at 170°C for 1 h	Li and Noike (1992)
WAS	Temperature: 115°C-180°C; Time: 5 min-200 min	VSS solubilisation = 30% and methane production of 1.9 times that of 'control' at 180°C for 1 h	Tanaka et al. (1997)
WAS	Temperature:130°C-170°C; Time: 60 min	COD solubilisation = 60%, total suspended solids (TSS) solubilisation = 51% and maximum BMP was 1.54 times the value obtained with 'control' at 170°C for 1 h	Valo et al. (2004)
WAS	Temperature: 130°C-170°C; Time: 30 min	Methane production was 1.8 times the value obtained with 'control' at 170°C for 30 min	Bougrier et al. (2006)

Dewatered WAS	Temperature: 130°C-190°C; Time: 15 min-60 min	Methane production was 1.4 times the value obtained with ‘control’ at 170°C for 30 min	Zhang et al. (2014)
WAS	Temperature: 140°C-180°C; Time: 120 min; pH: 3.3 and 9.3	Improvement in dewaterability at 180°C and pH of 3.3 Dissolved organic carbon concentration = 3,115 mg/L at 180°C (pH = 9.3) from an initial value of 88 mg/L in case of untreated SS Maximum volatile suspended solids (VSS) solubilisation = 45% at 180°C (pH = 9.3)	Malhotra and Garg (2019)
WAS	Temperaure:100°C-220°C; Time: 30 min	Maximum COD solubilisation = 51.2% at 220°C Maximum BMP was 1.8 times the value obtained with ‘control’ at 180°C	Jeong et al. (2019)
WAS	Temperature: 140°C-180°C; Time: 60 min	Better dewaterability of secondary sludge (SS) treated at 180°C with time to filter (TTF) as 4.7 s.L/g Maximum BMP was twice that obtained for control after TH at 160°C	Singh (2020)

Table 2. Summary of WAO treatment conditions and its impact on methane yields of WAS (Singh & Garg, 2022)

Substrate	Reaction conditions	Main findings	Reference
WAS	Temperature: 180°C-240°C; Time: 80 min; Pressure: 40 atm-60 atm; O_2 flow rate of 0.5 L/min-3 L/min.	At 240°C and 50 atm, total chemical oxygen demand (tCOD) reduction, total suspended solids (TSS) reduction and organic acid were found as 70%, 77% and 660 mg/L, respectively	Chung et al. (2009)
Thickened WAS	Temperature: 150°C; Time: 120 min; O_2 dose as 0 –120% of stoichiometric oxygen requirement (SOR).	VSS removal = 60% with 120% O_2 of SOR and AD of the treated SS gives total organic carbon (TOC) and VSS removal of 50% each at SOR of 100%	Abe et al. (2011)
Mixed sludge	Temperature: 220°C; Time: 120 min; Pressure: 19.7 atm; Oxidant: O_2	VSS and TSS reduction = 93% and 83%, respectively and Acetic acid concentration = 146 mg/g of dry sludge	Strong et al. (2011)
Thickened WAS	Temperature: 160°C-200°C; Time: 200 min; Pressure: 40 atm-80 atm; Oxidant: O_2	TSS, VSS and tCOD reduction = 71%, 92% and 63%, respectively at 200°C and O_2 pressure of 60 atm	Urrea et al. (2014)
Mixed sludge	Temperature: 220°C-240°C; Time: 60 min; O_2: biomass ratio (w/w) of 1:1–2:1	Acetic acid concentration = 2,750 mg/L was observed at 230°C and O_2 to biomass ratio of 2:1	Baroutian et al. (2015)
WAS	Temperature: 140°C-180°C; Time: 2 h-5 h; SOR: 0 - 50%	tCOD reduction = 56.5% at a temperature of 180°C after 5 h run under acidic condition and SOR = 50%	Malhotra and Garg (2019)

2.2.5 Economics of the existing WAO and TH technologies and those under development

Singh and Garg (2022) present a list of commercially available technologies in Table 3. As WAO is generally performed at higher temperatures and pressures than TH, the costs of WAO treatment are high due to higher reactor costs. This can be addressed by using catalytic WAO treatment, which may allow for operation of the reactor at less severe operational conditions. Moreover, corrosion of the WAO reactor could occur due to the production of acidic compounds as part of the WAO process. To tackle this issue, sufficient alkalinity should be present and the material for WAO systems should be chosen wisely (Singh & Garg, 2022).

According to a life cycle cost study carried out in the UK, AD, pyrolysis and WAO can work at a profit with the negative overall life cycle costs of -£65, -£291 and -£26/1,000 kg dry matter, respectively (Tarpani & Azapagic, 2018).

Table 3. Commercially available TH technologies (Singh & Garg, 2022)

S. No.	Processes	Treatment conditions	Major Outcomes
Thermal hydrolysis (TH)			
1	Cambi process	Temperature: 160-180°C; Pressure: 6 bar; Time: 20-30 min	50% more methane generation compared to conventional treatment. Destruction of pathogens and odourous bacteria. Improved biosolids dewatering Lower carbon footprint.
2	BioTHELYS ® process	Temperature: 165°C; Pressure: 9 bar; Time: 30 min	30-50% more biogas. 25-35% less dry matter. No odour. A pasteurised digestate.
Wet Air Oxidation (WAO)			
1	Athos process	Temperature: 235°C; Pressure: 45 bar; Time: < 60 min; Oxygen as an oxidant	A concentrate of inorganic matter accounting for only 1-2% of the initial liquid sludge quantity. Heavy metal stabilisation in a non-leachable form. Reduction in the greenhouse effect (50% drop in CO_2 emissions compared to incineration).
2	VerTech process	Temperature:275°C; Pressure: 100 bar; depth of reactor = 1,280 m	Accumulation of volatile fatty acids (VFAs) in liquid phase. Reduction of TSS 75-80%. Reduction of VSS almost 100%.
3	Zimpro process	Temperature: 200-300°C; Pressure: 40-60 bar; Time: 60 min.	75-90% of VSS oxidised or transformed to simple dissolved compounds.

2.4 Aspen Plus Models

Among all the AD models reported in the literature, only a few were simulated using Aspen Plus. Aspen Plus is a very beneficial tool in simulating the processes. It is often used to conceptualise and optimise the design, improve operational conditions and manage operations in various process industries. Furthermore, it can handle complex streams with multiple recycle streams. The software has a library system called Databank, which stores chemical, physical and dynamic properties of substances and compounds. Databank also includes phase equilibrium and pure component data for many polymers, solids and chemicals. Aspen Plus allows the user to integrate with the Microsoft Excel environment and FORTRAN programs for calculation purposes. Therefore, it is usually preferred over many other platforms.

Even though Aspen Plus offers all these advantages, it is still challenging to model biological processes using the software, especially processes such as AD, since such a system has not been integrated yet into the software (Nguyen, 2014). Several researchers have attempted to simulate an AD system to study the biogas and bioliquid generation, validating them with experimental data. Below is a review of some Aspen Plus models and methods reported in the literature.

2.4.1 Model A

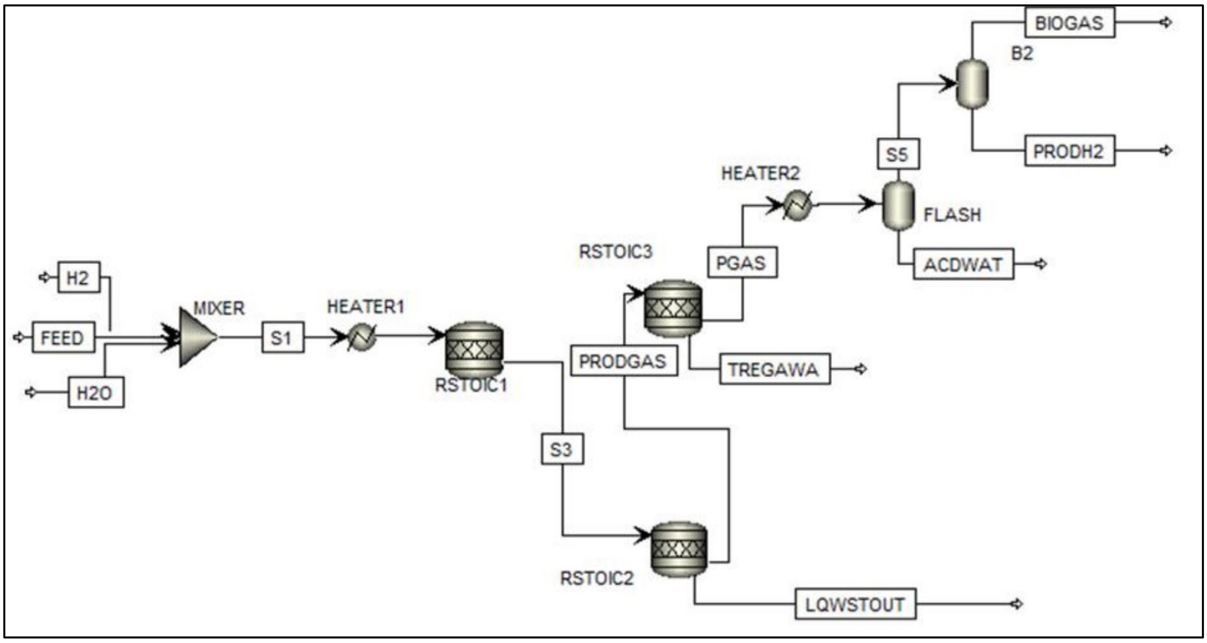


Figure 3. RStoic reactors model by Ravendran, Abdulrazik & Zailan (2019)

The above model was published by (Ravendran, Abdulrazik, & Zailan, 2019). They simulated AD to investigate biogas production and the environmental and economic potentials of cow manure. The model expresses the four AD stages in three reactor blocks, RSTOIC 1, RSTOIC 2 and RSTOIC 3, operating under 37, 45 and 47°C, respectively. Although all three reactors operate under 3 bars, each reactor plays a different role. The first reactor was used to carry out hydrolysis reactions. The second reactor was used for the acidogenesis and acetogenesis steps, while the last one was used for methanogenesis. A separation section is also introduced, as shown in Figure 3. The produced gas is heated and separated by flash separator in this section. The study focuses mainly on methane, hydrogen, and carbon dioxide.

This study was limited to specific components as only eight reactions were presented in the model. Although the study attempted to validate the model, the effects of some important complex compounds on the process were ignored. Moreover, the paper did not mention the type of reaction kinetics used in the model. Presenting all the reactions involved in the AD using only RStoic increases the gap between the experimental and simulated data since this type of block is mainly used for stoichiometric reactions where their rates are determined by, for example, the extent of reaction or conversation. According to the previous reaction models, AD has some power-law reactions that depend on kinetics. Therefore, those limitations may lead to significantly inaccurate results, especially for studies that examine AD products. (Nguyen, Heaven, & Banks, 2014) used only an RStoic reactor for simulating the AD process. The focus of the study was the economic analysis of the electricity and power requirement for an ammonia stripping process combined with a power plant. However, the study covers the AD treatment using the gas produced by the AD system. The generated gases were hydrogen sulphite (H_2S), oxygen (O_2), CH_4 and CO_2. Since the main aim was not to model AD, the authors followed a theoretical stoichiometric method introduced by (Symons & Buswell, 1933) called the Buswell equation. This method estimates the generation of four substances based on reaction stoichiometry. It is important to mention that both models used the non-random two-liquid model (NRTL) property method in Aspen Plus simulation.

2.4.2 Model B

(Rajendran, Kankanala, Lundin, & Taherzadeh, 2014) developed a novel approach and model simulating the AD based on the four pathways of sludge biodegradation. The research focused on describing the system's behaviour on the software model to obtain results close to reality (i.e., experimental data). Figure 4 presents the approach for AD process simulation using Aspen Plus.

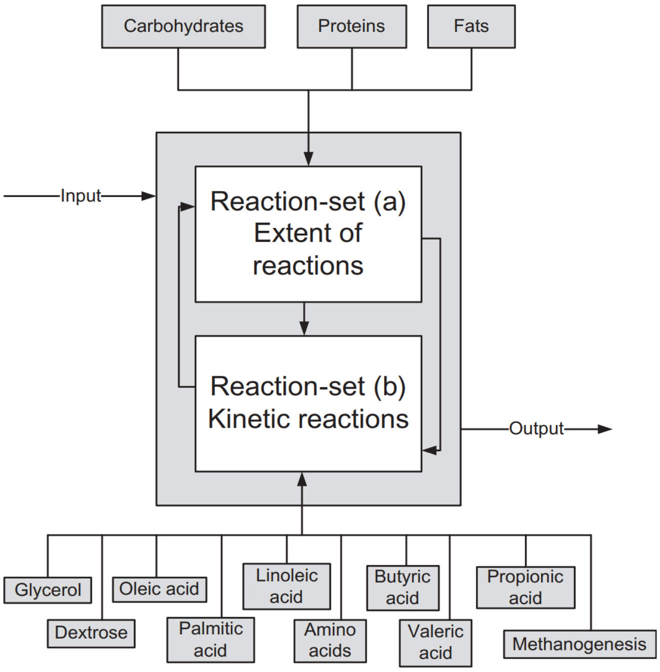


Figure 4. Process simulation model (PSM) novel approach developed by Rajendran et al. (2014).

The process simulation model (PSM) model uses 46 reactions to represent the biodegradation of sludge. The reactions cover pH, kinetic rates, loading rate, inhibitions, volume and ammonia. Here, the reactions are divided into two groups: 1) hydrolysis reactions based on stoichiometry reactions determined by the extent of reaction, and 2) acidogenic, acetogenic, and methanogenic reactions, which are expressed using kinetics. The concept behind the approach is to separate reaction groups in Aspen Plus based on the reaction order and blocks' function on the software. The hydrolysis reactions group is simulated using a separate RStoic block because these reactions occur first, which is described in many studies as the rate-limiting step. In contrast, the other group was simulated using rigorous continuously stirred tank (RCSTR) because this block presents the kinetic reactions. Rajendran et al. (2014) provided access to the model via the Swedish database http://hdl.handle.net/2320/ 12358.

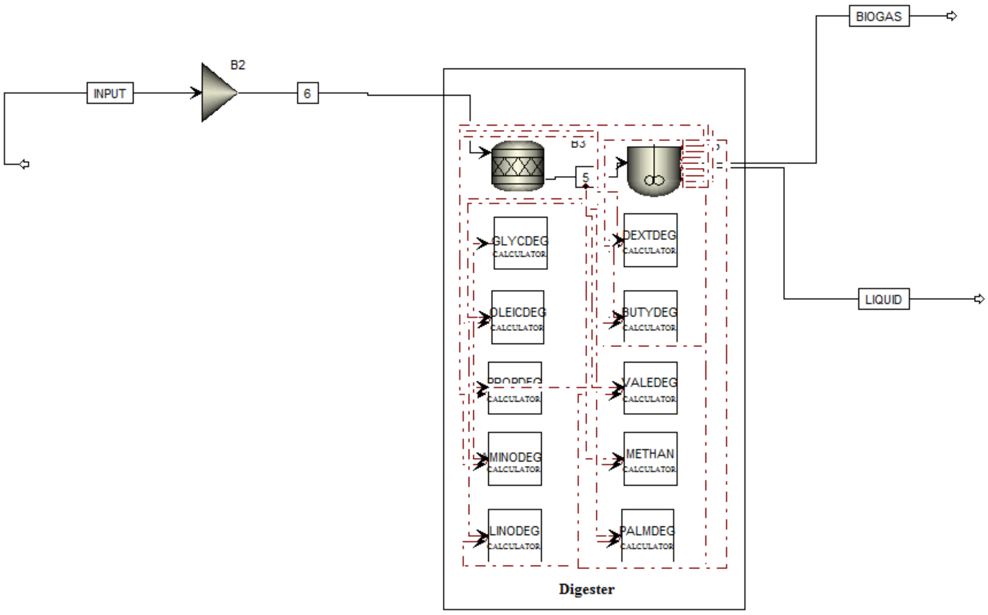


Figure 5. PSM Aspen Plus model developed by Rajendran et al. (2014)

Figure 5 shows how the calculator-blocks are connected to the model. In this case, reaction rates were determined for acidogenesis, acetogenesis, and methanogenesis using Fortran in the calculators. Based on the reactants (i.e., stream 5 components), calculators use different variables such as VFA, which involve pH calculation, ammonium ions, temperature, hydrogen, etc., for the computation. Although temperature, ammonia and pH were fixed, the calculators estimate ammonia inhibition and microorganic specific growth rates. The model included a combination of reaction models, proposed by Batstone et al. (2002), (Andrews, 1968; Angelidaki, Ellegaard, & Ahring, 1999). The reactions set used in those models are presented in Table 4 and Table 5.

Table 4. Hydrolysis reaction group (Rajendran et al. 2014)

No.	Compound	Hydrolysis reaction	Extent of reaction
1	Starch	$(C_6H_{12}O_6)_n + H_2O \rightarrow n\ C_6H_{12}O_6$	0.6 ± 0.2
2	Cellulose	$(C_6H_{12}O_6)_n + H_2O \rightarrow n\ C_6H_{12}O_6$	0.4 ± 0.1
3	Hemicellulose	$C_5H_8O_4 + H_2O \rightarrow 2.5\ C_2H_4O_2$	0.5 ± 0.2
4	Hemicellulose	$C_5H_8O_4 + H_2O \rightarrow C_5H_{10}O_5$	0.6 ± 0.0
5	Xylose	$C_5H_{10}O_5 \rightarrow C_5H_4O_2 + 3\ H_2O$	0.6 ± 0.0
6	Cellulose	$C_6H_{12}O_6 + H_2O \rightarrow 2\ C_2H_6O + 2\ CO_2$	0.4 ± 0.1
7	Ethanol	$2\ C_2H_6O + CO_2 \rightarrow 2\ C_2H_4O_2 + CH_4$	0.6 ± 0.1
8	Soluble protein	$C_{13}H_{25}O_7N_3S + 6\ H_2O \rightarrow 6.5\ CO_2 + 6.5\ CH_4 + 3\ H_3N + H_2S$	0.5 ± 0.2
9	Insoluble protein (I.P)	$I.P + 0.3337\ H_2O \rightarrow 0.045\ C_6H_{14}N_4O_2 + 0.048\ C_4H_7NO_4 + 0.047\ C_4H_9NO_3 + 0.172\ C_3H_7NO_3 + 0.074\ C_5H_9NO_4 + 0.111\ C_5H_9NO_2 + 0.25\ C_2H_5NO_2 + 0.047\ C_3H_7NO_2 + 0.067\ C_3H_6NO_2S + 0.074\ C_5H_{11}NO_2 + 0.07\ C_6H_{13}NO_2 + 0.046\ C_6H_{13}NO_2 + 0.036\ C_9H_{11}NO_2$	0.6 ± 0.1
10	Triolein	$C_{57}H_{104}O_6 + 3\ H_2O \rightarrow C_3H_8O_3 + 3\ C_{18}H_{34}O_2$	0.5 ± 0.2
11	Tripalmate	$C_{51}H_{98}O_6 + 8.436\ H_2O \rightarrow 4\ C_3H_8O_3 + 2.43\ C_{16}H_{34}O$	0.5 ± 0.3
12	Palmito-olein	$C_{37}H_{70}O_5 + 4.1\ H_2O \rightarrow 2.1\ C_3H_8O_3 + 0.9\ C_{16}H_{34}O + 0.9\ C_{18}H_{34}O_2$	0.6 ± 0.2
13	Palmito-linolein	$C_{37}H_{68}O_5 + 4.3\ H_2O \rightarrow 2.2\ C_3H_8O_3 + 0.9\ C_{16}H_{34}O + 0.9\ C_{18}H_{32}O_2$	0.6 ± 0.2

Table 5. Acidogenic, acetogenic, and methanogenic reactions group (Rajendran et al. 2014)

No.	Compound	Chemical reactions	Kinetic constant
Amino acid degradation reactions			
1	Glycine	$C_2H_5NO_2 + H_2 \rightarrow C_2H_4O_2 + H_3N$	$1.28 * 10^{-02}$
2	Threonine	$C_4H_9NO_3 + H_2 \rightarrow C_2H_4O_2 + 0.5\ C_4H_8O_2 + H_3N$	$1.28 * 10^{-02}$
3	Histidine	$C_6H_8N_3O_2 + 4\ H_2O + 0.5\ H_2 \rightarrow CH_3NO + C_2H_4O_2 + 0.5\ C_4H_8O_2 + 2\ H_3N + CO_2$	$1.28 * 10^{-02}$
4	Arginine	$C_6H_{14}N_4O + 3\ H_2O + H_2 \rightarrow 0.5\ C_2H_4O_2 + 0.5\ C_3H_6O_2 + 0.5\ C_5H_{10}O_2 + 4\ H_3N + CO_2$	$1.28 * 10^{-02}$
5	Proline	$C_5H_9NO_2 + H_2O + H_2 \rightarrow 0.5\ C_2H_4O + 0.5\ C_3H_6O_2 + 0.5\ C_5H_{10}O_2 + H_3N$	$1.28 * 10^{-02}$
6	Methionine	$C_5H_{11}NO_2S + 2\ H_2O \rightarrow C_3H_6O_2 + CO_2 + H_3N + H_2 + CH_4S$	$1.28 * 10^{-02}$
7	Serine	$C_3H_7NO_3 + H_2O \rightarrow C_2H_4O_2 + H_3N + CO_2 + H_2$	$1.28 * 10^{-02}$
8	Threonine	$C_4H_9NO_3 + H_2O \rightarrow C_3H_6O_2 + H_3N + H_2 + CO_2$	$1.28 * 10^{-02}$
9	Aspartic acid	$C_4H_7NO_4 + 2\ H_2O \rightarrow C_2H_4O_2 + H_3N + 2\ CO_2 + 2\ H_2$	$1.28 * 10^{-02}$
10	Glutamic acid	$C_5H_9NO_4 + H_2O \rightarrow C_2H_4O_2 + 0.5\ C_4H_8O_2 + H_3N + CO_2$	$1.28 * 10^{-02}$
11	Glutamic acid	$C_5H_9NO_4 + 2\ H_2O \rightarrow 2\ C_2H_4O_2 + H_3N + CO_2 + H_2$	$1.28 * 10^{-02}$
12	Histidine	$C_6H_8N_3O_2 + 5\ H_2O \rightarrow CH_3NO + 2\ C_2H_4O_2 + 2\ H_3N + CO_2 + 0.5\ H_2$	$1.28 * 10^{-02}$
13	Arginine	$C_6H_{14}N_4O_2 + 6\ H_2O \rightarrow 2\ C_2H_4O_2 + 4\ H_3N + 2\ CO_2 + 3\ H_2$	$1.28 * 10^{-02}$
14	Lysine	$C_6H_{14}N_2O_2 + 2\ H_2O \rightarrow C_2H_4O_2 + C_4H_8O_2 + 2\ H_3N$	$1.28 * 10^{-02}$
15	Leucine	$C_6H_{13}NO_2 + 2\ H_2O \rightarrow C_5H_{10}O_2 + H_3N + CO_2 + 2\ H_2$	$1.28 * 10^{-02}$
16	Isoleucine	$C_6H_{13}NO_2 + 2\ H_2O \rightarrow C_5H_{10}O_2 + H_3N + CO_2 + 2\ H_2$	$1.28 * 10^{-02}$
17	Valine	$C_5H_{11}NO_2 + 2\ H_2O \rightarrow C_4H_8O_2 + H_3N + CO_2 + 2\ H_2$	$1.28 * 10^{-02}$
18	Phenyalanine	$C_9H_{11}NO_2 + 2\ H_2O \rightarrow C_6H_6 + C_2H_4O_2 + H_3N + CO_2 + H_2$	$1.28 * 10^{-02}$
19	Tyrosine	$C_9H_{11}NO_3 + 2\ H_2O \rightarrow C_6H_6O + C_2H_4O_2 + H_3N + CO_2 + H_2$	$1.28 * 10^{-02}$
20	Typtophan	$C_{11}H_{12}N_2O_2 + 2\ H_2O \rightarrow C_8H_7N + C_2H_4O_2 + H_3N + CO_2 + H_2$	$1.28 * 10^{-02}$
21	Glycine	$C_2H_5NO_2 + 0.5\ H_2O \rightarrow 0.75\ C_2H_4O_2 + H_3N + 0.5\ CO_2$	$1.28 * 10^{-02}$
22	Alanine	$C_3H_7NO_2 + 2\ H_2O \rightarrow C_2H_4O_2 + H_3N + CO_2 + 2\ H_2$	$1.28 * 10^{-02}$
23	Cysteine	$C_3H_6NO_2S + 2\ H_2O \rightarrow C_2H_4O_2 + H_3N + CO_2 + 0.5\ H_2 + H_2S$	$1.28 * 10^{-02}$
Acidogenic reactions			
24	Dextrose	$C_6H_{12}O_6 + 0.1115\ H_3N \rightarrow 0.1115\ C_5H_7NO_2 + 0.744\ C_2H_4O_2 + 0.5\ C_3H_6O_2 + 0.4409\ C_4H_8O_2 + 0.6909\ CO_2 + 1.0254\ H_2O$	$9.54 * 10^{-03}$
25	Glycerol	$C_3H_8O_3 + 0.4071\ H_3N + 0.0291\ CO_2 + 0.0005\ H_2 \rightarrow 0.04071\ C_5H_7NO_2 + 0.94185\ C_3H_6O_2 + 1.09308\ H_2O$	$1.01 * 10^{-02}$
Acetogenic reactions			
26	Oleic acid	$C_{18}H_{34}O_2 + 15.2396\ H_2O + 0.2501\ CO_2 + 0.1701\ H_3N \rightarrow 0.1701\ C_5H_7NO_2 + 8.6998\ C_2H_4O_2 + 14.4978\ H_2$	$3.64 * 10^{-12}$
27	Propionic acid	$C_3H_6O_2 + 0.06198\ H_3N + 0.314336\ H_2O \rightarrow 0.06198\ C_5H_7NO_2 + 0.9345\ C_2H_4O_2 + 0.660412\ CH_4 + 0.160688\ CO_2 + 0.00055\ H_2$	$1.95 * 10^{-07}$
28	Isobutyric acid	$C_4H_8O_2 + 0.0653\ H_3N + 0.8038\ H_2O + 0.0006\ H_2 + 0.5543\ CO_2 \rightarrow 0.0653\ C_5H_7NO_2 + 1.8909\ C_2H_4O_2 + 0.446\ CH_4$	$5.88 * 10^{-06}$
29	Isovaleric acid	$C_5H_{10}O_2 + 0.0653\ H_3N + 0.5543\ CO_2 + 0.8044\ H_2O \rightarrow 0.0653\ C_5H_7NO_2 + 0.8912\ C_2H_4O_2 + C_3H_6O_2 + 0.4454\ CH_4 + 0.0006\ H_2$	$3.01 * 10^{-08}$
30	Linoleic acid	$C_{18}H_{32}O_2 + 15.356\ H_2O + 0.482\ CO_2 + 0.1701\ H_3N \rightarrow 0.1701\ C_5H_7NO_2 + 9.02\ C_2H_4O_2 + 10.0723\ H_2$	$3.64 * 10^{-12}$
31	Palmitic acid	$C_{16}H_{34}O + 15.253\ H_2O + 0.482\ CO_2 + 0.1701\ H_3N \rightarrow 0.1701\ C_5H_7NO_2 + 8.4402\ C_2H_4O_2 + 14.9748\ H_2$	$3.64 * 10^{-12}$
Methanogenic reactions			
32	Acetic acid	$C_2H_4O_2 + 0.022\ H_3N \rightarrow 0.022\ C_5H_7NO_2 + 0.945\ CH_4 + 0.066\ H_2O + 0.945\ CO_2$	$2.39 * 10^{-03}$
33	Hydrogen	$14.4976\ H_2 + 3.8334\ CO_2 + 0.0836\ H_3N \rightarrow 0.0836\ C_5H_7NO_2 + 3.4154\ CH_4 + 7.4996\ H_2O$	$2.39 * 10^{-03}$

Many researchers widely accepted the Rajendran et al. (2014) model because a PSM was valid for different types of waste. Since the model is accessible, it has been utilised and validated with experimental data by several studies (Harun, Hassan, Zainol, Ibrahim, & Hashim, 2019; Inayat et al., 2021; Menacho, Mazid, & Das, 2022; Wid, 2018). In addition, other studies have attempted to advance the model using different approaches (Bravo, Álvarez-Hornos, Penya-Roja, San-Valero, & Gabaldón, 2018; Lorenzo-Llanes, Pages-Diaz, Kalogirou, & Contino, 2020). For instance, Bravo et al. (2018) divided the reaction steps based on reaction order. They used the RStoic block for the hydrolysis and methanogenesis stages and RCSTR for the acidogenic and acetogenic phases.

3. Chemical composition and biomethane potential of WAS and anaerobic digestate

3.1 Introduction

The chemical composition and the biomethane potential of WAS and digestate with and without pretreatments were performed. Figure 6 presents the proposed process flow diagram of TH pretreatment of WAS (Fig. 6 above) and WAO treatment of AD digestate (Fig. 6 below) to be implemented at Sydney Water WWTPs. For both treatments, the substrate was concentrated by using a decanter centrifuge before performing the sludge treatments.

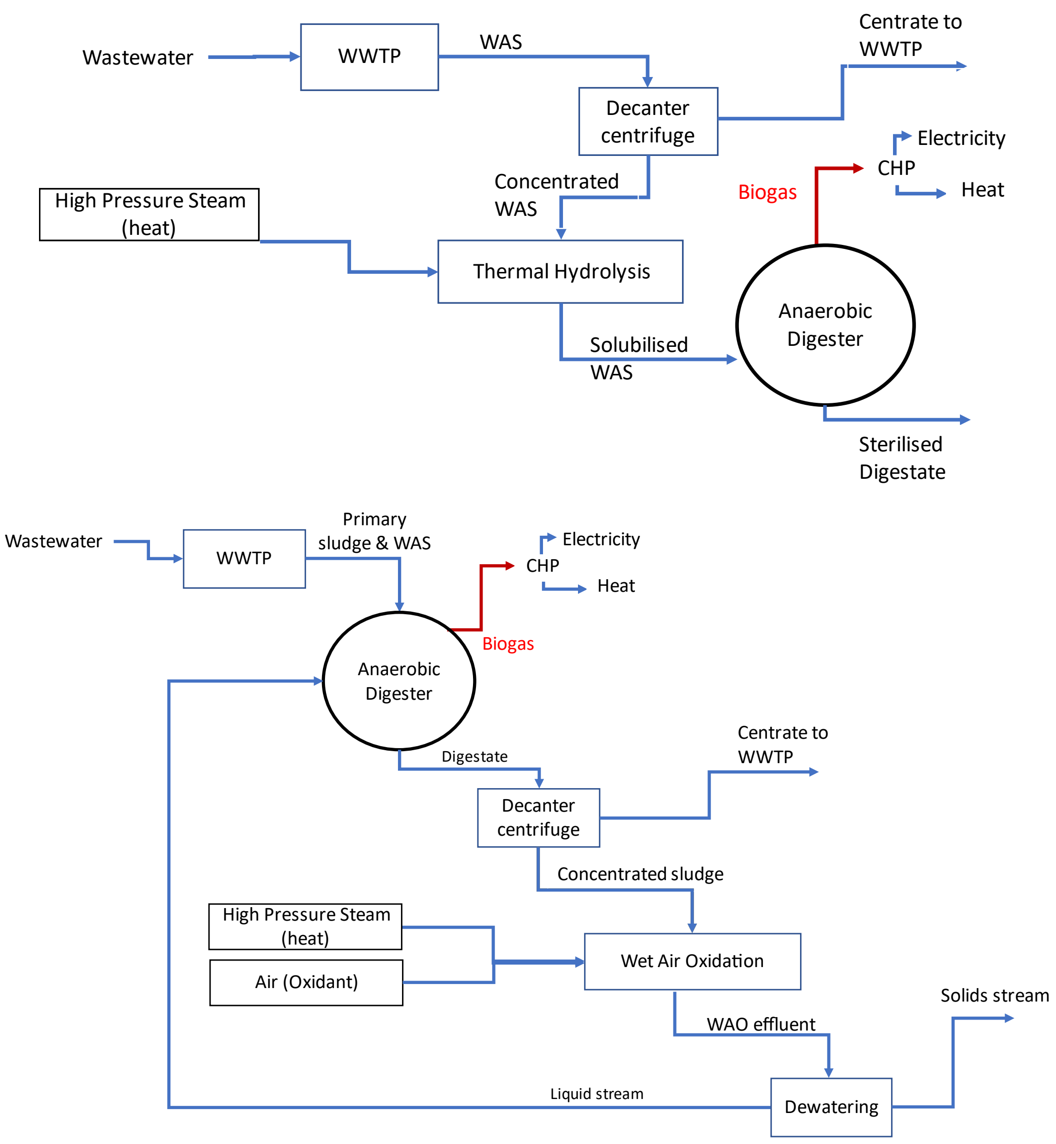


Figure 6. Process flow diagram of thermal hydrolysis (TH) of waste activated sludge (WAS) (above) and wet air oxidation (WAO) of anaerobic digestate concentrate (below) in the project.

3.2 Methodology

3.2.1 Inoculum

Mesophilically digested material from a full-scale AD plant (Luggage Point Wastewater Treatment Plant, Brisbane) treating both primary and secondary sludge was used as the inoculum for conducting biochemical methane potential studies. The inoculum was degassed by incubating at 37±1°C for a week to remove any remaining biodegradable fraction and residual methane.

3.2.2 Substrates

Raw WAS and anaerobic digestate were supplied by Sydney Water from its Glenfield WWTP in Sydney. Upon arrival at RMIT laboratory in Melbourne, the samples were stored at 4°C prior to further analyses/pretreatment. Untreated and pretreated samples were then shipped to Griffith University in Brisbane at 4°C.

3.2.3 TH pretreatment of WAS Concentrate (TH WAS) (Scenario 2)

The WAS substrate was centrifuged at 3,500 rpm for 1 hour and the concentrated WAS was then used in the TH pretreatment process. The operating conditions for TH are provided in Table 6.

Table 6. Pre-treatment conditions for TH of WAS Concentrate

Operating conditions	Required value
Temperature, °C	165
Angular velocity of agitator, rpm	300
Duration, minutes	30
Initial pressure (N_2), bars*	10
Maximum pressure recorded during run, bars	22

Note: * After purging with N_2

3.2.4 WAO of Digestate Concentrate (WAO Digestate Concentrate) (Scenarios 3 and 4)

The AD digestate sample (material obtained from AD of WAS), was centrifuged at 3,500 rpm for 1.5 hours and the Digestate Concentrate was then subjected to WAO. The pretreatment conditions for WAO under two operating conditions/extent of oxidation are provided in Table 7.

Table 7. Pre-treatment conditions for WAO of Digestate Concentrate (WAO-Digestate Concentrate)

Operating conditions	Run 1	Run 2
Temperature, °C	240	240
Angular velocity of agitator, rpm	300	300
Duration, minutes	30	30
Initial pressure (O_2), bars*	44	22
Maximum pressure recorded during run, bars	100	72
Oxygen loading, %	20%	10%

Note: * After purging with N_2

3.2.5 Biochemical methane potential of WAS Concentrate and Digestate Concentrate before and after TH and WAO, respectively.

Biochemical methane potential (BMP) tests were carried out in batch experiments using 160 mL glass bottles with working volume of 100 mL. To each assay, inoculum (80 g) and substrate were added to achieve an inoculum to substrate ratios (ISR) of 2.0. Distilled water was used to achieve the desired working volume. Assays with inoculum alone were used as controls. The experiments were performed in triplicate and incubated statically at 37±1°C in an incubator until no further biogas production was noticed for three consecutive measurements. Methane produced from the control assays (without sample) was subtracted from the sample assays. Cumulative methane production (mL) was corrected for standard temperature and pressure (STP) and methane yield was expressed as $Nm^3/kg\ VS_{added}$, determined as methane produced per unit of VS added to each assay.

3.3 Analytical methods

pH was measured with a pH meter (TPS pH meter). TS and VS were determined according to Standard Methods (APHA, 1998). Phosphate (PO_4-P) and ammonium nitrogen (NH_4-N) content in the samples were analysed using Quick chem 8000 Flow Injection Analyzer (FIA). Total Kjeldahl Phosphorus (TKP) and Total Kjeldahl Nitrogen (TKN) were analysed by using PerkinElmer, USA. The chemical composition of each sample was analysed using Fourier Transform Infra-Red (FTIR) spectrometer equipped with a multi-bounce Germanium Attenuated Total Reflection (Ge ATR) cell (Thermo Nicolet iS50). All spectra were obtained at a resolution of 4 cm^{-1}. The resultant data was corrected in Thermo Omnic with an ATR correction to account for the variable degree of penetration of the evanescent wave into the sample. All spectra were corrected according to the standard normal variate (SNV) method (Barnes, Dhanoa, & Lister, 1989).

Biogas composition (CH_4 and CO_2) was determined by using a gas chromatograph (GC), Shimadzu GC-2014, Japan. The GC was equipped with a Valco GC valve with a 1 mL sample loop, a Chin Carbon ST 100/120 packed column (2.0 m length; 1/16″ outer diameter, 1 mm inner diameter) and a thermal conductivity detector (TCD). The chromatograph's injector, oven, and detector temperatures were fixed at 75, 110 and 120°C, respectively, and 27.5 mL/min of argon at 723.8 kPa was used as a carrier gas (Paulose and Kaparaju, 2021).

The dissolved solids contents were measured from loss of weight on drying of liquor obtained after centrifuging the substrates at 3,500 rpm for 1 hour in a centrifuge and filtering through a 0.22 μm syringe filter. The total suspended solids (TSS) content was calculated from the total solids and total dissolved solids (TDS) content and by APHA method 2540D from vacuum filtration of a known amount of sample and weight loss on drying. The solids densities of the samples were calculated from the slurry density and solids concentrations, and total dissolved solids and density of the liquor. Slurry and liquor densities were measured using calibrated density cup and pycnometer, respectively.

Shear yield stress and flow curve measurements were conducted using a vane-in-cup (wide gap, cup diameter 30 mm, vane diameter and length 15 mm and 38 mm, respectively) or bob-in-cup (narrow gap, cup diameter 30 mm, bob diameter 28 mm, bob length 44 mm) using a TA instruments' HR3 rheometer. The TH WAS Concentrate, WAO Digestate Concentrate and Untreated WAS and Digestate Concentrate samples were hand mixed and loaded in the cup, the vane or bob immersed in the sample, and kept at equilibrium for 3 minutes. For shear yield stress, an angular velocity of 0.02 rad/s was applied for 3 minutes, and the torque values were recorded. The shear yield stress was calculated from the maximum torque value and vane geometry. For flow curve measurement, an angular velocity ramp (varying in the range 50 rad/s to 0.01 rad/s) was applied and torque values were recorded at 10 seconds intervals until a variation of less than 1% in three consecutive recording was achieved for each angular velocity.

4. Results and discussion

4.1 Chemical composition of WAS Concentrate and Digestate Concentrate with and without pre-treatments

The characteristics of the studied materials with and without pretreatments are presented in Table 8. The untreated whole WAS provided by Sydney Water had a TS content of 6.32% w/w with 86% of TS as VS. The total COD value of untreated whole WAS was 80.28 g/L whilst the soluble COD was 1.3 g/L. Prior to TH, WAS was concentrated using a centrifuge. The untreated WAS Concentrate had a TS and VS content of 9.41% and 8.32%, respectively (Table 9). TH of WAS Concentrate (TH WAS Concentrate) resulted in TS and VS of 9.01 and 8.18%, respectively. However, measured TSS for TH WAS Concentrate was 5.78%, which is significantly lower than the measured TS value of 9.01% in untreated WAS Concentrate. This difference can be attributed to the production of liquid or semi-solid organic compounds during TH pretreatment, which have boiling points higher than 100°C and did not evaporate during drying at 105°C in the oven. TH of WAS Concentrate improved the solubilisation of organic matter as indicated by an improved VS/TS ratio from 0.86 to 0.91. TH pretreatment also improved the organic acids content. The total VFA content increased from 187 mg/L in WAS Concentrate to 2,511 mg/L in TH WAS Concentrate. Acetic acid was the main VFA component and accounted for 48.5% of total VFA.

Digestate collected from the AD plant (untreated Whole Digestate) had TS and VS values of 1.83% and 1.37% w/w, respectively. Prior to WAO, digestate was concentrated into Concentrate (27% w/w) and Centrate (73% w/w) fractions using a centrifuge. Concentration of the digestate resulted in the distribution of chemical characteristics between these two fractions. Phosphorus was mainly present in Digestate Concentrate while ammonium N and potassium were present in the Centrate. The Digestate Concentrate had 7.19% and 5.64% w/w of TS and VS, respectively. The corresponding values for Centrate were 0.2% and 0.13%, respectively. The Digestate Concentrate used for partial WAO had a TS and VS of 10.2% and 7.27%, respectively (VS/TS ratio of 0.71). Partial WAO (20% O_2 loading) of the Digestate Concentrate (WAO 20% Digestate Concentrate) resulted in a decrease in both TS and VS content. Partial WAO also improved the solubilisation of organic matter with VS/TS ratio of 0.71 (Table 10). This was indicated by the increase in total VFA content (Table 9). The TS and VS values for the WAO Digestate Concentrate were 7.55% and 4.58%, respectively (VS/TS ratio of 0.61). Similar to the TH WAS Concentrate, the TSS content of the WAO Digestate Concentrate was 5.22%, which is significantly lower than that of the TS. Production of liquid or semi-solid organics during WAO may be the reason for the difference in the TS and TSS values for the WAO product (Table 8).

Table 8. Chemical composition of the WAS and digestate concentrate with and without pretreatments. Values are in mean ± SD of triplicates (* mg/L)

Sample/Parameter	WAS		AD Digestate			
	Untreated Whole WAS	TH WAS Concentrate	Untreated Whole Digestate	Untreated Digestate Concentrate	Untreated Digestate Centrate	WAO Digestate Concentrate
pH	6.7	6.01	7.21	6.30	7.12	8.75
Total solids, TS (% w/w)	6.32 ± 0.06	9.01 ± 0.04	1.83 ± 0.02	7.19 ± 0.02	0.20 ± 0.00	7.49 ± 0.13
Volatile solids, VS (% w/w)	5.44 ± 0.05	8.18 ± 0.02	1.37 ± 0.01	5.64 ± 0.16	0.13 ± 0.00	5.35 ± 0.05
VS/TS ratio	0.86	0.91	0.75	0.64	0.78	0.71
Total chemical oxygen demand, tCOD (g/L)	80.28	N/A	24.06	N/A	N/A	122.913
Soluble COD (g/L)	1.30	N/A	0.61	N/A	N/A	43.380
Total Carbon, C (% TS)	44.22	30.29	36.47	39.52	24.45	36.67
Total Carbon (g/kg)	11.04	27.29	7.17	28.43	0.24	16.85
Total Nitrogen, N (% TS)	8.53	5.73	5.60	6.12	4.11	6.62
Total Nitrogen (g/kg FM)	2.13	5.17	1.10	4.40	0.04	3.04
C/N ratio	5.18	5.29	6.51	6.45	5.95	
Total Phosphorus (as P g/kg)	1.47	1.63	0.44	1.43	131.50*	1.75
Total Kjeldahl Nitrogen (as N g/kg)	5.12	9.06	1.03	4.52	975.00*	3.41
Nitrite nitrogen, NO_2-N (mg/L)	0.00	1,912.50	0.00	2.45	1.02	0
Nitrate nitrogen, NOx-N (mg/L)	203.09	10,608.00	148.55	181.15	76.5	829.20
Ammonium nitrogen, NH_4-N (mg/L)	130.38	1,848.72	798.66	753.98	794.58	3155.24
Total phosphate, Total PO_4 (mg/L)	335.31	10,254.23	168.30	100.37	121.38	804.364
TVFA (mg/L)	187.59	2511.85	162.70	56.27	115.98	65.73
Acetic acid (mg/L)	100.52	1218.77	136.70	39.95	85.67	14.66
Propionic acid (mg/L)	55.15	504.39	17.10	7.75	11.31	0.00
Iso-Butyric acid (mg/L)	4.62	82.64	0.00	0.00	0.00	24.47
Butyric acid (mg/L)	9.37	321.70	8.90	3.16	6.79	0.00
Iso-Valeric acid (mg/L)	12.50	157.78	0.00	3.35	7.79	26.60
Valeric acid (mg/L)	5.43	133.08	0.00	2.07	4.44	0.00
4-Methyl valeric acid (mg/L)	0.00	80.79	0.00	0.00	0.00	0.00
Hexanoic acid (mg/L)	0.00	12.71	0.00	0.00	0.00	0.00

SD – Standard deviation

4.2 Physio-chemical and Shear rheological characterisation of TH WAS Concentrate and WAO Digestate Concentrate

The shear rheological behaviour of sewage sludges is described by the properties of shear yield stress (τ_y) and viscosity (η). The shear yield stress (τ_y) is defined as the minimum stress needed for the onset of flow; concentrated sludges demonstrate a shear yield stress due to the formation of a continuous particulate network. The shear yield stress is one of the three fitting parameters in the Herschel-Bulkley model, which is the most widely used model to describe the shear rheology of sewage sludge. The other two fitting parameters are: consistency (k), which is an indicator of viscosity (η), and flow behaviour index (n). The shear yield stress (τ_y) can be measured experimentally from the peak stress value during slow constant rate deformation or from the curve fit of the experimental rheograms (shear stress vs shear rate curves).
The Herschel-Bulkley model fitting parameters obtained from experimental flow curves (shear stress versus shear rate) of untreated and pretreated WAS Concentrate and Digestate Concentrate are presented in Table 9 and Table 10. The experimental flow curves (shear stress vs shear rate) and their Herschel-Bulkley model curve fits for different sludge and pretreated sludge samples are presented in Figure 7.

The Herschel-Bulkley model predicted shear yield stress (τy) values for the untreated WAS Concentrate and TH WAS Concentrate were 103.5 Pa and 2.1 Pa, respectively. The flow consistency (k, which is an indication of viscosity) values for untreated and TH WAS Concentrate were 18.8 and 0.75, respectively. As flow consistency is an indicator of apparent viscosity of the sludge samples, a reduction in flow consistency (k) corresponds to a reduction in viscosity as a result of the TH pretreatment. The reduction of shear yield stress and viscosity after TH can be attributed to the decrease in the TSS content (8.35% and 5.78% for untreated WAS and TH WAS, respectively) and the conversion of solids fractions of WAS into liquid and semi-liquid organic compounds in the TH. Reduction in yield stress and flow consistency index improves both mixing and pumping performance and reduces energy requirements of the system.

A similar trend was also observed from the comparison of shear rheological parameters of Digestate Concentrate and partially WAO. The Herschel-Bulkley predicted shear yield stress (τy) values for Digestate Concentrate and WAO Concentrate were 186 Pa and 0.02 Pa, respectively. The flow consistency (k) values for Digestate Concentrate and WAO Concentrate were 3.5 and 0.004, respectively, which is an indication of a dramatic reduction of viscosity (almost close to the viscosity of water) due to WAO. Similar to the trend observed for the TH WAS, WAO Digestate Concentrate also resulted in the decrease of the TSS content (10.11% and 5.22% for Digestate Concentrate feed and WAO Digestate product, respectively) and the conversion of solids fractions into liquid and semi-solid organic compounds, leading to a decrease in shear yield stress (τy) and flow consistency (k as an indicator of viscosity). It was observed a significant change in flow behaviour of WAO Digestate Concentrate (reached almost similar viscosity of water) and 2 orders of magnitude drop of yield stress and viscosity for TH WAS.

Table 9. Shear rheological and physicochemical properties of untreated Whole WAS, WAS Concentrate and TH WAS Concentrate

	WAS		
Properties	Untreated Whole WAS	WAS Concentrate	TH WAS Concentrate
TS (%, w/w)	4.61	9.41	9.19
VS (% w/w)	4.05	8.32	8.13
VS/TS	0.88	0.88	0.88
TSS (%, w/w)	4.45	9.26	5.78
TDS (% w/w)	0.16	0.16	N/A
Solids density, kg/m^3	1429	1429	N/A
Geometry	Wide gap vane	Wide gap vane	Cylindrical bob
Measured shear yield stress (τ_y), Pa	19.27	198	-
Herschel-Bulkley model parameters			
Shear yield stress (τ_y), Pa	11.01	103.52	2.15
Consistency (*k*), Pa.s^n	4.44	18.77	0.75
Flow index (*n*),-	0.38	0.45	0.55

Table 10. Shear rheological and physicochemical properties of untreated Whole Digestate along with WAO Digestate Concentrate at two oxygen loading conditions

Properties	Untreated Whole Digestate	WAO Pretreatment			
		Run 1 44 bar for 20% O_2 content		Run 2 22 bar for 10% O_2 content	
		WAO 20% Digestate Concentrate	WAO 20% Digestate Concentrate	WAO 10% Digestate Concentrate	WAO 10% Digestate Concentrate
TS (%, w/w)	1.61	10.2	7.55	10.6	7.68
VS (%, w/w)	1.11	7.27	4.58	7.6	4.67
VS/TS	0.69	0.71	0.61	0.72	0.61
TSS (%, w/w)	1.51	10.11	5.22	10.5	5.58
VSS (%, w/w)	-	-	-	-	2.87
VSS/TSS	-	-	-	-	0.52
TDS (%, w/w)	0.1	0.1	-	0.1	-
Solids density, kg/m^3	1418	1418	-	1418	-
tCOD (g/L)	24.06	-	-	130.938	122.913
sCOD (g/L)	0.61	-	47.150	2.575	43.380
Geometry	Cylindrical bob	Wide gap vane	Cylindrical bob	Wide gap vane	Cylindrical bob
Measured shear yield stress (τ_y), Pa	-	268	-	-	-
Herschel-Bulkley model parameters					
Shear yield stress (τ_y), Pa	0.006	186	0.023	-	0.22
Consistency (*k*), Pa.s^n	0.22	3.54	0.004	-	0.005
Flow index (*n*),-	0.39	0.69	0.99	-	0.94

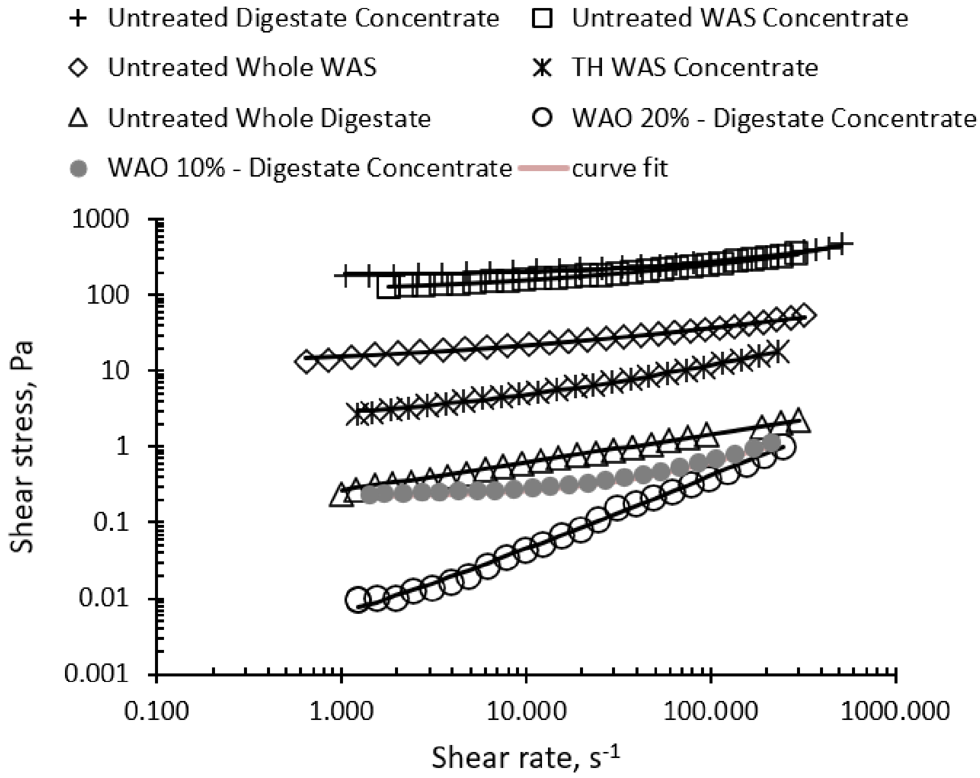


Figure 7. Experimental flow curves and their Herschel-Bulkley curve fits for untreated Digestate Concentrate, untreated waste activated sludge (WAS), untreated Digestate Concentrate, TH WAS and WAO Digestate Concentrate at 10% (WAO 10%) and 20% (WAO 20%) O_2 loading.

4.3 Chemical characterisation of TH and WAO pretreatments on WAS Concentrate and Digestate Concentrate using FTIR

ATR-FTIR spectroscopy was used to qualitatively determine the chemical composition in the surface of samples. The ATR-FTIR complete spectra of untreated whole WAS and digestate are shown in Figure . In general, both samples contained peptides and proteins, carbohydrates, fats, and aliphatic. The untreated Whole WAS spectrum is dominated by the peaks at 3,300 and 1,035 cm^{-1}, the stretching vibrations of O–H of hydroxyl group compounds (polyalcohols and saccharides), and at 3,100 cm^{-1}, the N-H stretching vibrations (proteins, peptides).

The peak at 2,950 cm^{-1} reflects alkyl chains (polyalcohols, saccharides, and fats), referred to the stretching vibration of C–H bonds. The intensity of peaks in this region decrease in the digestate. In the case of WAS, spectra also observed mid intensity peaks on 1,710 cm^{-1}, which reflect stretching C=O vibrations: aliphatic ketones—1,695–1,730 cm^{-1}; aromatic aldehydes—1,690–1,720 cm^{-1}; aliphatic–aromatic ketones—1,675–1,700 cm^{-1}; and amides—1,630–1,700 cm^{-1}. In the region 1,630 cm^{-1}, the mostly intense absorption band in the whole spectra is observed. This signal reflects stretching, asymmetrical vibrations of COO^- and is characteristic for peptides and proteins. This intensity of peaks in the untreated whole digestate had decreased.

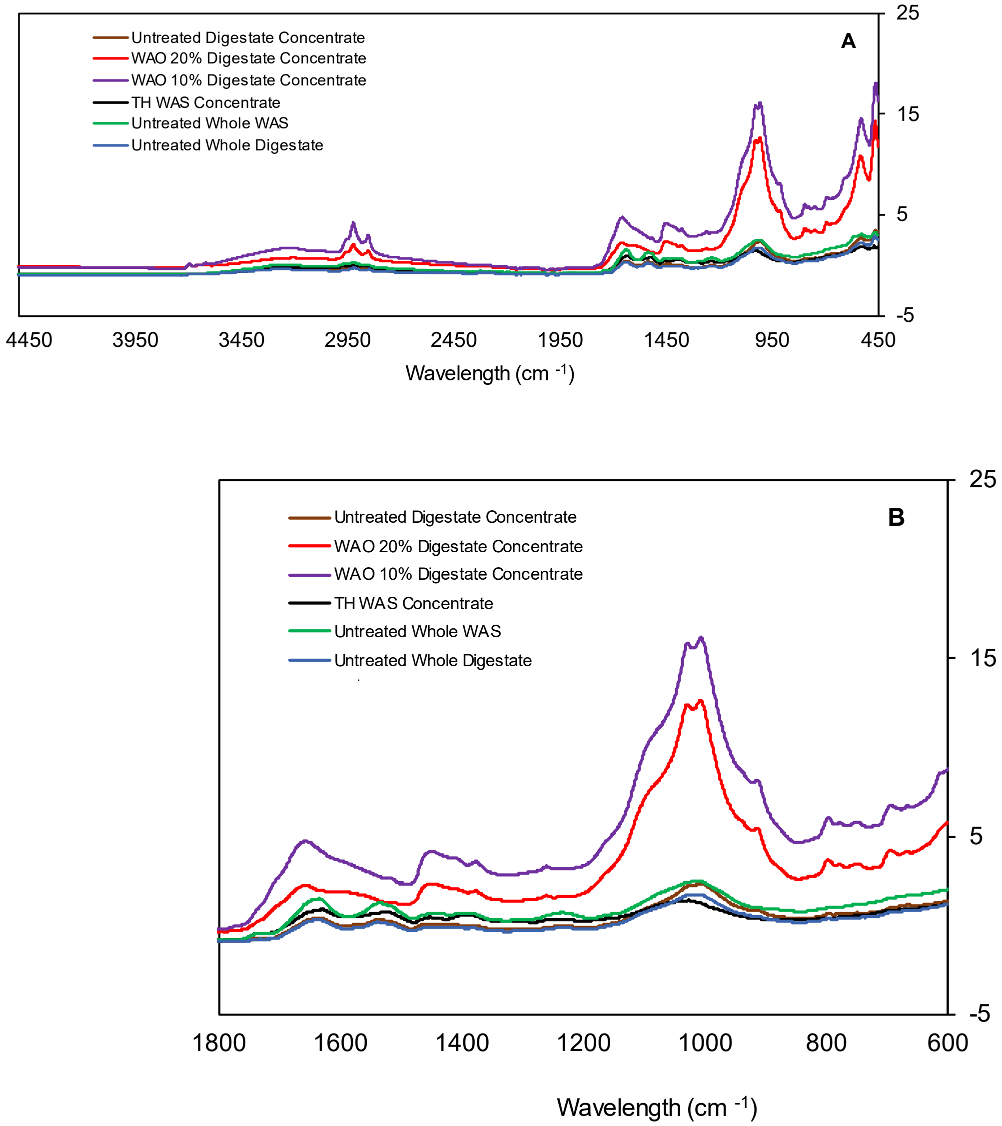


Figure 8. ATF-FTIR spectra of untreated and TH pretreated WAS Concentrate along with untreated and WAO of Digestate Concentrate. A – full wavelength spectrum (450 – 4,450 cm^{-1}), B – partial wavelength spectrum (600-1,800 cm^{-1}).

The next band occurs in 1,515 cm^{-1}. It is a sharp, intense peak reflecting corresponding symmetrical vibrations of NH^{+} bonds, characteristic for proteins. The relative intensity of this band in the digestate sample is decreased. There can also be observed absorption bands near 1,450 cm^{-1}: monosaccharides (deforming C–H bond vibration), ring, aromatic (mid-intense, 1,465–1,430 cm^{-1}), alkenes (C–H functional group), and amides (N–H group). In wave number 1,245 cm^{-1}, the region of deforming vibrations of NH^{+}, an absorption band is also observed. This also reflects peptides and proteins, and the higher intensity in WAS indicates the secretion of the extracellular polymeric substances. The last analysed region is 1,015 cm^{-1}, characteristic for C–O bond stretching vibrations in glucose. Occurrence of this peak is evidence of monosaccharides in the sample.

4.4 Biochemical Methane Potential (BMP)

The cumulative methane production rates and yields for WAS and Digestate with and without pretreatments are presented in Figure 9 and Table 11. Methane production started immediately in all assays except for WAO samples (Figure 8). Higher methane production rates were noticed for TH WAS Concentrate than with untreated Whole WAS or WAS Concentrate. TH of WAS Concentrate (8.5% TS) improved the kinetics of methane production and yields. After 35 days of incubation, methane yields of 378.10 NmL/g VS$_{added}$ was obtained for untreated WAS Concentrate. TH of WAS concentrate improved the methane yields by 10.1% (416.24 NmL/g VS_{added}).

Similarly, methane yields of 317.85 NmL/g VS_{added} were obtained for untreated Whole Digestate (Table 12). Separation of the digestate into concentrate and centrate fractions resulted in methane yields of 189.51 and 288.31 NmL/g VS_{added} for Digestate Concentrate and Digestate Centrate, respectively. WAO of the Digestate Concentrate at 200°C inhibited methane production at both 20% and 10% oxygen loadings. Negligible methane yields of 29.53 NmL/g VS_{added} and 50.35 NmL/g VS_{added} were obtained at 20% and 10% oxygen loading after a lag phase 25 and 12 days, respectively. On the other hand, WAO at 165°C and 20% oxygen loading resulted in methane yields of 117.32 NmL/g VS_{added} (21 d) with a shorter lag phase of 6 days. Chemical analysis showed the presence of 5-hydroxymethylfurfural (5-HMF) (4.7 to 9.97 mg/L) and furfurals (28-30.4 mg/L) in the WAO pretreated samples, Furfural and 5-HMF are furaldehydes and degradation products of pentoses and hexoses, respectively. These furan compounds can inhibit the hydrolytic enzymes as well as the growth and metabolism of the microorganisms used in the subsequent AD process. Furthermore, VFA concentration decreased from 17.06 mg/L in untreated WAS concentrate to 12.13 mg/L after TH. The corresponding values for untreated digestate concentrate before (16.38 mg/L) and after WAO were 15.99 mg/L at 20% and 162.56 mg/L at 10% oxygen loading. It has been noticed from Table 12 that acetic acid was primary VFA produced during the TH. However, the higher concentration of VFAs in WAO at 10% oxygen loading indicate unstable intermediate prior to complete oxidation to carbon dioxide within wet oxidation.

The results thus show that TH was shown to improve the methane yields of WAS Concentrate. On the other hand, WAO of Digestate Concentrate at 10% and 20% oxygen loading did not improve methane yields suggesting further optimisation of the WAO pretreatment in terms of solids loading, pretreatment temperature and/or oxygen concentration is required.

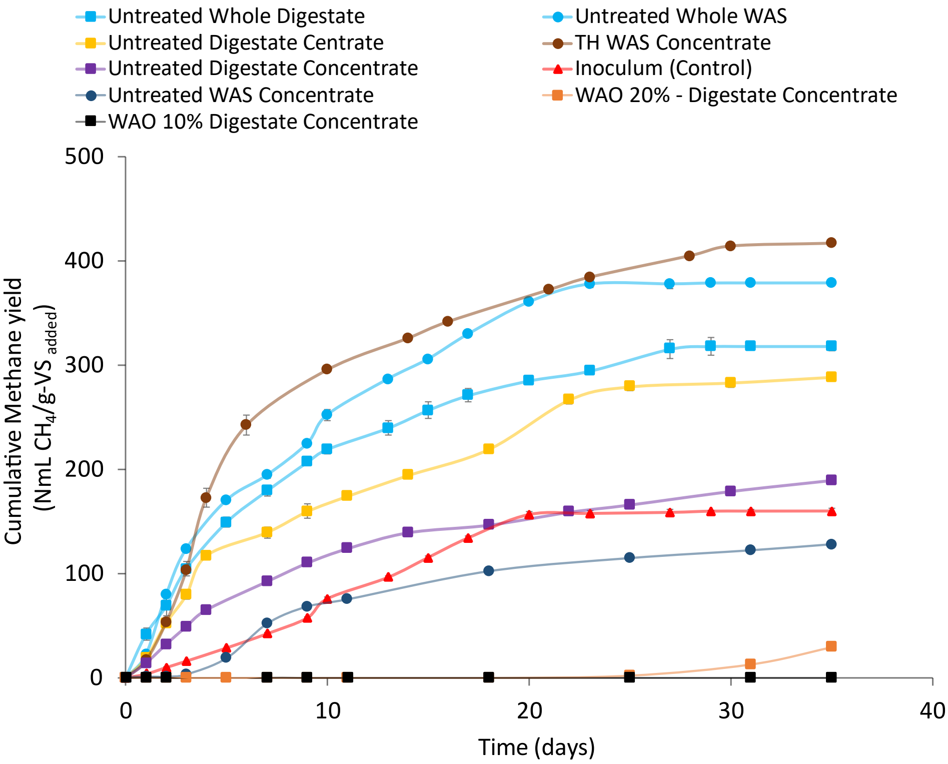


Figure 9. Cumulative methane yields and production rates obtained during anaerobic digestion of untreated and thermal hydrolysis (TH) pretreated WAS Concentrate along with untreated and wet air oxidation (WAO) of Digestate Concentrate at 10% and 20% oxygen loading incubated in batch assays at 37°C.

Table 11. Methane yields obtained during anaerobic digestion of WAS Concentrate before and after TH along with WAO of Digestate Concentrate at 10% and 20% oxygen loading incubated in batch assays at 37°C.

Sample	Methane yields (NmL CH_4/g VS_{added})	Methane yields (Nm^3/t waste)
Untreated Whole WAS	378.10 ± 1.20	20.58
Untreated WAS Concentrate	127.72 ± 0.69	5.00
TH WAS Concentrate	416.24 ± 0.55	34.05
Untreated Whole Digestate	317.85 ± 4.42	4.34
Untreated Digestate Centrate	288.31 ± 4.04	0.37
Untreated Digestate Concentrate	189.51 ± 4.04	10.70
WAO of Digestate Concentrate (@20% O_2 loading, 200⁰C)	29.53 ± 3.41	0.99
WAO of Digestate Concentrate (@10% O_2 loading, 200⁰C)	50.35± 8.21	2.69
WAO of Digestate Concentrate (@20% O_2 loading, 165⁰C)	117.32± 2.80	8.93

Table 12. Chemical composition of digestates at the end of BMP testing. Values are in mean ± SD of triplicates.

Sample/Parameter	WAS			Digestate				
	Untreated Whole WAS	Untreated WAS Concentrate	TH-WAS Concentrate	Untreated Whole Digestate	Digestate		WAO-Digestate Concentrate 20%	WAO-Digestate Concentrate 10%
					Untreated Digestate Concentrate	Untreated Digestate Centrate		
pH	7.71	7.62	7.66	7.64	7.63	7.72	7.62	7.57
TS (%)	2.50 ± 0.09	3.55± 0.03	2.57 ± 0.02	1.97 ± 0.13	4.01 ± 0.03	1.09 ± 0.02	2.11	2.05
VS (%)	1.74 ± 0.01	2.46 ± 0.02	1.81 ± 0.01	1.36 ± 0.02	2.86 ± 0.06	0.82 ± 0.01	1.46	1.40
Total Carbon (% TS)	36.52	33.51	36.60	34.29	32.08	36.46	32.83	NA
Total Carbon (g/kg Fresh)	9.12	7.07	9.42	6.74	3.49	14.62	6.93	NA
Total Nitrogen (% TS)	6.11	6.02	6.20	5.53	5.64	5.46	6.02	NA
Total Nitrogen (g/kg Fresh)	1.53	1.27	1.60	1.09	0.61	2.19	0.94	NA
Total Phosphorus (as P g/kg)	0.84	1.13	0.33	0.62	0.99	0.40	1.47	NA
Total Kjeldahl Nitrogen (as N g/kg)	1.45	2.00	0.45	1.17	1.99	0.64	2.87	NA
NO_2-N (mg/L)	0.00	0	0.00	0.00	0.00	0.00	0	NA
NOx-N (mg/L)	131.62	130.56	133.73	131.69	51.83	52.34	133.23	NA
NH_4-N (mg/L)	1907.40	79.69	2152.20	1468.80	1285.20	1815.60	66.96	NA
Total PO_4 (mg/L)	399.84	26.06	395.76	257.04	221.34	262.14	18.08	NA
TVFA (mg/L)	16.20	17.06	12.13	16.40	16.38	16.60	15.99	162.06
Acetic acid (mg/L)	10.30	11.94	7.92	10.20	11.40	10.26	11.28	125.34
Propionic acid (mg/L)	1.50	0	0	1.80	0	1.57	0	9.81
Iso-Butyric acid (mg/L)	0.00	0	0	0.00	0	0	0	0
Butyric acid (mg/L)	2.60	2.79	2.63	2.40	2.27	2.24	2.63	3.45
Iso-Valeric acid (mg/L)	0.00	0.57	0	0.00	0	0	0	0
Valeric acid (mg/L)	1.80	1.76	1.58	2.00	2.27	1.96	1.58	1.71
4-Methyl valeric acid (mg/L)	0.00	0	0	0.00	0	0	0	21.75
Hexanoic acid (mg/L)	0.00	0	0	0.00	0	0.57	0	0

5. Technical and economic feasibility of TH and partial WAO of sludge for biogas production

5.1 Introduction to feasibility study

Sydney Water has identified two potential treatments for improving the overall net energy from their biogas plants. At St Mary's wastewater treatment facility, the effect of TH on methane yields and net energy from WAS was evaluated. Similarly, the effect of partial WAO (using a lower oxygen concentration than that needed for complete oxidation and therefore attaining incomplete WAO) on digestate from AD plants treating WAS was also evaluated. In both studies, the amount of biosolids after WAO treatment and the costs associated with the transport of the biosolids for final disposal were compared with their respective business as usual scenarios.

To assess the feasibility of the proposed project, different scenarios were designed with an aim to evaluate the possible market opportunities for reducing the amount of biosolids to dispose and also an overall improvement in the net energy production. The studied scenarios will help in the process of decision-making. Based on the pretreatment options, four different scenarios were considered in this feasibility study and are presented below.

Scenario 1 – Business as usual (BAU) AD of primary sludge (PS) and waste activated sludge (WAS), and separation of digestate from the AD into solid (biosolids or Digestate Concentrate) and liquid (Centrate) fractions. Biosolids are exported to 300 km for final disposal, whilst the liquid fraction is recycled to the WWTP. The produced biogas is used for cogeneration in a combined heat and power (CHP) plant with electrical efficiency of 35%, noting that this CHP plant did not consider acid gas treatment while performing cost estimation. Renewable electricity with a value of $0.085/kWh is consumed behind the meter.

Scenario 2 – Similar to Scenario 1 but thickened WAS is subjected to TH pretreatment.

Scenario 3 – Similar to Scenario 1 but solid fraction of anaerobic digestate (Digestate Concentrate) is subjected to full WAO (100% oxygen loading) at 240°C. The liquid fraction (Centrate) from digestate solid-liquid separation is NOT used as process water for AD plant.

Scenario 4 - Similar to Scenario 1 but the solid fraction of anaerobic digestate (Digestate Concentrate) is subjected to partial WAO pretreatment at 240°C and 20% oxygen loading (Partial WAO Digestate Thickened). The liquid fraction (Centrate) from digestate solid-liquid separation is used as process water and recycled to AD plant.

It should be noted that Scenario 3 differs from the experimental work done in this study (Section 4). In Scenario 3, WAO is performed at 100% oxygen loading (Full WAO). For Scenario 4, based on laboratory tests, WAO is performed at 20% oxygen loading (Partial WAO) as the experimental results for partial WAO at 10% oxygen loading did not show any significant difference to the results obtained at 20% oxygen loading.

5.2 Drivers for this feasibility study

Annually, Sydney Water treats 500 GL of sewage and produces 180,000 tonnes of biosolids from its AD facilities, the generated biosolids are trucked inland and reused as compost or soil conditioner. However, the long trucking distance of around 300 km incur transport costs for biosolids disposal ($100/t) and significant greenhouse gas (GHG) emissions. Reduction in the volume of biosolids will have a direct impact on trucking costs and GHG emissions.

Sydney Water is constructing a TH pretreatment plant to reduce biosolids quantity and improve biosolids quality. However, current thermal pretreatment technologies, such as the TH process, are energy intensive and result in only a small positive net energy generation. Therefore, this project assessed the energy balance of TH of WAS with full and Partial WAO treatment of Digestate Concentrate obtained from AD of WAS. Both treatments were compared with the BAU scenario in terms of net energy balance, the reduction in biosolids amounts, the associated costs and GHG emissions from disposal.

5.3 Methodology for this feasibility study

For the feasibility study, techno-economic feasibility for the four scenarios mentioned in Section 5.1 are compared and analysed. For each scenario, a process flow diagram was developed using the process description details obtained from the literature and with the inputs from Sydney Water. A WAS feed flow rate of 86,400 kg/day was used for the mass balance calculations. The operating conditions of various process and equipment size were obtained from the literature and Sydney Water. Process simulation models for all scenarios were built using Aspen Plus software to solve mass and energy balance calculations. The stoichiometric reaction equations and reaction kinetic models required for simulating AD plant, TH unit, and WAO unit were obtained from the literature.

The process stream and operating condition results were used to size all the equipment in the process flow diagram for each scenario. The equipment sizes and operating conditions were used to estimate the plant's Glenfield capital cost for use in each scenario. The operating costs for each scenario were estimated using the raw materials consumption rate (if there are any), energy consumption rate, and waste disposal rate as variable costs and labour costs, and fixed capital dependent costs as fixed costs. The working capital (inventory fund required to run the WWTP) for each scenario was also calculated using the details available in the operating cost estimation tables. The fixed capital investment, operating cost and working capital were then used to construct a cash flow table for each scenario. The income in the cash flow table was calculated using the revenue generated from the sale of electricity produced from methane gas, Australian Carbon Credit Unit (ACCUs) and green certificates. All the data used to estimate the revenue is shown in Table 13. Electrical conversion efficiency of 35% was used for converting methane to electricity and heat (Suhartini, Lestari, & Nurika, 2019). The cash flow analysis was used to determine the profitability indicators like payback period and return on investment (ROI). A discounted cash flow analysis was also performed to obtain the net present value (NPV) and internal rate of return (IRR). The operating life of the plant was assumed to be 25 years for all scenarios in the cash flow analysis.

Table 13. Revenue Estimation Values

Term	Value	Unit
Methane value in electricity	9.68	kWh/m^3 CH_4
Methane value in heat	0.038	GJ/m^3 CH_4
1 tonne of coal	2.65	MWh
Greenhouse intensity of 1 MWh electricity from coal	0.986	tonne of CO_2/MWh
Carbon credits value	30	\$/tonne CO_2
Biomethane electricity sale price	0.085	\$/kWh
Green certificate	3	\$/GJ

5.4 Process descriptions and modelling

The following sections briefly describe the work done on different scenarios. Using the process simulation models developed in Aspen Plus, mass and energy balance models were constructed in Excel spreadsheet (Microsoft®). Both process simulation models and Excel spreadsheets are connected with the remaining techno-economic analysis (please refer to Appendix F). Every spreadsheet file has seven worksheets. The first file (worksheet 1.1) shows the compounds that are included in the calculations of TS, VS, inert material (ashes) and moisture content. Worksheet 1.2 presents the results obtained from the Aspen Plus simulation as a process flow diagram (PFD) form, while worksheet 1.3 estimates the methane production under different sludge feed rates. Worksheets 1.4, 1.5, 1.6 and 1.7 presents the economic analyses for the process flow diagrams, including capital cost, operation cost, working capital estimations, and cash flow analysis, respectively.

5.4.1 Process flow diagram

5.4.1.1 Process flow diagram for anaerobic digestion of primary sludge (PS) and waste activated sludge (WAS) mixture, and dewatering/concentration of anaerobic digestate (Scenario 1, BAU)

The process flow diagram for Scenario 1 is shown in Figure 10. In this scenario, a mixture of WAS and PS is fed to the anaerobic digester at 25°C and 1 bar. The biogas produced in the digester is then sent to CHP/cogeneration for heat and electricity generation. On the other hand, the digestate is sent to a dewatering unit to separate solids from the liquid phase. The solid phase (Concentrate) is then sent for disposal, while the liquid phase (Centrate) is recycled to the WWTP as process water for further treatment.

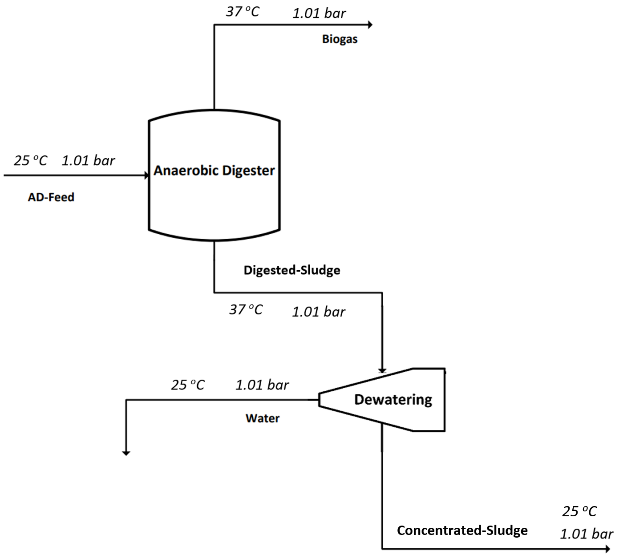


Figure 8. Process flow diagram for anaerobic digestion (AD) of waste activated sludge (WAS) and primary sludge (PS) followed by dewatering and concentration of the digestate (Scenario 1).

5.4.1.2 Process flow diagram for anaerobic digestion of thermal hydrolysis (TH) of waste activated sludge (WAS) Thickened (Scenario 2)

The process flow diagram for Scenario 2 is presented in Figure 11. In Scenario 2, The preheated WAS Thickened is fed to the TH unit which solubilises a considerable amount of solids in the sludge. The TH WAS Thickened is then fed to the anaerobic digester. The biogas produced in the anaerobic digester is sent to the CHP/cogeneration unit for heat and electricity generation. On the other hand, digestate from the anaerobic digester is dewatered to separate the solids from the liquid. The digestate concentrate is sent further for processing, while the centrate or liquid is recycled to the WWTP as process water.

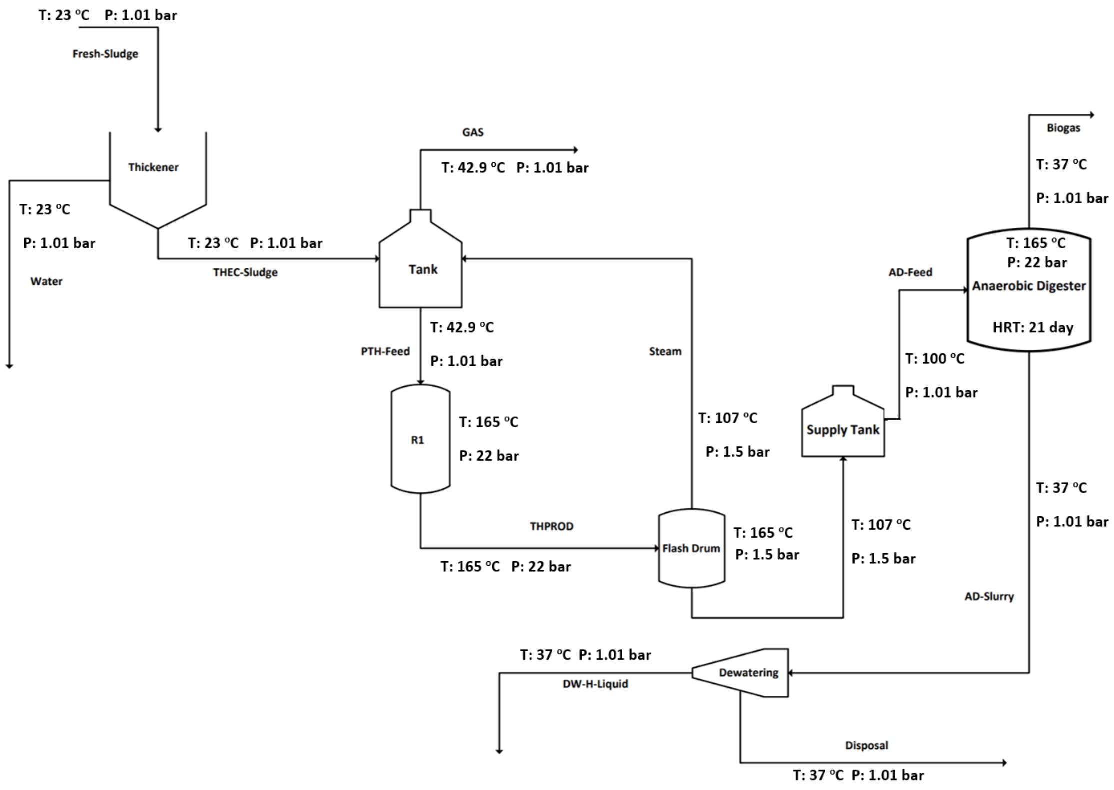


Figure 9. Process flow diagram for anaerobic digestion (AD) of thermal hydrolysis (TH) of waste activated sludge (WAS) thickened (Scenario 2).

5.4.1.3 Process flow diagram for anaerobic digestion (AD) of primary sludge (PS) and waste activated sludge (WAS) and wet air oxidation (WAO) of AD digestate concentrate (Scenarios 3 and 4)

Two different WAO treatment of AD digestate scenarios are shown in Figure 12: full WAO (100% oxygen loading) without recycling the liquid stream as process water for AD (Scenario 3) and partial WAO (20% oxygen loading) with recycle of liquid stream as process water for AD (Scenario 4). The WAO unit shown in the process flow diagram in Figure 12 will be used either for full WAO (Scenario 3) or partial WAO (Scenario 4). The feed WAS is pumped into the anaerobic digester under ambient conditions. The anaerobic digester operates under mesophilic temperature (37°C) to degrade the WAS. In the case of full WAO, the hydraulic retention time (HRT) of 21 days was used for AD. The corresponding value for partial WAO is 14 days as the recycled liquid stream is mixed with the incoming digester feed, leading to a lower HRT. Biogas produced during the AD is sent to the CHP cogeneration unit. On the other hand, digested sludge is sent to a thickening unit where it is thickened to contain 13 wt.% solids. The thickened sludge stream (Digestate Thickened) is then sent to a steam heater to prepare it for WAO.

During the WAO, temperature of Digestate Thickened is increased from 37°C to 160°C and operated at a pressure of 88 bar. In Aspen Model, WAO unit is designed as a gravity pressure vessel (GPV). The GPV is comprised of a long steel pipe, shaped like a test tube, of a fixed diameter. The annulus of an open-ended steel pipe creates updraft and is suspended within the test tube. This updraft protrudes above the test tube and descends to within a few feet of its concave bottom. Small bore steel pipes are suspended in the updraft to inject steam and chemicals, for temperature control, cathodic protection and cleaning. WAO treated Digestate Thickened is allowed to flow through the inner pipe downwards for 750 m.

Oxygen stream (100% purity) is fed at 15 bar into the WAO unit through a central pipe located inside the inner pipe. The reactions between oxygen and sludge at the bottom section of the WAO unit increase the sludge temperature up to about 240°C. Reacted sludge rises in the outer pipe and is removed as the product at the ground level. There is only one product stream from the WAO unit which contains gas and liquid products. A flash unit separates the gas and liquid phases after the WAO unit. In addition to separation, the flash unit also reduces the product pressure to 1 bar. The gas stream is sent to power generation while the liquid phase is recycled to the anaerobic digester.

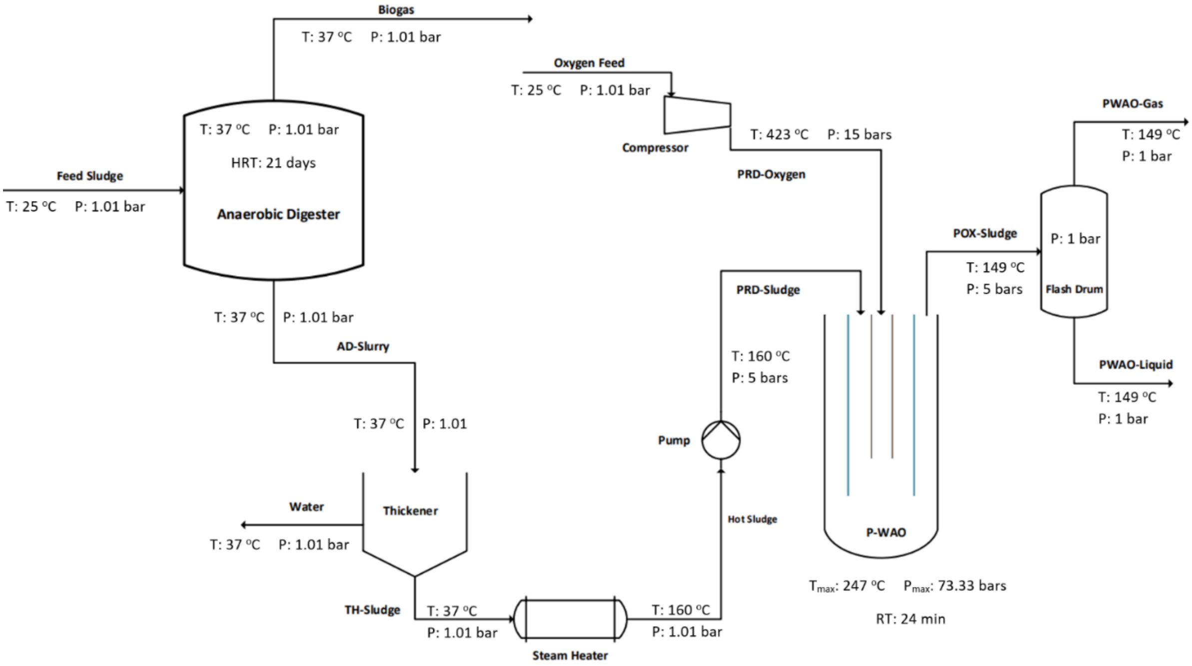


Figure 10. Process flow diagram for wet air oxidation of anaerobic digestate concentrate (Scenario 3/4).

5.4.2 Mass and energy balance calculations and economic analysis

Based on this project's aims, three main process options were studied: AD, WAO and TH. The four scenarios considered for this work are:

Scenario 1 – AD (original plant)
Scenario 2 – TH pretreatment of WAS followed by AD of the Digestate Concentrate
Scenario 3 – AD followed by 100% WAO of AD digestate thickened in a GPV unit with no recycle stream.
Scenario 4 – AD followed by partial WAO (20% oxygen) of AD digestate thickened in a GPV with liquid stream from partial WAO outlet recycled as process water to AD plant.

These scenarios helped in studying three process alternatives for the existing AD plants of Sydney Water: 1) addition of TH (Scenario 2), 2) addition of full WAO (Scenario 3) in a GPV unit, and 3) addition of partial WAO in a GPV unit with recycling liquid stream from partial WAO to the AD plant (Scenario 4).

Using Aspen Plus software, these scenarios were simulated to carry out the mass flow and energy consumption calculations. The capital costs of process plants for each scenario were estimated based on the process flow diagram and the results of the mass and energy balance calculations. Cash flow analysis was then carried out after determining the operating cost, working capital and income due to revenue sources i.e., electricity, ACCUs and green certificates. The economic feasibility indicators, like net present value (NPV) and internal rate of return (IRR), were used to recommend the best process alternative.

The WAO process is simulated based on the concept of the process rather than the equipment itself. Due to the lack of reported studies, assumptions were made on chemical reactions to complete the process simulation modelling. Pure oxygen is used in the WAO process at a ratio of 0.2 tonnes per tonne of dry sludge to achieve a partially oxidised environment for partial WAO (20% oxygen loading) and 1.0 tonnes per tonne of sludge for full WAO process (100% oxygen loading). For the TH process, CAMBI technology was selected. However, there is a knowledge gap in the literature regarding reaction kinetics and stoichiometry. Therefore, the kinetics and stoichiometry for the hydrolysis phase of the AD were used in the TH simulation. It is a valid assumption as the literature suggests that some of the reactions of the AD hydrolysis phase are similar to those in TH.

In all studied scenarios, the outlet pressure of pumps that transports the sludge was assumed to be 3.5 bar. The heat from units such as thickeners and dewatering units is assumed to be conserved. Depending on the feed stream temperature, the outlet stream temperature for the equipment, like the thickener or dewatering unit, was estimated. Hence, inlet temperature equals to the outlet temperature for all these units.

The economic analysis was conducted by estimating the equipment and plant cost using the data from literature and the CAPCOST spreadsheet available in a chemical engineering design textbook (Turton, Bailie, Whiting, & Shaeiwitz, 2008). The fixed capital cost included the Sydney Water's Glenfield AD plant and CHP. Operating cost was analysed based on manufacturing and non-manufacturing costs. Working capital was included to evaluate the inventory fund available to meet the plant operation. Finally, the plant revenue was estimated using two sources of income: ACCUs, green certificates and sales of electricity from biogas as shown in Table 13. The revenue and operating costs were used in a cash flow analysis to determine profitability indicators like ROI, payback period and IRR. For more information regarding the calculation of the economic analysis, please refer to Appendix F.

Due to the lack of reliable and accurate data, the following assumptions were made in conducting the mass and energy balance calculations in various scenarios.

- The experimental work in this project did not identify the compounds and components of the sludge feed. Therefore, the fresh feed in all scenarios is assumed to be WAS whose composition was obtained from the literature (Rajendran et al. 2014).
- Fresh sludge in all scenarios is assumed to be fed at 23°C.
- Salt formed was assumed to be among "Inert/Ashes."
- It was assumed that GPV does not lose heat to the surroundings, i.e., energy conserved in the system.
- Due to the lack of reaction kinetic data and models for partial WAO, data for complete oxidation were used in the process simulation.
- It is assumed that the infrastructure cost includes piping, electrical and construction costs.
- Based on the advice given by Sydney Water, it was assumed that the disposal cost is $100 AUD per tonne of sludge.
- It was assumed that the plant life is 25 years for all the scenarios.

5.4.3 Process modelling using Aspen Plus

5.4.3.1 AD model

A process model was developed using Aspen Plus to simulate the four reaction stages during the AD process. They are hydrolysis, acidogenesis, acetogenesis and methanogenesis (Rajendran et al., 2014). The model allows the user to input the sludge compositions and digester operating conditions like temperature, pressure, and HRT. The model results provide an estimation of the gas and liquid products and their concentrations. Supplying a valid feed composition is important for the simulation to estimate reliable product composition. Therefore, it is essential to identify the commonly found and dominant components in the feed and product streams and include them in the simulation. Table 14 presents the components found in the feed and product streams of the anaerobic digester and their categories. In contrast to experimental work and industry practice, there is no definition of WAS in Aspen Plus as feed is characterised based on fat, carbohydrate, protein and volatile fatty acids (VFA) contents. To differentiate between the different type of sludges, feed composition based on Table 16 is required. Because of the project time limitation, the experimental work did not cover the analysis of the sludge fed to the anaerobic digestion (i.e. WAS). Meaning that the compositions of WAS that provided the concentrations of starch (carbohydrate), tripalmitin (fat) or protein were not determined in the experiment work. However, these details are needed as input for Aspen Plus simulation. Therefore, the Aspen Plus simulation results cannot be comparable to the experimental findings. In this study, the digester feed composition reported in the literature (Rajendran et al. 2014) was used in the Aspen Plus simulation. In future work, experimental work should include such analysis to generate the data required for Aspen Plus simulation. Upon provision of different quantities of the listed components, the Aspen Plus model can be simply updated for a new feed stream composition.

Table 14. Components present in the feed and product streams of the anaerobic digester.

Category	Components in the system
Carbohydrates	Cellulose, Hemicellulose, Starch and Xylose, Glucose, Dextrose
Fat	Triolein, Tripalmitin, Palmito-olein and Palmito-linolein
Protein	Soluble and Insoluble Proteins
Volatile Fatty Acids (VFAs)	Acetic acid, Propionic acid, Butyric acid, Valeric acid, Palmitic acid and Linoleic acid
Amino Acids	Asparagine, Glutamine, Arginine, Histidine, Lysine, Tyrosine, Tryptophan, Phenylalanine, Cysteine, Methionine, Threonine, Serine, Leucine, Isoleucine, Valine, Glutamic acid, Aspartic acid, Glycine, Alanine and Proline
Others (gases, ester, acid-base, Inert)	Ammonia, methane, carbon dioxide, hydrogen, benzene, phenol, furfural, hydrogen sulphide, methyl mercaptan, formamide, indole, carbonic acid, Inert and ethyl cyanoacetate.

In the Aspen Plus modelling, the chemical and physical properties of the main components listed in Table 14 were considered. The anaerobic digester operating conditions used in the simulation were 37°C and 1 atm with an HRT of 21 days. Two reaction blocks, RStoic and RCSTR, were used in Aspen Plus to simulate the reactions in the anaerobic digester (Figure 13). RStoic estimates the reaction products and their concentrations based on reaction conversion, and it was used to model the hydrolysis phase of the digestion. RCSTR estimates the conversion using reaction kinetics, and it was used to model the reactions in acidogenesis, acetogenesis and methanogenesis.

The model mentioned above also considers the inhibition caused by butyric acid, ammonia, and pH. Although two reaction blocks are used in the Aspen Plus simulation, they simulate the process that occurs in one piece of equipment, an anaerobic digester. Many previous studies have validated this approach by comparing the simulation results with experimental results. As mentioned above, the gas product is used for power generation, and the liquid product is treated in the dewatering unit for removing solids for disposal or in the WAO process for further destruction of organics in the sludge.

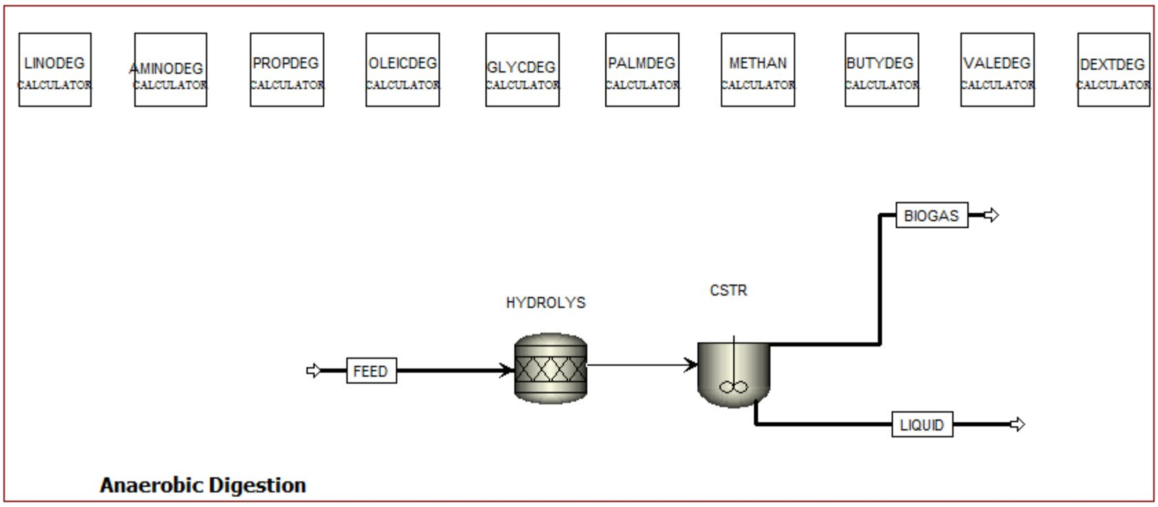


Figure 13. AD simulation using Aspen Plus (Rajendran et al., 2014)

5.4.3.2 Wet Air Oxidation Model

A model for the WAO process was developed to simulate the process that occurs in a GPV. A GPV is an underground piece of equipment reactor consisting of three pipes: oxygen pipe, sludge inlet pipe and sludge outlet pipe (Figure 14). The equipment can be extended to a great depth, depending on the application. There is a significant gap in the literature regarding GPV technology. Most papers found are patented and related to bioethanol formation processes. The main purpose of this technology is to convert complex carbohydrates to simpler sugar molecules using hydrostatic pressure, temperature, and low pH conditions. For this report, the RMIT team developed a novel simulation model to evaluate the products expected in the process outlet stream. The simulation considered the equipment's dimensions, fluid dynamics, and heat transfer. Figure 14 shows a schematic diagram of the GPV, including an illustration of material flow in and out of the equipment. Sludge enters the inner vertical pipe at the top of the GPV while oxygen flows through a tube placed inside the inner pipe. As suggested by Sydney Water, the depth of the GPV was set at 750 m. The sludge and oxygen experience a pressure change of about 73 bars at this depth. Both feed sludge and oxygen are pumped/compressed downwards to prevent backflow and allow the oxidised sludge to move upwards in the outer pipe.

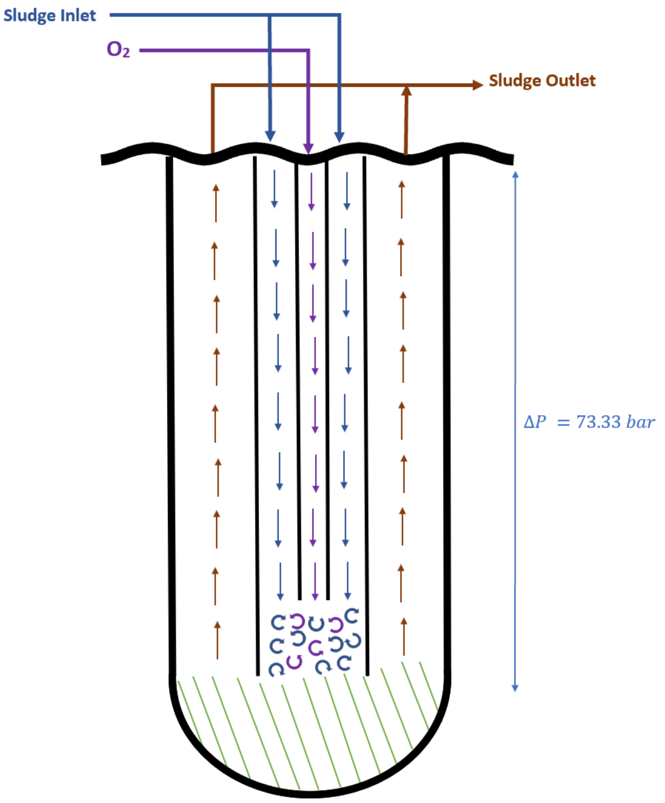


Figure 14. Sludge flow in gravity pressure vessel (GPV)

The developed simulation model used for WAO is shown in Figure 14. Both Figure 14 and 15 are presented with different colours to provide a better understanding of the fluid flow directions. Here, the blue line refers to the Digestate Thickened sludge that enters the equipment, while the purple colour shows the oxygen path downwards. The brown colour indicates the reacted WAO Digestate Thickened sludge that moves upwards to exit the GPV. The bottom of the equipment is highlighted with a green colour to show the area where the sludge is reacting. It is important to point out that the sludge starts reacting once it is mixed with oxygen (i.e. mixing point above the green area). In the model, sludge is heated by high-pressure steam to about 160°C and pumped downwards at 88 bars. Oxygen is compressed to 15 bars, raising the gas temperature to approximately 423°C, and sent to the reactor. The residence times of the sludge in the downward and upward pipes were estimated using chosen liquid hold-up models in Aspen Plus.

The WAO reactions in the GPV were simulated using a plug flow reactor (PFR) module available in Aspen Plus. This PFR module simulates the reaction in the process more accurately since the reactions occur in a pipe-shaped reactor. It also can simulate the sludge flow appropriately in both the downward and upward directions.

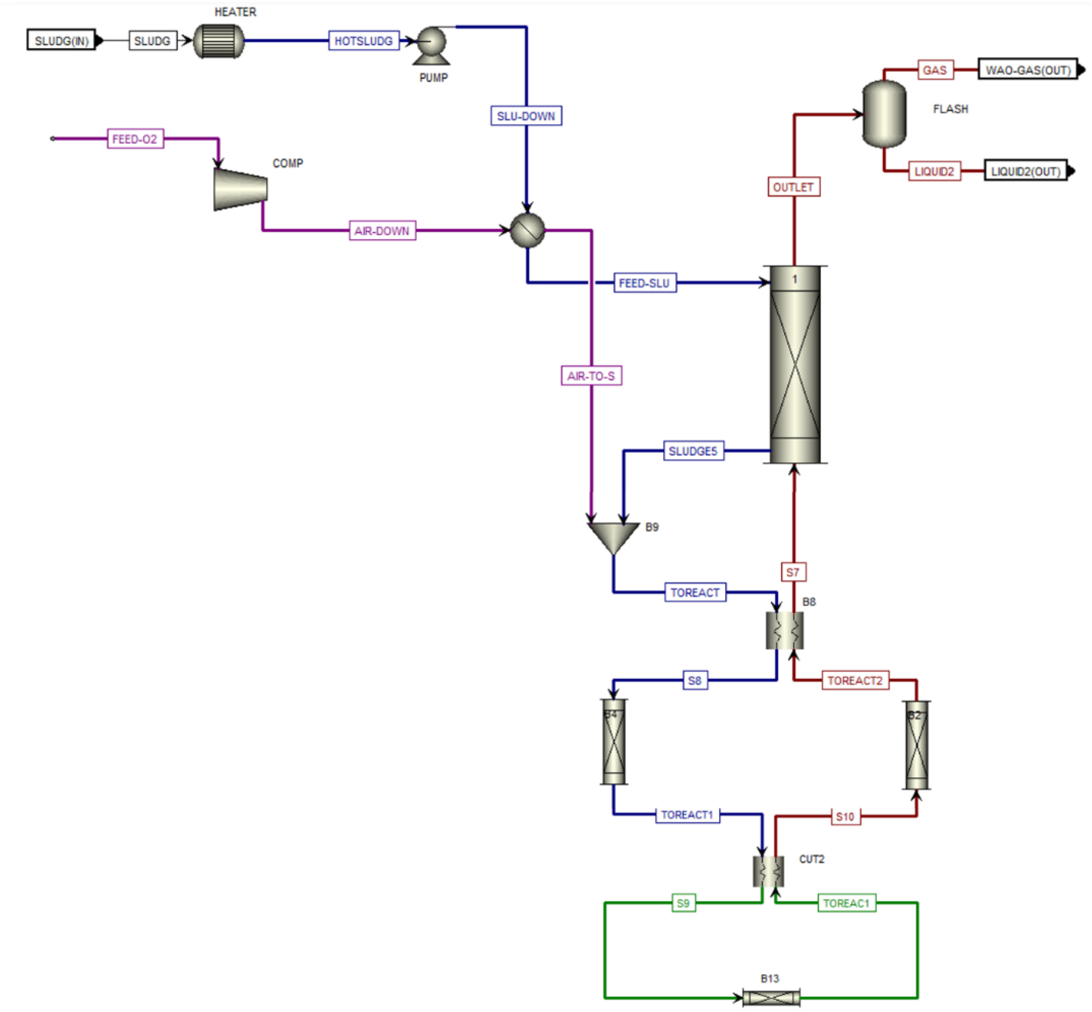


Figure 15. GPV process simulation model in Aspen Plus

The reactor dimensions and heat transfer between the pipes were considered in the simulation, as shown in Figure 16. Although Sydney Water provided the length of the GPV, other dimensions were assumed for this simulation. In the GPV, heat is transferred to the feed sludge (160°C) while it is flowing downwards in two ways, (1) from oxygen flowing in the centre of the pipe at 423°C, which ensures co-current heat transfer, and (2) from the reacted sludge that moves upward at approximately 240°C, which ensures counter-current heat transfer.

Minimal information is available in the literature regarding the reactions that occur in the partial WAO process. However, a few studies reported it to be a reaction involving only one component: glucose. For the simulation in this work, reactions involving glucose and dextrose were considered, and reaction kinetics reported by Park et al. (2016). Other reactions, such as the degradation of cellulose, hemicellulose, and proteins, were simulated using data reported for full WAO in the literature. Furthermore, this work considers some of the reactions related to VFAs, resulting in carbon dioxide and water production. All the reactions used in the WAO simulation are presented in Appendix A.

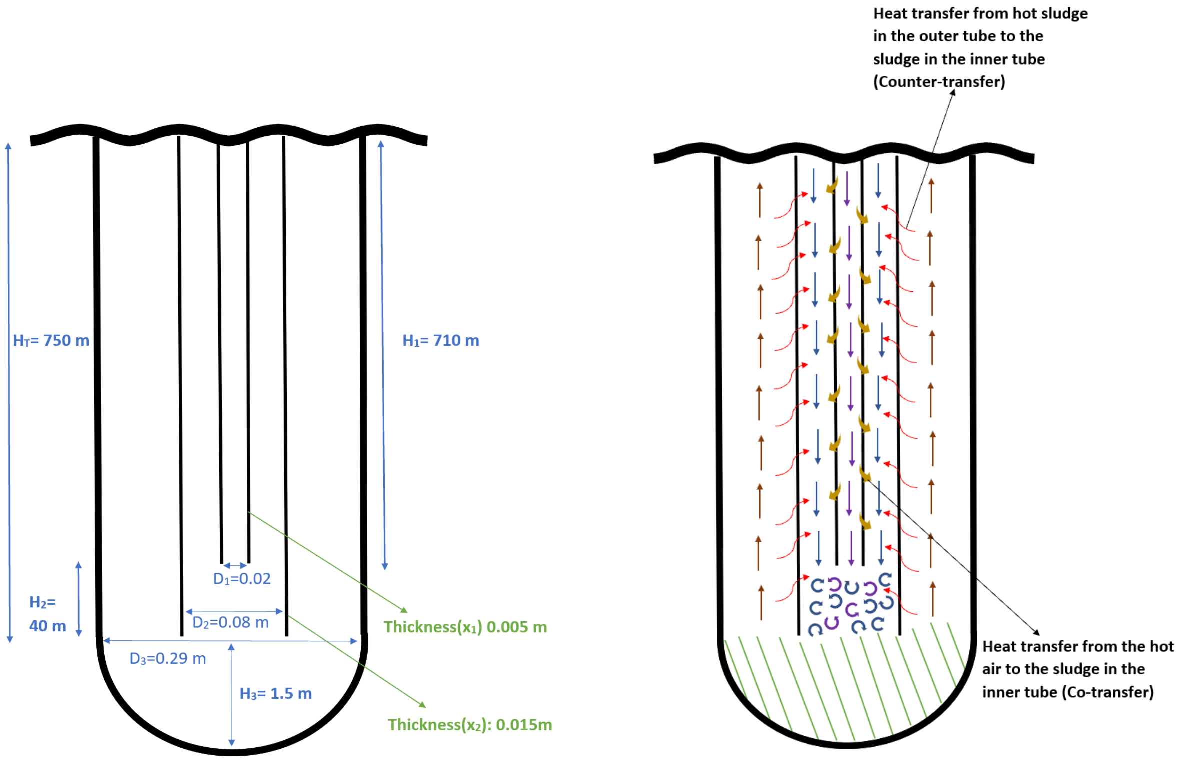


Figure 16. GPV dimensions and heat transfer considered for Aspen Plus simulation, purple = oxygen, blue = concentrated DS, brown = WAO-DS sludge

Further work was done to incorporate acid hydrolysis into the simulation of the GPV technology. A study was conducted on furan formation (Furfural and HMF) when acid treatment is used in the GPV technology. In acid hydrolysis, cellulose is converted to glucose, while hemicellulose is converted to xylose. With the addition of acid, both glucose and xylose led to the production of furans. The kinetics used in studying the above reactions were obtained from a study on a continuous pipe reactor under high pressure and temperature (Franzidis, Porteous, & Anderson, 1982; Köchermann, Mühlenberg, & Klemm, 2018). The formation reaction of glucose, xylose and furans is shown in Appendix B. Sydney Water advised that sulphuric acid of 1 to 2 wt% concentration was used for treating the reacted sludge. However, no information was available regarding the depth at which the acid would be injected. Injection depth is crucial as it influences the residence time of the reaction mixture and, therefore, the extent of the reaction. Therefore, an additional study was conducted to find the optimal injection point (injection depth).

5.4.4 Preliminary Aspen Plus Simulation Results

5.4.4.1 AD simulation results

The composition of feed was used in the simulation is shown in Table 15. The flow rate of the feed was 86,400 kg/day (1 kg/s) with a digester operating condition of 37°C, 1 atm, and HRT of 21 days. The simulation results for the liquid and gas product streams are shown in Table 16 and Table 17, respectively. The mass percentage of methane in the gas product stream is 62 wt%, close to the 60 wt% reported in industrial operations. This finding confirms the validity of the reaction blocks, reactions, and kinetics used in the simulation.

Table 15. Feed stream composition used in the simulation

FEED Stream	Composition (wt.%)
Water	94.00
Dextrose	1.00
NH_3	0.10
Cellulose	2.20
Hemicellulose	1.00
Triolein	0.04
Tripalmitin	0.04
Sn-1-Palmito-2-Linolein	0.04
Soluble Protein	0.30
Insoluble Protein	0.18
Inert	1.10
Summary	
FEED Stream	**Composition (wt.%)**
Water	94.00
NH3	0.10
Carbohydrate	4.20
Protein	0.48
Fats	0.12
Inert	1.10

Table 16. Liquid product stream composition

Liquid Stream	Mass (kg/day)	wt.%	Mole (kmole/day)	mol.%
WATER	80995.59	94.45	4495.94	98.28
CH_4	158.19	0.18	9.86	0.22
CO_2	1120.40	1.31	25.46	0.56
VFAs	684.51	0.80	8.56	0.19
NH_3	108.26	0.13	6.36	0.14
Amino acids	0.06	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00
H_2S	19.98	0.02	0.59	0.01
Benzene	0.63	0.00	0.01	0.00
Protein	103.68	0.12	0.74	0.02
Carbohydrates	1014.25	1.18	6.79	0.15
Fats	10.37	0.01	0.01	0.00
ethyl cyanoacetate	157.33	0.18	1.39	0.03
Ethanol	429.75	0.50	9.33	0.20
Glycerol	4.11	0.00	0.04	0.00
Inert	950.40	1.11	9.50	0.21
O_2	0.00	0.00	0.00	0.00
Other	0.00	0.00	0.00	0.00
Total	85757.52	100.00	4574.58	100.00

Table 17. Gas product stream composition

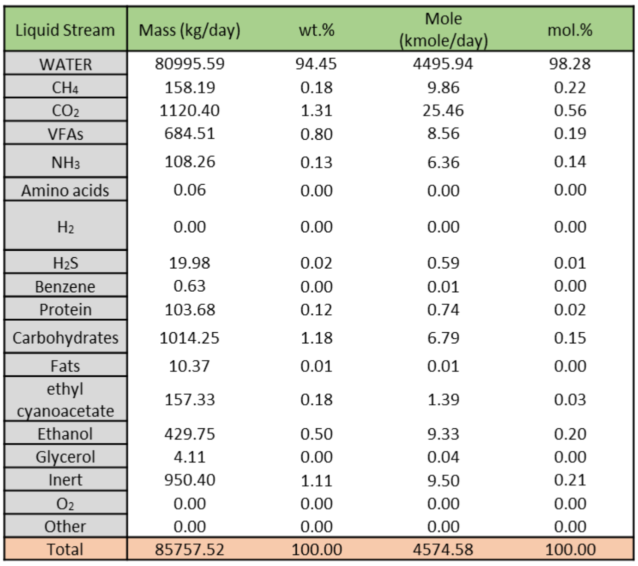
Gas Stream	Mass (kg/day)	wt.%	Mole (kmole/day)	mol.%
WATER	33.59	5.46	1.86	6.17
CH_4	381.30	62.04	23.77	78.65
CO_2	191.97	31.23	4.36	14.43
VFAs	0.20	0.03	0.00	0.01
NH_3	0.78	0.13	0.05	0.15
Amino acids	0.00	0.00	0.00	0.00
H_2	0.04	0.01	0.02	0.06
H_2S	3.19	0.52	0.09	0.31
Benzene	1.24	0.20	0.02	0.05
Protein	0.00	0.00	0.00	0.00
Carbohydrates	0.00	0.00	0.00	0.00
Fats	0.00	0.00	0.00	0.00
ethyl cyanoacetate	0.00	0.00	0.00	0.00
Ethanol	2.30	0.37	0.05	0.17
Glycerol	0.00	0.00	0.00	0.00
Inert	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00
Other	0.00	0.00	0.00	0.00
Total	614.61	100.00	30.22	100.00

5.4.4.2 WAO simulation results

The digestate from the AD is concentrated by removing 13% of its water and then sent to the WAO unit for oxidation. The digestate thickened temperature was 37°C and the oxygen temperature was 25°C. Table 18 shows the compositions of the digestate thickened stream. Partial WAO (20% oxygen loading) was implemented in the simulation. Table 19 and Table 20 show the compositions of the liquid and gas product streams, respectively, from the partial WAO unit. It can be seen that some of the carbohydrates and a small amount of proteins degraded, resulting in the formation of CO_2. Due to the exothermic reaction, the maximum temperature attained by the sludge was approximately 240°C. It was obtained from the area at which reacted sludge starts to move upwards in the equipment as shown in Figure 14 or the green stream (TO-REACT1) in Figure 15. The simulation shows that sludge spent about 24 minutes between the inlet and outlet. The outlet stream temperature was 149°C, and the stream was then transported to the flash drum for phase separation. Similar results and stream conditions can be expected for full WAO. Furthermore, it was found that inhibitors were formed after the partial WAO, where compounds such as furfural and 5-hydroxymethylfurfural (5-HMF) were detected in the liquid outlet stream.

The acid hydrolysis study (Figure 17) on partial WAO showed that hemicellulose is readily converted to xylose. The xylose is converted to furfural when sulphuric acid is introduced close to the surface. The injection of 1 wt% acid/sludge at 50 m below the surface will eventually lead to the consumption of most of the xylose in the sludge. On the other hand, the cellulose is hard to convert, and therefore, a higher residence time is needed (i.e., higher acid injection depth). Due to the project timeline, after consultation with Sydney Water, an injection depth of 20 m and an acid concentration of 1 wt% were selected for this study.

Table 18. Compositions of the PWAO feed stream



Liquid Stream	Mass (kg/day)	wt.%	Mole (kmole/day)	mol.%
WATER	80995.59	94.45	4495.94	98.28
CH_4	158.19	0.18	9.86	0.22
CO_2	1120.40	1.31	25.46	0.56
VFAs	684.51	0.80	8.56	0.19
NH_3	108.26	0.13	6.36	0.14
Amino acids	0.06	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00
H_2S	19.98	0.02	0.59	0.01
Benzene	0.63	0.00	0.01	0.00
Protein	103.68	0.12	0.74	0.02
Carbohydrates	1014.25	1.18	6.79	0.15
Fats	10.37	0.01	0.01	0.00
ethyl cyanoacetate	157.33	0.18	1.39	0.03
Ethanol	429.75	0.50	9.33	0.20
Glycerol	4.11	0.00	0.04	0.00
Inert	950.40	1.11	9.50	0.21
O_2	0.00	0.00	0.00	0.00
Other	0.00	0.00	0.00	0.00
Total	85757.52	100.00	4574.58	100.00

Table 19. Liquid products from PWAO

Liquid Stream	Mass (kg/day)	wt.%	Mole (kmole/day)	mol.%
WATER	21385.80	82.95	1187.09	95.30
CH_4	7.45	0.03	0.46	0.04
CO_2	657.41	2.55	14.94	1.20
VFAs	671.89	2.61	8.07	0.65
NH_3	28.17	0.11	1.65	0.13
Amino acids	0.08	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00
H_2S	4.10	0.02	0.12	0.01
Benzene	0.21	0.00	0.00	0.00
Protein	117.08	0.45	0.87	0.07
Carbohydrates	1183.18	4.59	7.58	0.61
Fats	6.28	0.02	0.01	0.00
ethyl cyanoacetate	0.00	0.00	0.00	0.00
Ethanol	122.00	0.47	2.65	0.21
Glycerol	6.67	0.03	0.07	0.01
Inert	875.06	3.39	8.73	0.70
O_2	0.00	0.00	0.00	0.00
Other	715.48	2.78	13.42	1.08
Total	25780.86	100.00	1245.67	100.00

Table 20. Gas products from PWAO

Gas Stream	Mass (kg/day)	wt.%	Mole (kmole/day)	mol.%
WATER	31.66	5.54	1.76	11.52
CH_4	30.75	5.38	1.92	12.56
CO_2	504.67	88.27	11.47	75.15
VFAs	0.26	0.05	0.00	0.02
NH_3	0.64	0.11	0.04	0.25
Amino acids	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00
H_2S	0.97	0.17	0.03	0.19
Benzene	0.51	0.09	0.01	0.04
Protein	0.00	0.00	0.00	0.00
Carbohydrates	0.00	0.00	0.00	0.00
Fats	0.00	0.00	0.00	0.00
ethyl cyanoacetate	0.00	0.00	0.00	0.00
Ethanol	1.56	0.27	0.03	0.22
Glycerol	0.00	0.00	0.00	0.00
Inert	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00
Other	0.70	0.12	0.01	0.05
Total	571.72	100.00	15.26	100.00

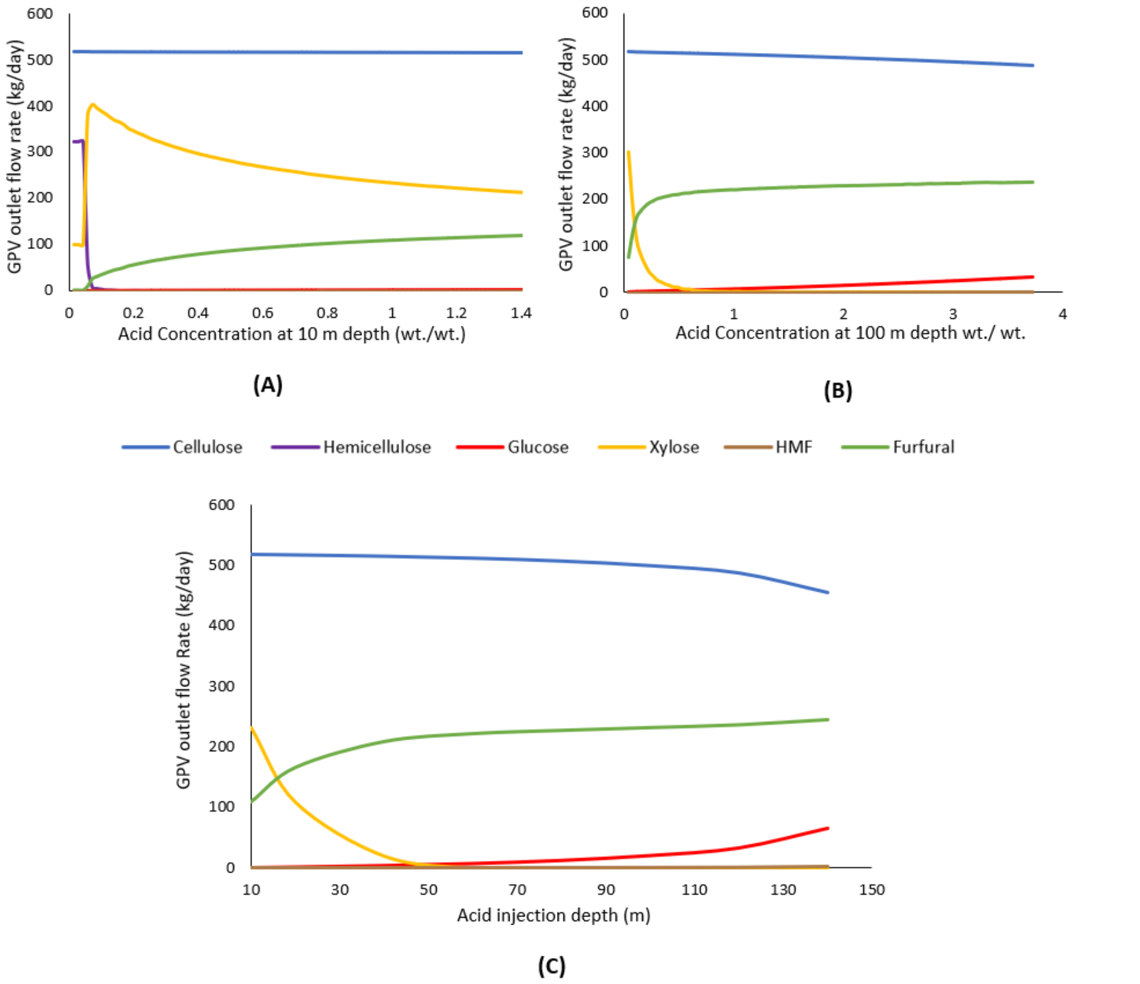


Figure 17. Acid hydrolysis study results: (A) acid concentration effect at an injection depth of 10 m below the surface, (B) acid concentration effect at an injection depth of 100 m below the surface, (C) acid injection depth effect for an acid concentration of 1 wt%.

5.5. Mass and Energy balance calculations for all scenarios

5.5.1 Scenario 1 - (BAU) Anaerobic Digestion (AD)

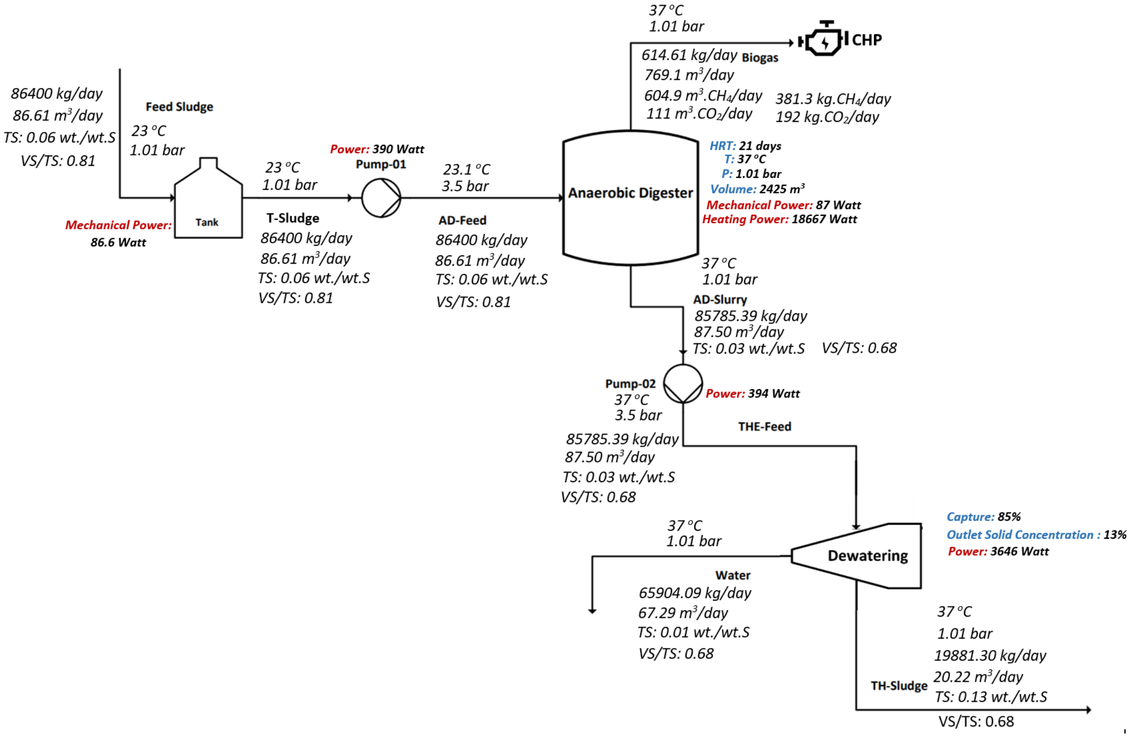


Figure 18. Process flow diagram for anaerobic digestion (AD) plant (Scenario 1).

Scenario 1 is the typical AD plant without any pre- or post-treatment processes. This scenario is considered a base scenario, which will be used for comparison. Figure 18 presents the process flow diagram for Scenario 1. Feed WAS with a flow rate of 86.4 tonne per day is pumped to feed tank and then fed to the AD reactor. The digestate from AD is then pumped to a dewatering unit which separates 85% of the TS. The biosolids are then trucked for disposal. The results of mass and energy balance calculations are shown in the Figure 18. The WAS feed rate to the AD is 84.6 tonnes/day and it contains 94 wt.% liquid and 6 wt.% total solids. The AD process generates about 605 m^3/day of methane while consuming approximately heating energy of 18.7 kJ/s. The digested sludge from AD is sent to a dewatering unit, generating waste with 13 wt.% solids.

5.5.2 Scenario 2 - Anaerobic Digestion (AD) and Thermal Hydrolysis (TH) processes

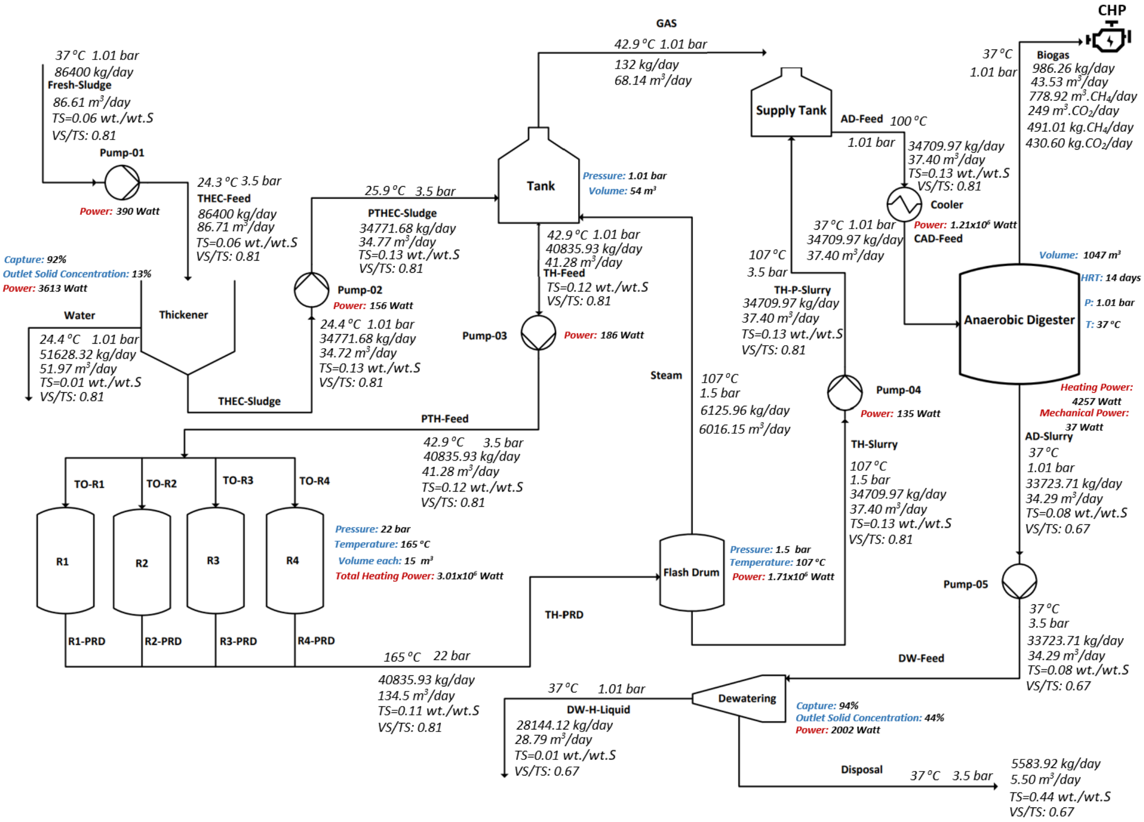


Figure 19. Process Flow Diagram - Thermal hydrolysis of WAS and anaerobic digestion processes (Scenario 2).

In Scenario 2, a TH pretreatment process is added to Scenario 1. The WAS feed with a flow rate of 86.6 tonne per day is firstly thickened to contain 13 wt% TS and sent to a preheating supply-tank. The WAS Thickened in this supply tank is preheated using the hot gas, recycled from the TH flash drum. Some solids in the WAS evaporate due to the high temperature, leading to a decrease in TS concentration by 1%. Four TH reactors are operated in parallel for treating the WAS Thickened. Each TH reactor volume is estimated to be around 15 m³. After the TH step, the product stream consisting of sludge with solubilised solids, is recombined and sent to a flash drum for phase separation. The water vapour or steam is recycled to the TH supply tank while the liquid stream is sent to the AD supply tank. It is important to know that the steam from the flash separator could contain CO_2 and other gases. Hence, not all the gases in this stream will dissolve in the sludge in the TH supply tank. Therefore, the TH supply tank has a vent gas stream. A cooling step is used before the AD to adjust the temperature of the liquid-gas sludge mixture. The AD volume is estimated by the required HRT while considering extra volume for safety consideration. Methane is generated from the AD step along with CO_2. The liquid sludge from the AD is then dewatered to separate the solids. Finally, the dewatering process captures approximately 94% of the total solids while consuming power of around 2 kWh per hour.

5.5.3 Scenario 3 - AD then 100% WAO-DS process with no recycle stream

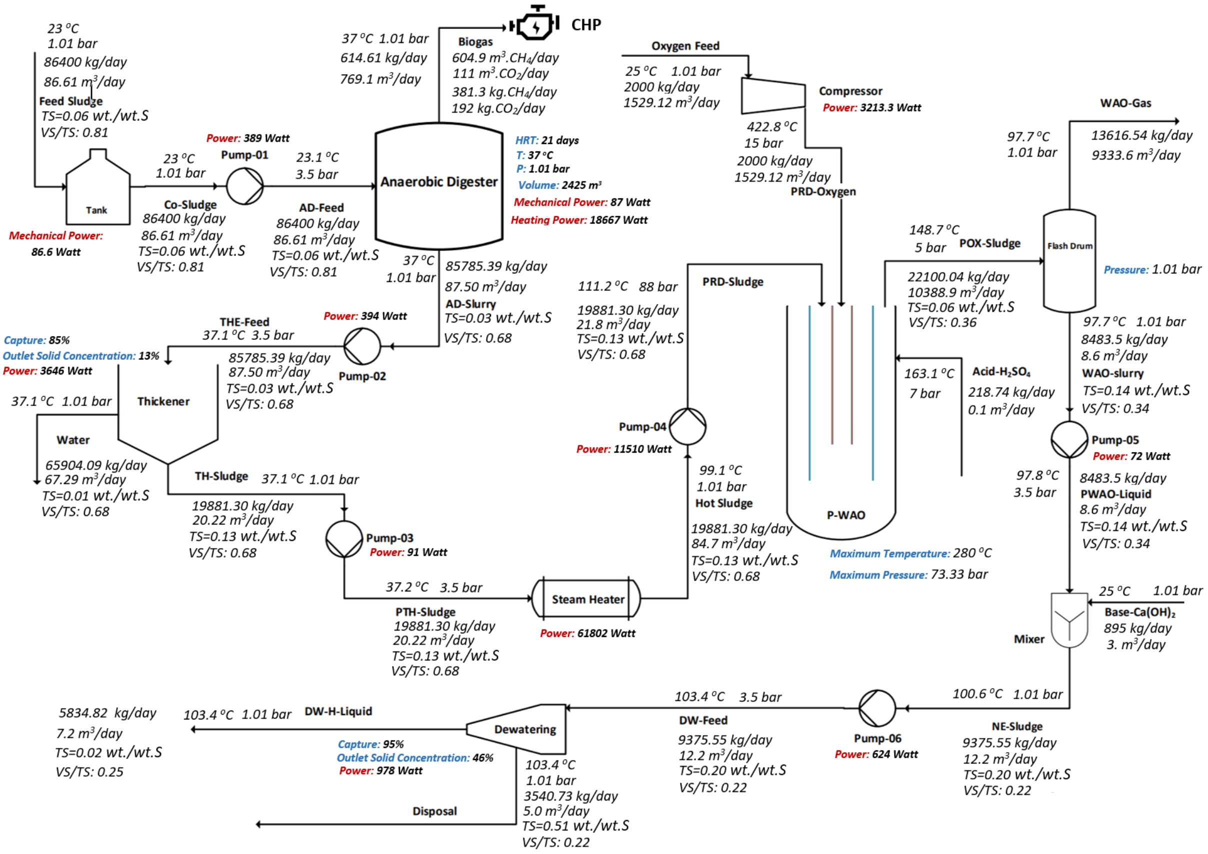


Figure 20. PFD-AD and 100% WAO-DS with no recycle stream (Scenario 3).

In Scenario 3, full WAO of the digestate obtained from AD plant (Scenario 1) is examined. In this scenario, the liquid product obtained from the dewatering process following a WAO is not recycled but instead is directed to disposal. This scenario was considered to estimate the cost of adding full WAO to the original AD plant. The Feed WAS is pumped to the supply tank and fed to AD plant. The AD digestate is then fed to the thickener to reduce its liquid content. The Digestate Thickened stream from the thickener is heated first and then pumped using a cavitation pump into the WAO unit. Excess oxygen is also fed into the WAO unit to ensure complete oxidation. The sludge subjected to the WAO treatment leaves the unit with a low solid concentration. The partially oxidised sludge from WAO is then mixed with calcium hydroxide to neutralise the percarbonic acid, which otherwise would lead to a pH between 4.5-5. In the next step, partial WAO treated sludge is dewatered to capture 95% of the solids. The results of the process's mass and energy balance calculations are shown in Figure 20. The sludge feed rate used in this scenario was 86.4 tonnes per day.

5.5.4 Scenario 4 - Anaerobic digestion and PWAO plant (20% WAO) with recycle stream to main AD

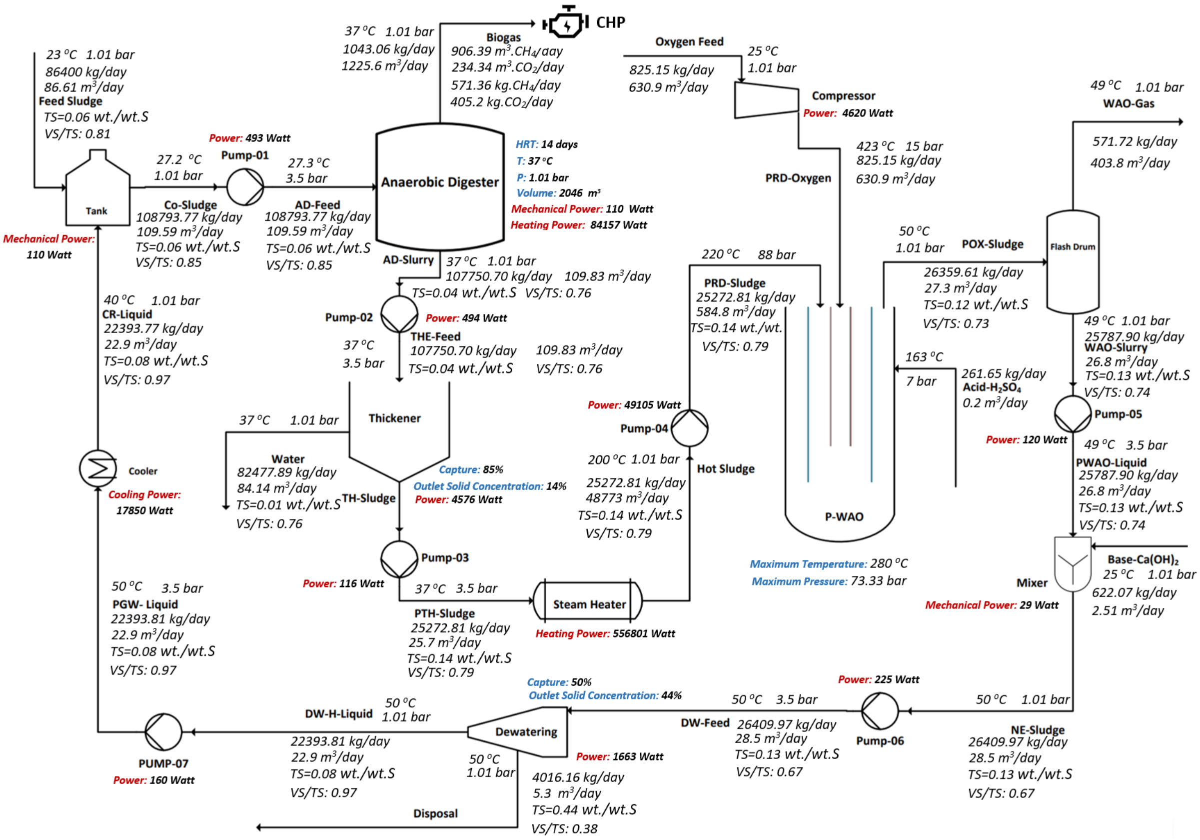


Figure 21. Process Flow Diagram - Anaerobic digestion and 20% partial wet air oxidation (PWAO) with recycle stream to main AD (Scenario 4).

In Scenario 4, partial WAO with recycling the liquid stream obtained from the separation of partial WAO product using a dewatering system are added to the original AD plant. Unlike Scenario 3, HRT for the AD in Scenario 4 is decreased to adjust to the increased feed sludge volume from the recycle stream. Thus, HRT of the AD plant in Scenario 4 is similar to that of Scenario 3. In addition, there are some ancillary equipment used in this scenario that are need for recycling the process water. The total feed fed to the AD plant consists of 80 wt% fresh WAS and 20% recycled process water from the partial WAO process. Furthermore, the dewatering step in the recycle stream captured 50% of the input solids. It is important to understand that these factors significantly impact the factors like the energy consumption by the cooler and pumps, as well as the volume of methane generated in AD. Mass and energy results for this scenario are presented in Figure 21. The fresh feed sludge was 86.4 tonnes per day, similar to those for all scenarios in this study.

Table 21. Mass and energy summary for all four scenarios

Mass balance	Units	Scenario 1 BAU (AD)	Scenario 2 (TH + AD)	Scenario 3 (AD + Full WAO)	Scenario 4 (AD + Partial WAO)
Total feedstock treated	t/d	86.4	86.4	86.4	86.4
Additional water/ recycled process water	kL/d	-	-	-	22.9
Biogas produced	m³/d	769	1,108	769	1,226
Methane produced	m³/d	605	779	605	906
Energy in methane produced	GJ/d	23	29.6	23	34.4
Electricity generated	kWh/d	5,856.4	7,540.7	5,856.4	8,770.1
	GJ/d	21.1	27.1	21.1	31.6
Heat generated	kWh/d	6,388.9	8,222.2	6,388.9	8,777.8
	GJ/d	23	29.6	23	34.4
Carbon dioxide produced from digester	t/d	0.192	0.43	0.192	0.405
Carbon dioxide produced from pre/post-treatment	t/d	-	0.085	2.15	0.505
Parasitic demand—electrical	kWh/d	110	161	490	1,484
Parasitic demand—heat	kWh/d	448	116,284	1,931	16,990
Import of electricity	kWh/d	-	-	-	-
Import of heat (natural gas)	kWh/d	-	-	-	-
Exportable electricity—grid	kWh/d	-	-	-	-
Exportable heat	kWh/d	-	-	-	-
Digestate production	t/d	85.7	33.7	85.7	107.7
Solid digestate	t/d	2.9	2.6	2.9	4.1
Liquid digestate	t/d	82.8	31.1	82.8	103.6

6. Financial Analysis

6.1 Capital Costs

Capital costs of equipment in the process flow diagrams were estimated using the exponential method and CAPCOST spreadsheet tool available from Turton et al. (2018). The estimations used the information obtained through the sizing of major and minor equipment in the process flow diagram and CHP/cogeneration units. The above estimations provided the total capital expenditure (CapEx total), which is the sum of all equipment purchased and installation costs. In addition, development and commissioning, EPSM, infrastructure, and Glenfield fees which represents 10, 15, 10, 17% of total CapEx, respectively, were calculated and added to total CapEx and contingency (inherent 22% and contingent 28% risks) to obtain the total capital investment (TCI). In all estimations, the plant was assumed to be located in NSW and a Glenfield development. The worksheets showing the capital cost estimations are in Appendix F for all four scenarios. The summary of the capital cost components and TCI values for all four scenarios are shown in Table 22. The TCI values for scenarios 1, 2, 3 and 4 are $49.1, $86.1, $127.3 and $139.2 million, respectively. The TCI value for scenario 1 (BAU case) is the lowest. The TCI value for scenario 4 (AD then 20% WAO with recycle stream) is the highest due to additional equipment like GPV, heaters, pumps, coolers and compressors. The addition of recycle stream has increased the number of equipment and capacities, resulting in different TCI for scenario 3 (AD then 100% WAO).

6.2 Operating cost

The operating cost (OpEx) estimation summarises the cost of processing sludge through various operations like anaerobic digestion, thermal hydrolysis, and wet air oxidation in the treatment plant. It also includes the cost of transporting the biosolids from the treatment plant to the storage locations. A $100/tonne rate was used for biosolids transportation and storage. The OpEx estimation also includes operating and management staff salaries and overheads, plant maintenance, insurance and taxes, and plant depreciation. The annual depreciation was calculated by taking the plant operating life as 25 years. The annual OpEx estimated for the four scenarios are shown in the Appendix F. The summary of the OpEx for all four scenarios is shown in Table 22. The annual OpEx for scenarios 1, 2, 3 and 4 are $10.02, $23.3, $18.96, and $21.69 million, respectively. The annual OpEx for scenario 1 is the lowest whereas the annual OpEx for other scenarios are higher due to the additional utility consumptions in the additional process equipment like TH and WAO units, and additional fluid moving equipment, heat exchangers and phase separators. The annual OpEx for scenario 2 (TH then AD) is the highest due to the relatively higher amount of heating involved in the TH process. The main difference between scenarios 3 and 4 is the additional materials fed to the system by the recycle stream in scenario 4. Thus, higher energy is needed to transport and heat the materials, and higher pump power for GPV.

6.3 Working capital

Working capital for a typical processing plant is defined as the additional capital required to maintain the plant's operation over a short period until it starts earning an income (Sinnott & Towler 2013). The components included in the working capital estimation are usually raw material stockpile, finished product stockpile, material in progress, debtors, creditors and wages. Working capital is considered a relatively low-risk investment and fully recoverable at the end of the plant life. In the current study, working capital is the additional capital reserve that is required to maintain the WWTPs until a steady operation is achieved. Not all the working capital components mentioned above will apply fully to the working capital estimation for the wastewater treatment plant. The working capital estimations for the four scenarios are shown in the Appendix F.

6.4 Cash flow analysis

The cash flow analysis was carried out by constructing a cash flow table that shows the calculation of annual cash flow for each year of the plant's operating life (25 years). In addition to the TCI, working capital, annual operating costs, and the annual revenue were included in the analysis. The sources of revenue included are green certificates, Australian Carbon Credit Units and electricity supplied to the grid. A straight-line method was used for calculating annual depreciation, which was then used to calculate taxable income. The cash flow diagram was plotted using the cumulative cash flow values and the number of years. Usually, the cash flow diagram is used to determine the payback period for a typical process plant that produces an income-generating product. The payback period is the time taken for the cumulative cash flow equal to the sum of TCI and working capital. In the current work, no payback period was determined regardless of the scenarios under consideration. The main reason is that a WWTP plant does not produce enough income to generate profits. Its primary income sources are only those mentioned above. The calculation of return on investment (ROI) is also not applicable to this work for the reason mentioned above. Therefore, ROI and payback period values are not included in Table 22.

6.5 NPV and IRR

Two other essential profitability indicators from the cash flow analysis are net present value (NPV) and internal rate of return (IRR). NPV is the sum of future cash flows in today's dollars and is calculated using a nominal rate of return called the 'nominal discount rate'. A higher NPV value for a project will make the project attractive for investment by a business. The annual cash flow values calculated in the cash flow table were used to calculate the annual discounted cash flow or present value using a nominal discount rate of 12% in this work. The sum of the annual discounted values gave the NPV.

IRR is the nominal discount rate that makes the NPV zero. In other words, IRR shows the actual rate of return provided by the project's cash flows. The IRR is compared with a company's hurdle rate (rate of return expected by a company on its investment) to determine the financial attractiveness of the project. NPV is a measure of cash profit, while IRR is a measure of the efficiency of capital utilisation. In this work, IRR was calculated by determining the nominal discount that made the NPV zero. The Goal-seek tool in MS Excel was used to determine the IRR values.

The cash flow tables used to calculate the NPV and IRR for four scenarios are shown in Appendix F. The summary of NPV and IRR values for all four scenarios is shown in Table 22. The NPV values for scenarios 1, 2, 3 and 4 are -$78.5, -$164.5, -$169.1 and -$191 million, respectively. All NPV values are negative due to very low incomes in all four scenarios compared to the higher capital investment and operating costs. The financial attractiveness of the project can be found by determining the scenario with the lowest negative NPV. Scenario 1 has the lowest negative NPV of -$78.5 million. Among the other three scenarios that include sludge pretreatments, scenario 2, which includes TH then AD, has the next lowest negative NPV, ascribed to the second lowest TCI, and second highest ACCUs and electricity sales of about $0.08 and $0.23 million per year, respectively.

The IRR values for scenarios 2, 3 and 4 are 93%, 85%, and 88%, respectively (Table 22). The IRR value for scenario 1 is not included because it was higher than 100%, which was considered unrealistic. Among the other three scenarios, scenario 2, which includes TH then AD, has the highest IRR, ascribed to the highest electricity sales for this scenario.

Thus, the NPV and IRR values for the four scenarios indicate that scenario 2 (TH then AD) will lead to the least negative NPV and higher IRR values. The higher income generated in this scenario is due to the larger quantity of electricity produced.

6.6 Overview of financial analyses of biogas plant

Table 22. Economic analysis summary for all four scenarios

Project parameters	Scenario 1	Scenario 2	Scenario 3	Scenario 4
Unit	$Million or ($Million/year)			
CapEx				
Infrastructure	3.2	5.6	8.3	9.1
Development and Commissioning costs	3.1	5.4	7.9	8.7
Glenfield	5.3	9.4	13.8	15.1
EPCM fee	4.6	8.0	11.9	13
CapEx total (including above terms)	31.5	55.1	81.5	89.2
CapEx contingency (inherent 22% and contingent risks 28%)	17.7	31	45.8	50.1
CapEx total including contingency	49.1	86.1	127.3	139.2
Total Capital Investment (TCI) required	49.1	86.1	127.3	139.2
OpEx				
Electricity	0.04	8.93	0.19	1.33
Digestate processing	0.03	0.01	0.03	0.04
Biosolids disposal	0.73	0.20	0.13	0.28
O&M cost	1.47	2.58	3.82	4.18
Other OpEx	7.75	11.58	14.79	15.86
Total OpEx	10.02	23.30	18.96	21.69
Revenue				
Green certificate	0.03	0.03	0.03	0.04
ACCUs	0.06	0.08	0.06	0.1
Electricity—grid	0.18	0.23	0.18	0.27
Revenue—total	0.27	0.35	0.27	0.4
Net profit (P_{NET})				
IRR (% pa)	Very high	93	85	88
NPV ($)	-78.5	-164.5	-169.1	-191

6.7 Summary of findings of technoeconomic analysis

The main objective of this project is to compare the technical and economic aspects of the conventional AD process with two treatment approaches, TH pretreatment of WAS and WAO treatment of AD digestate. Two new processes were designed incorporating the TH and WAO to achieve the objective, and detailed process flow diagrams (PFD) were constructed. After a comprehensive literature review, a detailed description of the two processes was prepared, incorporating the stream flow rates, composition, physical properties, and operating conditions. Information like separation ratio and component recovery data were obtained from the literature to complete the mass and energy balance calculations for all the separation equipment in the PFD. Reaction kinetics models were required to complete the mass balance calculations for reactors like anaerobic digester, TH, and WAO units. Due to the limited scope of experimental work to generate the reaction kinetic data in the current project, the kinetic models were obtained from the literature. Although detailed reaction stoichiometric equations and kinetic models are available in the literature for all four phases of anaerobic digestion, only limited information was available for TH and WAO reactions. The energy balance calculations were carried out for phase separation equipment and reactors, fluid moving equipment like pumps and compressors, heat exchangers, and mixers.

All the mass and energy balance calculations were carried out using Aspen Plus software. The process models for the above four scenarios were successfully constructed and simulated. The simulation models developed in this work are valuable tools for conducting numerous alternative process analyses like scale-up, scale-down, capacity variation, composition variation and operating condition variation. It must be pointed out that the simulation of some of the reactors relied on approximate kinetic models due to the absence of reliable data in the literature. However, these knowledge gaps provide ample opportunities for further research to generate reliable kinetic data and use them in the simulation models, thereby improving the accuracy of the overall process models. For the first time, the construction of Aspen Plus simulation models for TH and WAO treatment operations has been done in this work.

The WAO treatment of AD digestate was assumed to occur in an underground concentric pipe flow reactor (GPV) to use the geothermal heat to power the WAO reactions. The pipe flow reactor's depth used was 750 m. However, despite the additional interest in acid hydrolysis of the WAO products it took a long time to determine the optimum concentration of acid and injection depth as no industry data was available. Five different acid concentrations at 10 different depths from the ground level were investigated to obtain the right location for injecting acid and with a minimal amount of acid was identified which provided minimal inhibition to the AD process. The results generated in the above analysis will be valuable to Sydney Water in its further studies on WAO.

The mass and energy balance results generated in the process simulation studies were used to size every piece of equipment in all four scenarios. The sizing calculations enabled reliable estimation of the equipment capital cost, thereby eliminating the approximations usually found in the literature based on the overall plant capacity and exponential method. Similarly, the OpEx and working capital estimations were also carried out using detailed methods compared to the usual way of estimating them as an approximate fraction of the CapEx. The profitability indicators like NPV and IRR were determined using detailed discounted cash flow analysis compared to approximate methods used in the literature.

Using the process simulation models developed in Aspen Plus, mass and energy balance models were constructed in an MS Excel spreadsheet. The objective of building process simulation models in MS Excel is to connect them with the remaining techno-economic analysis steps, which include major and minor equipment sizing, capital and operating cost estimations and cash flow analysis. The construction of a spreadsheet for carrying out all the above calculations enables us to develop a comprehensive techno-economic tool for the first time to compare the advantages of TH and WAO as alternative treatment options. There is no such tool available in the open literature at present. This MS Excel file also has a functionality of changing the feed flow rate to provide easy to use tool for Sydney Water. However, for changing any other operating parameters, Aspen Plus software is required to be utilised for new operating conditions.

6.8 Suggestions for further work

- More work is required on the WAO reaction kinetics to increase the accuracy of the data obtained from the simulation.
- It was observed from the experiment that the destruction of the total solids was around 25% after using the GPV. However, the Aspen Plus simulation showed lower total solids destruction under the same conditions while all the oxygen was consumed after the mixing point. This would suggest that other reactions occur in the GPV which are independent of oxygen. Such reactions may occur due to the high temperature leading to more solids destruction, hence this is also an area for further study.
- For methane production, lower operating temperature (<240°C) in GPV should be tested (Zhang, Li, & Li 2021).

7. Industry Reference Group (IRG) Engagement

Introduction

The IRG was established in the first week after the funding agreement was signed with the RACE for 2030 CRC. Online IRG meetings were conducted using MS Teams to update members on project progress and seek advice on how outcomes aligned with the needs of business. A brief summary of the IRG meetings is presented below.

IRG Members

1. Water Services Association of Australia.
2. City of Gold Coast
3. Hunter Water Corporation
4. Water Corporation (WA)
5. South Australia Water.

IRG Meeting #1

The first IRG Meeting was held on 10th February 2022. The meeting commenced with an introduction of IRG members and the Project Team. Thereafter, project aim and objectives along with the role of IRG members in the project were presented. A brief background about the fast-track project was presented. Sydney Water produced 500 GL/year of sewage from AD which consists of 180,000 tons/year of biosolids. As a result, discarding these wastes lead to high transportation costs and show significant greenhouse gas emissions. Overall, the energy produced by Sydney Water AD plants is used to run the treatment plant electricity requirements (about 40-60%). Therefore, the goal of this project is to improve the solids treatment to reduce the transportation costs and energy requirements for the wastewater treatment plant by using the two proposed treatment techniques. Currently, Sydney Water has Thermal Hydrolysis (TH) pretreatment facility at St Mary's WWTP. This plant is said to be energy neutral which means that there is no positive gain in the energy in spite of adding a pretreatment process. Therefore, WAO treatment will be included in this project and shall be compared with TH. A comprehensive techno-economic analysis study to improve the profitability and financial viability of the whole biogas plant will be conducted. This study will begin with pretreatment of sewage sludge using the above two proposed pretreatment technologies. The biogas potential using AD of untreated and pretreated biomass will be determined in the lab-scale biochemical methane potential studies. Finally, technoeconomic feasibility studies using Aspen Plus modelling will be conducted.

In addition, rheology and chemical composition analysis of the sewage before and after treatment/AD will also be used to understand the effect of treatment on the substrates involved in this study. These studies are useful to design the pumps and mixers in the hydrolysis and biogas plants. Careful consideration is also required in terms of process parameters of the treatment especially temperature control is highly recommended.

The fluctuations in temperature range can result in calcinated material which can inhibit the biogas process. Currently the parameters are based on literature. However, through the project progress the effects of treatment process parameters will also be assessed.

IRG Meeting #2

The IRG meeting was held on 21st March 2022. The goal of the project is to have a full-scale WAO treatment facility for treating digestate obtained from a biogas plant treating waste activated sludge. Although WAO does not reduce the energy in terms of methane, the main idea is to incorporate WAO to destroy the biosolids produced in the digestate and recycle the solids to AD which will further improve the biogas yields.

This was observed in the chemical composition of the WAO treated samples. We could notice that there is still some organic material left in the digestate, which still has potential to produce biogas. However, the WAO treatment conditions used in the present study were close to partial oxidation as opposed to complete oxidation, which would have completely destroyed the organic material making it unsuitable for AD. In addition, PWAO reduces the dewatering costs, pasteurisation costs, less drying requirement, etc. Through initial screening TH pretreatment seems to be energy neutral and WAO has the potential to be energy positive. However, its unlikely to predict the most suitable treatment without a complete technical feasibility study which is the main objective of this project. In addition, WAO is effective at breaking down cellulose rich wastes, cardboard/FOGO and can be utilised as cosubtrates to further improve biogas production. This can be achieved by adjusting the carbon-nitrogen ratio of the substrates. For instance, waste sludge is a nitrogen rich source which can be mixed with carbon rich wastes (cardboard/FOGO) to adjust the C:N ratio in the range of 20-40. In addition to biogas production, the rheology of the substrates involved must also be measured to assess the flow characteristics which will be useful for mixing and pumping systems.

IRG Meeting # 3

The IRG meeting was held on 31st May 2022. The meeting was commenced with reintroduction of IRG Team members to Dani Alexander, the Program Leader at RACE. In this meeting, the chemical composition of biomass before and after the two treatment methods were presented. Later, the results of the BMP experiments for TH were presented. BMP results for WAO treatment are still ongoing, and results will be updated by next month. Through the end of the meeting the following points were suggested by the IRG members which will be addressed in the following meeting. Codigestion of WAS along with cellulose rich wastes such as FOGO/cardboard can be used to effectively degrade cellulose rich wastes. However, we must contact suppliers of real FOGO waste if we want to understand the rheological properties and methane yields as FOGO subjected to WAO. Griffith has already prepared FOGO based on the literature data. Furthermore, WAO is able to convert all complex organic material into organic acids, which may produce inhibitors. Despite improvement in biogas yields from the WAO treated biomass (WAO-DS) we must ensure that WAO can meet the environmental standards before disposal of residual solids. However, the preferable treatment method among TH and WAO can only be determined once a through techno-economic analysis of both treatment process were completed. While FOGO can be an excellent source of cellulose rich wastes to improve biogas yields, pumping solids into reactor can be challenging. In this case, using WAO is a suitable treatment as it can degrade even the most complex solid material. In addition, the solid waste produced after the AD process must also be hygienic to lower the cost of further treatment before using as compost. In other words, the treatment applied must be effective enough to reduce the dewatering expenses at the end of the AD process.

IRG Meeting # 4

The final IRG Meeting was 12th August. The introduction and project progress were reported by A/Prof Prasad Kaparaju (Griffith University), the Sydney Water representative and Professor Rajarathinam Parthasarathy (RMIT University). According to the plan, there were 4 scenarios developed which incorporated Scenario 1 (BAU - AD + Dewatering), Scenario 2 (AD then TH), Scenario 3 (AD + WAO) and Scenario 4 (AD + PWAO + recycle stream). From the treatments, TH only increased solubilisation by 30% while WAO reduced suspended solids by 50%, thereby improving the sCOD seventeen times. However, the methane potential of WAO didn't show any improvement. Thermal pretreatment however showed improved methane yields by 30%. The role of thermal treatment is to reduce the biosolids so it can be easily transferred through belt conveyor for transportation into other reactors. However, if methane yields should be excluded, the WAO treatment was much more feasible than TH in reducing biosolids. Scenario 3 (AD + 100% WAO without recycle) and Scenario 4 (AD + 20% PWAO with recycle stream) had the same size digestor capacity leading to lower chance of economic growth. As suggested by IRG members, Scenario 3 should have a larger anaerobic digestor capacity compared to Scenario 4 as there is an additional recycle stream. On the other hand, for the same digestor capacity we can either increase the retention time or build a larger digestor. Increasing the size will increase capital costs but the OpEx remains the same. Also, there was a discrepancy in the volume of the AD digestor because with the use of TH, the volume must be lower with higher solid concentration compared to the base Scenario 1. Another suggestion was made regarding the CapEx. that base scenario had 8 times lower CapEx compared to another scenario which looked like too big a difference. However, the digester volume and inclusion of treatment components, pumps that contributed to the CapEx as well. However, further investigation is required to understand the costs. The base case scenario consisting of AD and a dewatering/concentration unit ($10.9 million) is still very low. These values were compiled from literature therefore further modification from realistic values is required. The increase in CapEx from Scenario 1 to Scenario 4 will only be feasible if the NPV is positive. Also operating costs for Scenarios 2-4 shouldn't be much higher than the base case scenario unless it incorporated additional electricity costs and biosolids transportation. In addition, further optimisation of the PWAO process parameters is required to improve the methane yields in addition biosolid reduction so more energy can be produced in the AD.

8. Conclusion and Recommendation

A rapid review of the effect of TH on WAS (TH-WAS) and WAO treatment of anaerobic digestate (WAO-DS) demonstrated that TH can greatly enhance the biodegradability and biogas production from WAS and thereby reduce the hydraulic residence time in the AD. The rheological characteristics of TH-WAS showed to improve the dewaterability of the TH WAS. On the other hand, WAO treatment of AD digestate (WAO-DS) was found to be very effective in sludge reduction and conversion of sludge into useful products such as acetic acid. However, both TH and WAO treatments are energy-intensive processes and thus a thorough economic and technical feasibility of two treatments was essential.

Experimental results showed that TH could improve the methane yields of concentrated WAS (416.24 mL/g VS_{added}) by 10.1% through improved solubilisation of organic matter. On the other hand, WAO of the concentrated digestate at 200°C resulted in a reduction of both TS and VS content but would inhibit methane production at both 20% and 10% oxygen loading. However, WAO at 165°C and at 20% oxygen loading can produce methane yields of 117.32 mL/g VS_{added} with a lag phase of 6 days. The decrease in methane yields after WAO at 20% oxygen loading was 62% compared to untreated sample.

The economic analyses consisting of the total capital investment, operating costs, net present value and internal rate of return for all four scenarios showed that an AD plant with TH (Scenario 2) would be a better option than an AD plant with WAO treatment (Scenarios 3 and 4). However, both TH and WAO technologies were shown to be more expensive that the existing base case scenario (Scenario 1).

Both experimental and simulation studies thus suggests that further optimisation of the treatment conditions for WAO in terms of solids loading, treatment temperature and oxygen concentration needs to be determined. A pilot-scale study is needed to examine the effects of above process parameters before scaling up scenarios can be evaluated. As only a few studies on GPV were performed and the data is limited or seldom reported in literature, we recommend that Sydney Water further investigates optimising the process parameters for GPV technology.

9. References

Anaya Menacho, W, Mazid, AM & Das, N 2022, 'Modelling and analysis for biogas production process simulation of food waste using Aspen Plus', *Fuel*, vol. 309, p. 122058.

Andrews, JF 1968, 'A mathematical model for the continuous culture of microorganisms utilising inhibitory substrates', *Biotechnology and Bioengineering*, vol. 10, no. 6, pp. 707–723.

Angelidaki, I, Ellegaard, L & Ahring, BK 1999, 'A comprehensive model of anaerobic bioconversion of complex substrates to biogas', *Biotechnology and Bioengineering*, vol. 63, no. 3, pp. 363–372.

Appels, L., Baeyens, J., Degrève, J., & Dewil, R. (2008). Principles and potential of the anaerobic digestion of waste-activated sludge. *Progress in energy and combustion science, 34*(6), 755-781.

Balaji, R K, Rajan, K P & Ragula, U B R 2020, 'Modeling & optimisation of renewable hydrogen production from biomass via anaerobic digestion & dry reformation. International Journal of Hydrogen Energy', vol. 45, no. 36, pp. 18226–18240. https://doi.org/10.1016/j.ijhydene.2019.08.209

Bravo, D, Álvarez-Hornos, FJ, Penya-roja, JM, San-Valero, P & Gabaldón, C 2018, 'Aspen Plus process-simulation model: Producing biogas from VOC emissions in an anaerobic bioscrubber', *Journal of Environmental Management*, vol. 213, pp. 530–540.

Andrews, J. F. (1968). A mathematical model for the continuous culture of microorganisms utilizing inhibitory substrates. *Biotechnology and bioengineering, 10*(6), 707-723.

Angelidaki, I., Ellegaard, L., & Ahring, B. K. (1999). A comprehensive model of anaerobic bioconversion of complex substrates to biogas. *Biotechnology and bioengineering, 63*(3), 363-372.

Barnes, R., Dhanoa, M. S., & Lister, S. J. (1989). Standard normal variate transformation and de-trending of near-infrared diffuse reflectance spectra. *Applied spectroscopy, 43*(5), 772-777.

Bhoite, G. M., & Vaidya, P. D. (2018a). Improved Biogas Generation from Biomethanated Distillery Wastewater by Pretreatment with Catalytic Wet Air Oxidation. *Industrial & Engineering Chemistry Research, 57*(7), 2698-2704. doi:10.1021/acs.iecr.7b04281

Bhoite, G. M., & Vaidya, P. D. (2018b). Iron-catalyzed wet air oxidation of biomethanated distillery wastewater for enhanced biogas recovery. *Journal of Environmental Management, 226*, 241-248. doi:https://doi.org/10.1016/j.jenvman.2018.08.048

Bougrier, C., Albasi, C., Delgenès, J.-P., & Carrère, H. (2006). Effect of ultrasonic, thermal and ozone pre-treatments on waste activated sludge solubilisation and anaerobic biodegradability. *Chemical Engineering and Processing: Process Intensification, 45*(8), 711-718.

Bravo, D., Álvarez-Hornos, F., Penya-Roja, J., San-Valero, P., & Gabaldón, C. (2018). Aspen Plus process-simulation model: Producing biogas from VOC emissions in an anaerobic bioscrubber. *Journal of environmental management, 213*, 530-540.

Castro, Y. A., & Agblevor, F. A. (2020). Effect of wet air oxidation on the composition and biomethanation of water hyacinth. *Biomass Conversion and Biorefinery*, 1-12.

Dwyer, J., Griffiths, P., & Lant, P. (2009). Simultaneous colour and DON removal from sewage treatment plant effluent: Alum coagulation of melanoidin. *Water Research, 43*(2), 553-561. doi:https://doi.org/10.1016/j.watres.2008.10.053

Dwyer, J., Kavanagh, L., & Lant, P. (2008). The degradation of dissolved organic nitrogen associated with melanoidin using a UV/H2O2 AOP. *Chemosphere, 71*(9), 1745-1753. doi:https://doi.org/10.1016/j.chemosphere.2007.11.027

Dwyer, J., Starrenburg, D., Tait, S., Barr, K., Batstone, D. J., & Lant, P. (2008). Decreasing activated sludge thermal hydrolysis temperature reduces product colour, without decreasing degradability. *Water Research, 42*(18), 4699-4709. doi:https://doi.org/10.1016/j.watres.2008.08.019

Franzidis, J.-P., Porteous, A., & Anderson, J. (1982). The acid hydrolysis of cellulose in refuse in a continuous reactor. *Conservation & Recycling, 5*(4), 215-225.

Harun, N., Hassan, Z., Zainol, N., Ibrahim, W. H. W., & Hashim, H. (2019). Anaerobic digestion process of food waste for biogas production: a simulation approach. *Chemical Engineering & Technology, 42*(9), 1834-1839.

Higgins, M. J., Beightol, S., Mandahar, U., Suzuki, R., Xiao, S., Lu, H.-W., . . . Murthy, S. N. (2017). Pretreatment of a primary and secondary sludge blend at different thermal hydrolysis temperatures: Impacts on anaerobic digestion, dewatering and filtrate characteristics. *Water Research, 122*, 557-569. doi:https://doi.org/10.1016/j.watres.2017.06.016

Hii, K., Baroutian, S., Parthasarathy, R., Gapes, D. J., & Eshtiaghi, N. (2014). A review of wet air oxidation and thermal hydrolysis technologies in sludge treatment. *Bioresource technology, 155*, 289-299.

Hu, L., Guo, N., Hu, M., Yan, Z., Gao, R., Ma, J., . . . Peng, L. (2020). *Wet air oxidation (WAO) on sludge treatment and its first industrial application in China*. Paper presented at the IOP Conference Series: Earth and Environmental Science.

Inayat, A., Shahbaz, M., Khan, Z., Inayat, M., Mofijur, M., Ahmed, S., . . . Ahmad, A. A. (2021). Heat integration modeling of hydrogen production from date seeds via steam gasification. *International journal of hydrogen energy, 46*(59), 30592-30605.

Ko¨chermann, J., Mu¨hlenberg, J., & Klemm, M. (2018). Kinetics of hydrothermal furfural production from organosolv hemicellulose and d-xylose. *Industrial & Engineering Chemistry Research, 57*(43), 14417-14427.

Lorenzo-Llanes, J., Pages-Diaz, J., Kalogirou, E., & Contino, F. (2020). Development and application in Aspen Plus of a process simulation model for the anaerobic digestion of vinasses in UASB reactors: Hydrodynamics and biochemical reactions. *Journal of Environmental Chemical Engineering, 8*(2), 103540.

Menacho, W. A., Mazid, A. M., & Das, N. (2022). Modelling and analysis for biogas production process simulation of food waste using Aspen Plus. *Fuel, 309*, 122058.

Nguyen, H. H. (2014). *Modelling of food waste digestion using ADM1 integrated with Aspen Plus*. University of Southampton,

Nguyen, H. H., Heaven, S., & Banks, C. (2014). Energy potential from the anaerobic digestion of food waste in municipal solid waste stream of urban areas in Vietnam. *International Journal of Energy and Environmental Engineering, 5*, 365-374.

Pham, V. T., Wu, P.-H., Guan, C.-Y., Chang, C.-C., Liu, B.-L., Chang, C.-Y., & Yu, C.-P. (2021). Biogas Production and Microbial Communities in the Anaerobic Digestion of Sewage Sludge Under Hydrothermal Pretreatment with Air and a Catalyst. *BioEnergy Research, 14*(3), 828-843.

Rajendran, K., Kankanala, H. R., Lundin, M., & Taherzadeh, M. J. (2014). A novel process simulation model (PSM) for anaerobic digestion using Aspen Plus. *Bioresource technology, 168*, 7-13.

Ravendran, R., Abdulrazik, A., & Zailan, R. (2019). *Aspen Plus simulation of optimal biogas production in anaerobic digestion process*. Paper presented at the IOP Conference Series: Materials Science and Engineering.

Singh, D. K., & Garg, A. (2022). 18 - A review on hydrothermal pretreatment of sewage sludge: Energy recovery options and major challenges. In C. Hussain & S. Hait (Eds.), *Advanced Organic Waste Management* (pp. 297-314): Elsevier.

Suhartini, S., Lestari, Y., & Nurika, I. (2019). *Estimation of methane and electricity potential from canteen food waste*. Paper presented at the IOP conference series: earth and environmental science.

Symons, G., & Buswell, A. (1933). The methane fermentation of carbohydrates1, 2. *Journal of the american chemical society, 55*(5), 2028-2036.

Tarpani, R. R. Z., Alfonsin, C., Hospido, A., & Azapagic, A. (2020). Life cycle environmental impacts of sewage sludge treatment methods for resource recovery considering ecotoxicity of heavy metals and pharmaceutical and personal care products. *Journal of Environmental Management, 260*, 109643.

Tarpani, R. R. Z., & Azapagic, A. (2018). Life cycle costs of advanced treatment techniques for wastewater reuse and resource recovery from sewage sludge. *Journal of Cleaner Production, 204*, 832-847.

Thomsen, L. B. S., Anastasakis, K., & Biller, P. (2022). Wet oxidation of aqueous phase from hydrothermal liquefaction of sewage sludge. *Water Research, 209*, 117863.

Turton, R., Bailie, R. C., Whiting, W. B., & Shaeiwitz, J. A. (2008). *Analysis, synthesis and design of chemical processes*: Pearson Education.

Wid, N. (2018). Phosphorus Recovery from Anaerobically Digested Liquor of Screenings. *Anaerobic Digestion Processes: Applications and Effluent Treatment*, 177-182.

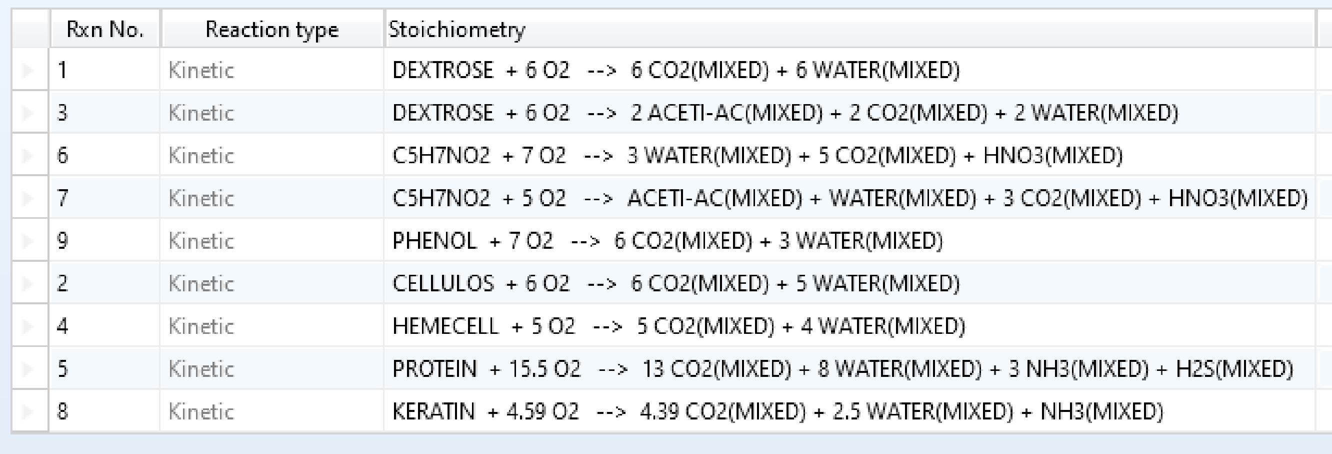
Wilson, C. A., & Novak, J. T. (2009). Hydrolysis of macromolecular components of primary and secondary wastewater sludge by thermal hydrolytic pretreatment. *Water Research, 43*(18), 4489-4498. doi:https://doi.org/10.1016/j.watres.2009.07.022

Zhang, D., Feng, Y., Huang, H., Khunjar, W., & Wang, Z.-W. (2020). Recalcitrant dissolved organic nitrogen formation in thermal hydrolysis pretreatment of municipal sludge. *Environment International, 138*, 105629. doi:https://doi.org/10.1016/j.envint.2020.105629

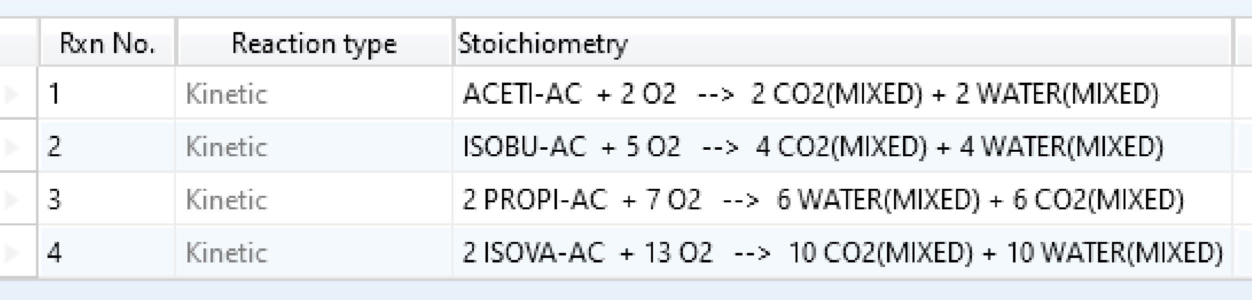
Zhang, Y., Li, H., & Li, D. (2021). Maximize methane recovery from sludge anaerobic digestion by combining an optimal wet air oxidation process. *Renewable Energy, 179*, 359-369.

10. APPENDICES

Appendix A. Partial WAO Reactions

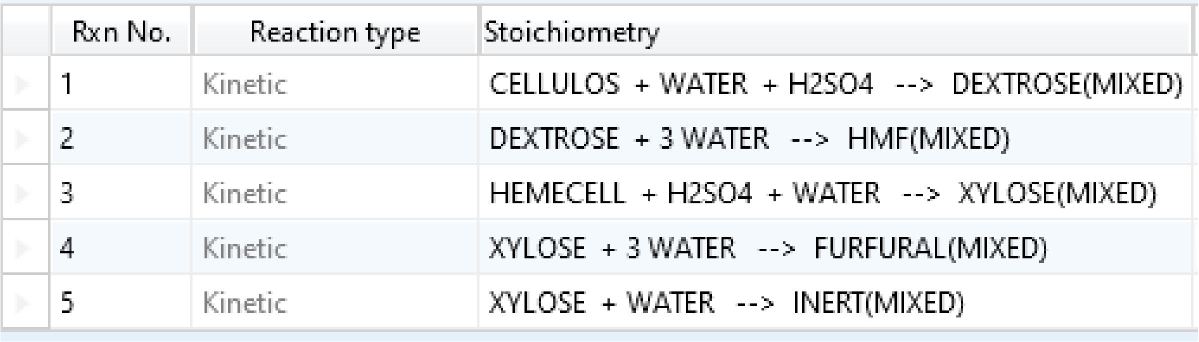


Rxn No.	Reaction type	Stoichiometry
1	Kinetic	DEXTROSE + 6 O2 --> 6 CO2(MIXED) + 6 WATER(MIXED)
3	Kinetic	DEXTROSE + 6 O2 --> 2 ACETI-AC(MIXED) + 2 CO2(MIXED) + 2 WATER(MIXED)
6	Kinetic	C5H7NO2 + 7 O2 --> 3 WATER(MIXED) + 5 CO2(MIXED) + HNO3(MIXED)
7	Kinetic	C5H7NO2 + 5 O2 --> ACETI-AC(MIXED) + WATER(MIXED) + 3 CO2(MIXED) + HNO3(MIXED)
9	Kinetic	PHENOL + 7 O2 --> 6 CO2(MIXED) + 3 WATER(MIXED)
2	Kinetic	CELLULOS + 6 O2 --> 6 CO2(MIXED) + 5 WATER(MIXED)
4	Kinetic	HEMECELL + 5 O2 --> 5 CO2(MIXED) + 4 WATER(MIXED)
5	Kinetic	PROTEIN + 15.5 O2 --> 13 CO2(MIXED) + 8 WATER(MIXED) + 3 NH3(MIXED) + H2S(MIXED)
8	Kinetic	KERATIN + 4.59 O2 --> 4.39 CO2(MIXED) + 2.5 WATER(MIXED) + NH3(MIXED)



Rxn No.	Reaction type	Stoichiometry
1	Kinetic	ACETI-AC + 2 O2 --> 2 CO2(MIXED) + 2 WATER(MIXED)
2	Kinetic	ISOBU-AC + 5 O2 --> 4 CO2(MIXED) + 4 WATER(MIXED)
3	Kinetic	2 PROPI-AC + 7 O2 --> 6 WATER(MIXED) + 6 CO2(MIXED)
4	Kinetic	2 ISOVA-AC + 13 O2 --> 10 CO2(MIXED) + 10 WATER(MIXED)

Appendix B. Partial WAO Reactions



Rxn No.	Reaction type	Stoichiometry
1	Kinetic	CELLULOS + WATER + H2SO4 --> DEXTROSE(MIXED)
2	Kinetic	DEXTROSE + 3 WATER --> HMF(MIXED)
3	Kinetic	HEMECELL + H2SO4 + WATER --> XYLOSE(MIXED)
4	Kinetic	XYLOSE + 3 WATER --> FURFURAL(MIXED)
5	Kinetic	XYLOSE + WATER --> INERT(MIXED)

Appendix C. Total elemental analysis of WAS and Digestate before and after tested treatments.

Elements (mg/kg TS)	WAS feed		Digestate from Anaerobic Digester		
	WAS as such	TH WAS Concentrate	Whole Digestate	Concentrate	Centrate
Al	0.00	2,690.85	0.00	7,708.01	345.95
As	0.74	3.00	1.29	5.35	0.45
B	18.06	3.79	24.62	0.00	4.78
Ba	107.59	103.91	199.05	213.58	19.27
Ca	8,689.20	7,999.88	20,764.80	17,158.73	2,840.42
Cd	0.59	0.39	0.84	0.78	0.06
Co	0.00	1.48	1.30	3.50	0.25
Cr	11.06	12.75	20.78	23.39	2.20
Cu	249.15	279.28	387.30	438.17	32.52
Fe	8,071.01	5,675.22	20,114.81	19,032.27	1,039.38
K	7,328.47	3,992.25	7,144.49	2,308.17	1,493.57
Mg	5,209.77	3,391.52	5,362.11	4,294.14	1,005.51
Mn	122.24	117.96	215.56	226.84	32.86
Mo	20.63	8.52	24.44	21.54	0.88
Na	2,131.11	1,720.82	7,151.90	2,011.20	661.53
Ni	14.84	114.09	24.96	23.66	2.00
P	24,967.51	18,232.58	25,182.36	20,194.85	1,964.11
Pb	8.38	7.07	17.80	18.16	1.79
S	7,451.01	8,468.18	11,699.50	12,618.99	1,040.14
Se	0	3.72	5.62	5.36	0.28
Zn	521.51	581.59	799.07	916.94	58.35

Appendix D. Total elemental analysis in post-BMP assays of WAS feed and WAS digestate with and without tested treatments.

Elements (mg/kg TS)	WAS feed		WAS Digestate			
	WAS as such	TH WAS Concentrate	Whole Digestate	Concentrate	Centrate	WAO Digestate Concentrate
Al	2.52	13511.21	2.26	16,798.47	12,572.52	8,019.90
As	10.12	18.28	7.70	11.79	16.75	0.00
B	0.81	1.94	0.00	0.00	0.94	0.00
Ba	245.95	498.54	243.27	505.14	459.92	235.22
Ca	18,902.24	44,145.46	20,186.19	44,993.48	48,627.73	18,920.28
Cd	1.04	2.10	1.02	1.47	2.28	0.73
Co	5.35	13.02	4.71	10.45	9.99	0.00
Cr	38.41	107.40	38.60	58.31	73.23	43.62
Cu	626.56	1276.62	577.58	1,116.55	1,366.19	556.60
Fe	20,320.02	44,662.75	21,929.37	49,812.27	42,356.02	22,097.94
K	13,965.99	28,799.32	12,384.02	11,063.01	35,987.62	13,679.56
Mg	5,946.10	11,508.45	5,642.00	9,366.01	9,315.14	7,392.37
Mn	177.80	353.17	189.81	460.75	287.32	156.36
Mo	16.36	39.12	18.78	43.53	30.22	15.02
Na	11,335.95	24,718.41	11,428.89	10,200.47	33,797.96	12,450.00
Ni	44.21	158.64	33.34	46.78	69.55	39.20
P	36,234.51	70,124.10	31,266.24	51,323.98	64,643.43	38,129.37
Pb	31.94	67.06	33.35	51.15	65.51	28.78
S	16,938.32	38,646.61	16,464.51	31,963.58	32,124.31	16,782.72
Se	5.25	8.50	3.93	10.41	7.62	0.00
Zn	1,029.18	6,913.78	990.17	2,074.15	1,885.06	890.40

Appendix E. FTIR Spectral analysis for all samples

Peak number	X (cm^{-1})	Y (A)	Appearance	Group	Compound Class	Comments
Inoculum Concentrate						
1	2,924.92	0.03	medium	C-H stretching	alkane	
2	1,628.79	0.05	medium	C=C stretching	cyclic alkene	
3	1,392.5	0.04	medium	C-H bending	aldehyde	
4	1,027.06	0.06		C-F stretch	Aliphatic fluoro compounds	
Untreated Digestate Concentrate						
1	2,923.47	0.02	medium	C-H stretching	alkane	
2	1,639.89	0.03	strong	C=C stretching	alkene	monosubstituted
3	1,007.41	0.06	strong	C-F stretch	Aliphatic fluoro compounds	
Untreated WAS Concentrate						
1	2,924.15	0.01	medium	C-H stretching	alkane	
2	1,007.19	0.03		C-F stretch	Aliphatic fluoro compounds	
Wet air oxidation 20%						
1	2,923.81	0.07	medium	C-H stretching	alkane	
2	1,658.82	0.08	medium	C=C stretching	cyclic alkene	
3	1,453.8	0.08	medium	C-H bending	alkane	methyl group
4	1,007.31	0.31		C-F stretch	Aliphatic fluoro compounds	
5	797.94	0.11	strong	C-H bending	1,2,3-trisubstituted	
6	534.74	0.27	strong	C-I stretching	halo compound	
7	465.9	0.35				
Wet air oxidation 10% oxygen loading						
1	3,694.48	0.03	medium, sharp	O-H stretching	alcohol	Free
2	3,223.17	0.06	strong, broad	O-H stretching	alcohol	intermolecular bonded
3	2,923.23	0.12	medium	C-H stretching	alkane	
4	2,852.8	0.09	medium	C-H stretching	alkane	
5	1,655.13	0.13	medium	C=C stretching	cyclic alkene	

6	1,452.86	0.12	medium	C-H bending	alkane	methyl group
7	1,028.57	0.38	strong	C-F stretch	Aliphatic fluoro compounds	
8	1,006.96	0.39	strong	C-F stretch	Aliphatic fluoro compounds	
9	797.27	0.16	medium	C=C bending	alkene	Trisubstituted
10	750.27	0.16	strong	C-H bending	monosubstituted	
11	694.51	0.18	strong	C-Br stretching	halo compound	
12	531.1	0.35	strong	C-I stretching	halo compound	

Appendix F. Economic analysis

Appendix F1. Capital Cost Estimation ($AUD million)

Equipment	**Scenario 1: Business as usual of anaerobic digestion of WAS**	**Scenario 2: TH of WAS followed by AD**	**Scenario 3: AD of WAS followed by 100% WAO**	**Scenario 4: AD of WAS followed by 20% WAO**	**Source**
AD Feed Tank[1]	0.69	0.42	0.69	0.79	Sydney Water
AD Tank	8.02	5.53	8.02	7.24	Sydney Water
Press Belt Filter (Dewatering)	0.60	0.34	0.16	0.30	Sydney Water
Transportation Pumps	0.06	0.15	0.15	0.18	Aspen plus; CAPCOST spreadsheet
CHP	1.02	2.62	1.02	1.53	Abu-Orf & Goss (2012); Suhartini, Lestari and Nurika (2019)
Solid bowl Centrifuge (Thickener)	-	0.21	0.21	0.24	Sydney Water
TH reactor & steam boiler	-	8.29	-	-	Aspen plus & CAPCOST spreadsheet
Flash Drum	-	-	0.03	0.06	Aspen plus & CAPCOST spreadsheet
TH Feed Tank	-	0.29	-	-	Sydney Water
Flash drum with Cooler	-	0.34	-	-	Aspen plus & CAPCOST spreadsheet
Cooler	-	-	-	0.04	Aspen plus & CAPCOST spreadsheet
Steam heater	-	-	0.05	0.05	Aspen plus & CAPCOST spreadsheet

Equipment	Scenario 1: Business as usual of anaerobic digestion of WAS	Scenario 2: TH of WAS followed by AD	Scenario 3: AD of WAS followed by 100% WAO	Scenario 4: AD of WAS followed by 20% WAO	Source
GPV reactor	-	-	14.04	16.50	Environment, (1985) & Industry Quote
Compressors			2.13	2.13	Aspen plus & CAPCOST spreadsheet
Cavity high-pressure pump	-	-	0.09	0.19	Aspen plus & CAPCOST spreadsheet
Neutralization Tank	-	-	0.31	0.52	Sydney Water

	Scenario 1: Business as usual of anaerobic digestion of WAS	Scenario 2: TH of WAS followed by AD	Scenario 3: AD WAS followed by 100% WAO	Scenario 4: AD of WAS followed by 20% WAO
Including Development and Commissioning costs	$18,330,432	$32,116,744	$47,473,687	$51,935,721
Including EPCM fee	$22,913,040	$40,145,930	$59,342,108	$64,919,651
Including Infrastructure	$26,120,866	$45,766,360	$67,650,003	$74,008,402
including Glenfield	$31,467,242	$55,133,743	$81,496,495	$89,156,321
Contingency-inherent risk	$6,922,793	$12,129,424	$17,929,229	$19,614,391
Most likely delivery cost	$38,390,035	$67,263,167	$99,425,724	$108,770,711
Contingency - Contingent risk	$10,749,210	$18,833,687	$27,839,203	$30,455,799
Base estimate + contingency	$49,139,245	$86,096,854	$127,264,927	$139,226,510

Appendix F2: Operating Cost Estimation

Scenario 1: Business as usual of anaerobic digestion of WAS

	Unit cost, $/(tonne or kJ)	Consumption in tonne (or kJ)/(year)	$million/ year	$/ tonne product
Utilities				
Electricity	0.0001	733,803,223	0.04	1.36
Chemicals and substances for AD (Anaerobic Digestion Utility)	1$ per feed tonne to AD		0.0315	1.000
	Subtotal		0.07	2.36
Process Labour				
Process Labour	(5 operators/shift, 40 hr/week, 4 shifts, 100,000 per year)		$2.00	$63.42
Supervision labour	15% of process labour		$0.30	$9.51
Payroll overheads	20% of process labour		$0.40	$12.68
Plant overheads	20% of process labour and supervision labour		$0.46	$14.59
Annual cost of maintenance	Estimated at 3% of total fixed capital		$1.47	$46.75
Annual depreciation	Calculation in Spreadsheet (Annual Depreciation)		$1.97	$62.33
Insurance and local taxes	Estimated at 4% of total fixed capital		$1.97	$62.33
	Subtotal		$8.57	$271.60
Other Expenses				

Transportation		100	7,256.67407	$0.73	$23.01
	Subtotal			$0.73	$23.01
Direct Manufacturing Cost				$9.37	$296.97
Non-manufacturing cost					
R & D (Research and Development expenses)	2.0% of annual manufacturing cost			0.19	5.94
Corporate administration	5.0% of annual manufacturing costs			0.47	14.85
	Subtotal			0.66	20.79
General Expenses Non-manufacturing Cost				0.66	20.79
Total Operation Cost				10.02	$317.76

Appendix F2: Operating Cost Estimation

Scenario 2: TH of WAS followed by AD

	Unit cost, $/(tonne/ m3 or kJ)	Consumption in tonne or kJ)/(year)	$million/year	$/tonne product
Utilities				
Electricity	0.0001	153,007,896,060.48	$8.93	$283.02
Chemicals and substances for AD (Anaerobic Digestion Utility)	1$ per feed tonne to AD		$0.01	$0.40
	Subtotal		$8.94	$283.43
Process Labour				
Process Labour	(50operators/shift, 40hr/week, 4 shifts, 100,000 per year)		$2.00	$63.42
Supervision labour	15% of process labour		$0.30	$9.51
Payroll overheads	20% of process labour		$0.40	$12.68
Plant overheads	20% of process labour and supervision labour		$0.46	$14.59
Annual cost of maintenance	Estimated at 3% of total fixed capital		$2.58	$81.90
Annual depreciation	Calculation in Spreadsheet (Annual Depreciation)		$3.44	$109.20
Insurance and local taxes	Estimated at 4% of total fixed capital		$3.44	$109.20
	Subtotal		$12.63	$400.52
Other Expenses				
Transport	100	2,036.550224	$0.20	$6.46
	Subtotal		$0.20	$6.46
Direct Manufacturing Cost			$21.77	$690.40
Non-manufacturing cost				
R & D (Research and Development expenses)	2.0% of annual manufacturing cost		$0.44	$13.81
Corporate administration	5.0% of annual manufacturing costs		$1.09	$34.52
	Subtotal		$1.52	$48.33
General Expenses Non-manufacturing Cost			$1.52	$48.33
Total Operation Cost			$23.30	$738.73

Appendix F2: Operating Cost Estimation

Scenario 3: AD WAS followed by 100% WAO

	Unit cost, \$/(tonne/ m^3 or kJ)	Consumption in tonne or kJ)/(year)	\$million/year	\$/tonne product
Raw Materials				
Liquid Oxygen	0.13	558,128.24	0.07	2.30
$Ca(OH)_2$	352.09	326.66	0.12	3.65
H_2SO_4	347.61	79.84	0.03	0.88
	Subtotal		0.22	6.83
Utilities				
Electricity	0.0001	3,193,368,647.20	0.19	5.91
Chemicals and substances for AD	1\$ per feed tonne to AD		0.032	1.00
	Subtotal		0.22	6.91
Process Labour				
Process Labour	(5 operators/shift, 40hr/week, 4 shifts, 100,000 per year)		2.00	63.42
Supervision labour	15% of process labour		0.30	9.51
Payroll overheads	20% of process labour		0.40	12.68
Plant overheads	20% of process labour and supervision labour		0.46	14.59
Annual cost of maintenance	Estimated at 3% of total fixed capital		3.82	121.07
Annual depreciation	Calculations in Spreadsheet (Annual Depreciation)		5.09	161.42
Insurance and local taxes	Estimated at 4% of total fixed capital		5.09	161.42
	Subtotal		17.16	544.11
Other Expenses				
Transport	100.00	1,292.37	0.13	4.10
	Subtotal		0.13	4.10
Direct Manufacturing Cost			17.72	561.95
Non-manufacturing cost				
R & D (Research and Development expenses)	2.0% of annual manufacturing cost		0.35	11.24
Corporate administration	5.0% of annual manufacturing costs		0.89	28.10
	Subtotal		1.24	39.34
General Expenses Non-manufacturing Cost			1.24	39.34
Total Operation Cost			18.96	601.28

Appendix F2: Operating Cost Estimation

Scenario 4: AD of WAS followed by 20% WAO

	Unit cost, \$/(tonne/ m^3 or kJ)	Consumption in tonne or kJ)/(year)	\$million/year	\$/tonne product
Raw Materials				
Liquid Oxygen	0.13	301,180.45	0.04	1.24
$Ca(OH)_2$	352.09	227.05	0.08	2.53
H_2SO_4	347.61	95.50	0.03	1.05
	Subtotal		0.15	4.83
Utilities				
Electricity	0.0001	22,722,302,360	1.33	42.03
Chemicals and substances for AD (Anaerobic Digestion Utility)	1\$ per feed tonne to AD		0.04	1.26
	Subtotal		1.37	43.29
Process Labour				
Process Labour	(5 operators/shift, 40hr/week, 4 shifts, 100,000 per year)		2.00	63.42
Supervision labour	15% of process labour		0.30	9.51
Payroll overheads	20% of process labour		0.40	12.68
Plant overheads	20% of process labour and supervision labour		0.46	14.59
Annual cost of maintenance	Estimated at 3% of total fixed capital		4.18	132.45
Annual depreciation	Calculation in Spreadsheet (Annual Depreciation)		5.57	176.59
Insurance and local taxes	Estimated at 4% of total fixed capital		5.57	176.59
	Subtotal		18.47	585.84
Other Expenses				
Transport	100.00	2773.196947	0.28	8.79
	Subtotal		0.28	8.79
Direct Manufacturing Cost			20.27	642.75
Non-manufacturing cost				
R & D (Research and Development expenses)	2.0% of annual manufacturing cost		0.41	12.85
Corporate administration	5.0% of annual manufacturing costs		1.01	32.14
	Subtotal		1.42	44.99
General Expenses Non-manufacturing Cost			1.42	44.99
Total Operation Cost			21.69	687.74

Appendix F3: Revenue Estimation

Scenario 1: Business as usual of anaerobic digestion WAS

Term	Value	Units
Methane production	220,783.2093	m^3/year
Methane value in electricity	9.68	kWh/m^3 CH_4
Methane value in heat	0.038	GJ/m^3 CH_4
1 tonne of Coal	2.65	MWh
1 MWh Coal electricity	0.986	tonne of CO_2/MWh
Carbon credits value	30	$/tonne CO_2
Methane cost electricity	0.085	$/kWh
Green Certificate	3	$/GJ
Revenue	**$million/year**	**$/tonne product**
Revenue 1: Carbon credits	0.06	2.00
Revenue 2: CH_4 Electricity Cost	0.18	5.76
Revenue 1: Green Certificate	0.03	0.80
TOTAL	0.27	8.56

Scenario 2: TH of WAS followed by AD

Term	Value	Units
Methane production	284,307.39	m^3/year
Methane value in electricity	9.68	kWh/m^3 CH_4
Methane value in heat	0.038	GJ/m^3 CH_4
1 tonne of Coal	2.65	MWh
1 MWh Coal electricity	0.986	tonne of CO_2/MWh
Carbon credits value	30	$/tonne CO_2
Methane cost electricity	0.085	$/kWh
Green Certificate	3	$/GJ
Revenue	**$million/year**	**$/tonne product**
Revenue 1: Carbon credits	0.08	2.58
Revenue 2: CH_4 Electricity Cost	0.23	7.42
Revenue 1: Green Certificate	0.03	1.03
TOTAL	0.35	11.03

Scenario 3: AD WAS followed by 100% WAO

Term	Value	Unit
Methane production	220,783.2093	m^3/year
Methane value in electricity	9.68	kWh/m^3 CH_4
Methane value in heat	0.038	GJ/m^3 CH_4
1 tonne of Coal	2.65	MWh
1 MWh Coal electricity	0.986	tonne of CO_2/MWh

Carbon credits value	30	\$/tonne CO_2
Methane cost electricity	0.085	\$/kWh
Green Certificate	3	\$/GJ
Revenue	**\$million/year**	**\$/tonne product**
Revenue 1: Carbon credits	0.06	2.00
Revenue 2: CH_4 Electricity Cost	0.18	5.76
Revenue 1: Green Certificate	0.03	0.80
TOTAL	0.27	8.56

Scenario 4: AD of WAS followed by 20% WAO

Term	**Value**	**Units**
Methane production	330,831.34	m^3/year
Methane value in electricity	9.68	kWh/m^3 CH_4
Methane value in heat	0.038	GJ/m^3 CH_4
1 tonne of Coal	2.65	MWh
1 MWh Coal electricity	0.986	tonne of CO_2/MWh
Carbon credits value	30	\$/tonne CO_2
Methane cost electricity	0.085	\$/kWh
Green Certificate	3	\$/GJ
Revenue	**\$million/year**	**\$/tonne product**
Revenue 1: Carbon credits	0.095	3.00
Revenue 2: CH_4 Electricity Cost	0.272	8.63
Revenue 1: Green Certificate	0.038	1.20
TOTAL	0.40	12.8

Appendix F4: Cash Flow

Scenario 1: Business as usual of anaerobic digestion of WAS

Year	Annual total capital expenditure A_{TC}	Annual sales income A_S	Annual fixed expenses A_{FE}	Annual variable expenses A_{VE}	Annual total expenses $A_{TE}=A_{FE}+A_{VE}$	Annual Cash Income $A_{CI}=A_S-A_{TE}$	Annual depreciation A_D	Annual Income tax $A_{IT}=(A_{CI}-A_D)T$	Annual Net Cash Income $A_{NCI}=A_{CI}-A_{IT}$	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Cumulative Annual Cash Flow $\sum A_{CF}$
2022	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
2023	$50.19	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	-$50.19	-$50.19
2024	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$55.05
2025	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$59.91
2026	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$64.77
2027	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$69.63
2028	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$74.49
2029	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$79.35
2030	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$84.21
2031	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$89.07
2032	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$93.93
2033	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$98.79
2034	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$103.65
2035	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$108.51
2036	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$113.37
2037	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$118.23
2038	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$123.09
2039	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$127.95
2040	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$132.81
2041	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$137.67
2042	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$142.53
2043	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$147.39

2044	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$152.25
2045	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$157.11
2046	$0.00	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$4.86	-$161.97
2047	-$1.05	$0.27	$7.26	$0.80	$8.06	-$7.79	$1.97	-$2.93	-$4.86	-$3.81	-$165.78

Scenario 2: TH of WAS followed by AD

Year	Annual total capital expenditure A_{TC}	Annual sales income A_S	Annual fixed expenses A_{FE}	Annual variable expenses A_{VE}	Annual total expenses $A_{TE}=A_{FE}+A_{VE}$	Annule Cash Income $A_{CI}=A_S-A_{TE}$	Annual depreciation A_D	Annual Income tax $A_{IT}=(A_{CI}-A_D)T$	Annual Net Cash Income $A_{NCI}=A_{CI}-A_{IT}$	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Cumulative Annual Cash Flow $\sum A_{CF}$
2022	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
2023	$86.10	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	-$86.10	-$86.10
2024	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$98.72
2025	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$111.34
2026	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$123.96
2027	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$136.58
2028	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$149.20
2029	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$161.82
2030	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$174.44
2031	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$187.06
2032	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$199.68
2033	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$212.30
2034	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$224.92
2035	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$237.54
2036	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$250.16
2037	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$262.78
2038	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$275.40
2039	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$288.02
2040	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$300.64
2041	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$313.26

2042	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$325.88
2043	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$338.50
2044	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$351.12
2045	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$363.74
2046	$0.00	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$12.62	-$376.36
2047	-$2.07	$0.35	$10.71	$9.14	$19.85	-$19.50	$3.44	-$6.88	-$12.62	-$10.55	-$386.91

Scenario 3: AD WAS followed by 100% WAO of digestate

Year	Annual total capital expenditure A_{TC}	Annual sales income A_S	Annual fixed expenses A_{FE}	Annual variable expenses A_{VE}	Annual total expenses $A_{TE}=A_{FE}+A_{VE}$	Annual Cash Income $A_{CI}=A_S-A_{TE}$	Annual depreciation A_D	Annual Income tax $A_{IT}=(A_{CI}-A_D)T$	Annual Net Cash Income $A_{NCI}=A_{CI}-A_{IT}$	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Cumulative Annual Cash Flow $\sum A_{CF}$
2022	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
2023	$127.26	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	-$127.26	-$127.26
2024	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$135.26
2025	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$143.25
2026	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$151.25
2027	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$159.24
2028	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$167.23
2029	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$175.23
2030	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$183.22
2031	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$191.22
2032	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$199.21
2033	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$207.20
2034	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$215.20
2035	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$223.19
2036	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$231.18
2037	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$239.18
2038	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$247.17
2039	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$255.17
2040	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$263.16
2041	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$271.15
2042	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$279.15
2043	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$287.14
2044	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$295.13
2045	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$303.13

2046	$0.00	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$7.99	-$311.12
2047	-$1.91	$0.27	$13.31	$0.56	$13.87	-$13.60	$5.09	-$5.61	-$7.99	-$6.09	-$317.21

Scenario 4: AD of WAS followed by 20% WAO of digestate

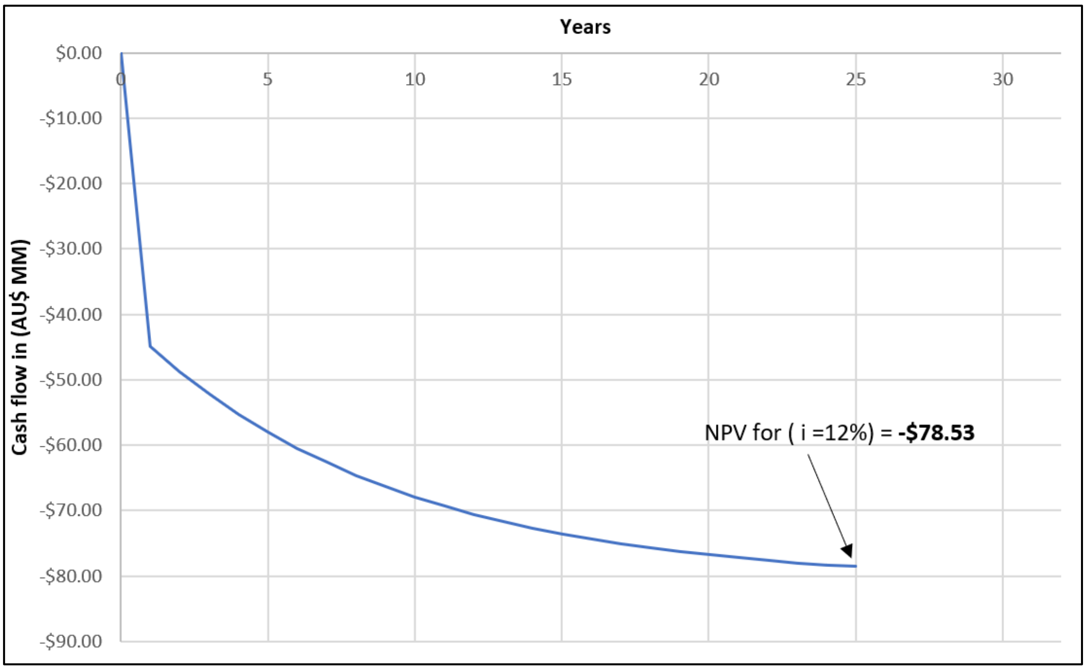
Year	Annual total capital expenditure A_{TC}	Annual sales income A_S	Annual fixed expenses A_{FE}	Annual variable expenses A_{VE}	Annual total expenses $A_{TE}=A_{FE}+A_{VE}$	Annual Cash Income $A_{CI}=A_S-A_{TE}$	Annual depreciation A_D	Annual Income tax $A_{IT}=(A_{CI}-A_D)T$	Annual Net Cash Income $A_{NCI}=A_{CI}-A_{IT}$	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Cumulative Annual Cash Flow $\sum A_{CF}$
2022	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
2023	$141.39	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	-$141.39	-$141.39
2024	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$150.72
2025	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$160.05
2026	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$169.38
2027	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$178.71
2028	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$188.03
2029	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$197.36
2030	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$206.69
2031	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$216.02
2032	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$225.35
2033	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$234.68
2034	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$244.01
2035	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$253.34
2036	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$262.67
2037	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$272.00
2038	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$281.33
2039	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$290.66
2040	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$299.99
2041	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$309.32
2042	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$318.65
2043	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$327.98

2044	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$337.31
2045	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$346.64
2046	$0.00	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$9.33	-$355.97
2047	-$2.16	$0.40	$14.32	$1.79	$16.12	-$15.71	$5.57	-$6.39	-$9.33	-$7.17	-$363.14

Appendix F5: NPV and IRR Calculations

Scenario 1: Business as usual of anaerobic digestion of WAS

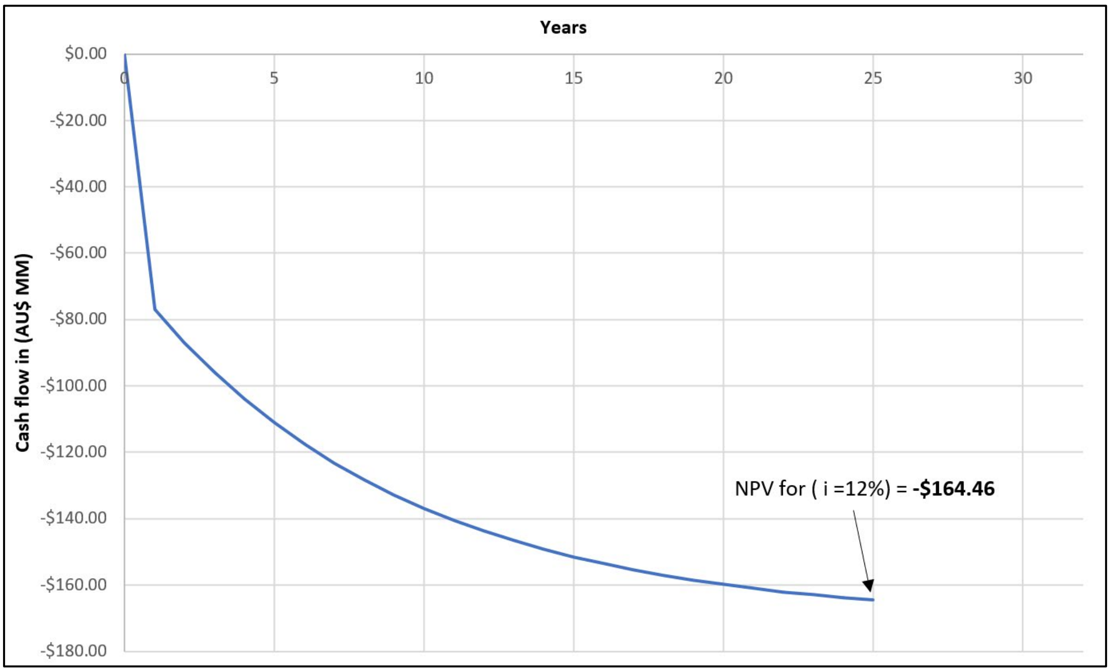
Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D For i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1	$0.00	$0.00
2023	1	-$50.19	0.893	-$44.81	-$44.81
2024	2	-$4.86	0.798	-$3.87	-$48.68
2025	3	-$4.86	0.712	-$3.46	-$52.14
2026	4	-$4.86	0.636	-$3.09	-$55.23
2027	5	-$4.86	0.567	-$2.76	-$57.99
2028	6	-$4.86	0.507	-$2.46	-$60.45
2029	7	-$4.86	0.452	-$2.20	-$62.65
2030	8	-$4.86	0.404	-$1.96	-$64.61
2031	9	-$4.86	0.361	-$1.75	-$66.37
2032	10	-$4.86	0.322	-$1.56	-$67.93
2033	11	-$4.86	0.287	-$1.40	-$69.33
2034	12	-$4.86	0.257	-$1.25	-$70.58
2035	13	-$4.86	0.229	-$1.11	-$71.69
2036	14	-$4.86	0.205	-$0.99	-$72.68
2037	15	-$4.86	0.183	-$0.89	-$73.57
2038	16	-$4.86	0.163	-$0.79	-$74.36
2039	17	-$4.86	0.146	-$0.71	-$75.07
2040	18	-$4.86	0.130	-$0.63	-$75.70
2041	19	-$4.86	0.116	-$0.56	-$76.27
2042	20	-$4.86	0.104	-$0.50	-$76.77
2043	21	-$4.86	0.093	-$0.45	-$77.22
2044	22	-$4.86	0.083	-$0.40	-$77.62
2045	23	-$4.86	0.074	-$0.36	-$77.98
2046	24	-$4.86	0.066	-$0.32	-$78.30
2047	25	-$3.81	0.059	-$0.22	-$78.53



Scenario 2: TH of WAS followed by AD

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D for i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1	$0.00	$0.00
2023	1	-$86.10	0.893	-$76.87	-$76.87
2024	2	-$12.62	0.798	-$10.06	-$86.93
2025	3	-$12.62	0.712	-$8.98	-$95.92
2026	4	-$12.62	0.636	-$8.02	-$103.94
2027	5	-$12.62	0.567	-$7.16	-$111.10
2028	6	-$12.62	0.507	-$6.39	-$117.49
2029	7	-$12.62	0.452	-$5.71	-$123.20
2030	8	-$12.62	0.404	-$5.10	-$128.30
2031	9	-$12.62	0.361	-$4.55	-$132.85
2032	10	-$12.62	0.322	-$4.06	-$136.91
2033	11	-$12.62	0.287	-$3.63	-$140.54
2034	12	-$12.62	0.257	-$3.24	-$143.78
2035	13	-$12.62	0.229	-$2.89	-$146.67
2036	14	-$12.62	0.205	-$2.58	-$149.25
2037	15	-$12.62	0.183	-$2.31	-$151.56
2038	16	-$12.62	0.163	-$2.06	-$153.62
2039	17	-$12.62	0.146	-$1.84	-$155.46
2040	18	-$12.62	0.130	-$1.64	-$157.10
2041	19	-$12.62	0.116	-$1.47	-$158.56

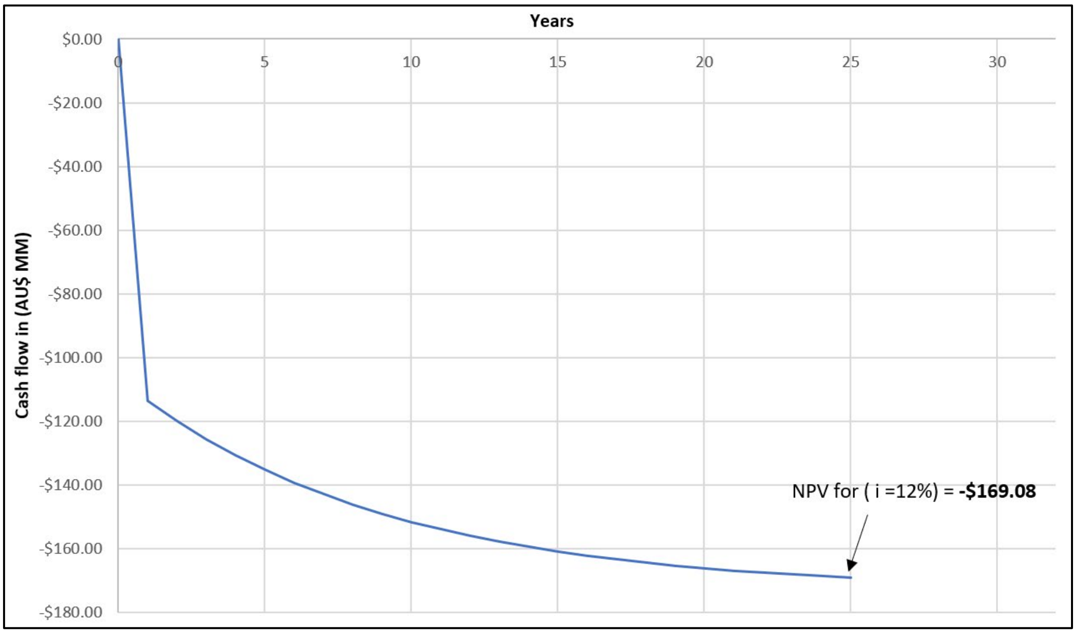
2042	20	-$12.62	0.104	-$1.31	-$159.87
2043	21	-$12.62	0.093	-$1.17	-$161.04
2044	22	-$12.62	0.083	-$1.04	-$162.08
2045	23	-$12.62	0.074	-$0.93	-$163.01
2046	24	-$12.62	0.066	-$0.83	-$163.84
2047	25	-$10.55	0.059	-$0.62	-$164.46



Scenario 3: AD WAS followed by 100% WAO of digestate

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D For i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1	$0.00	$0.00
2023	1	-$127.26	0.893	-$113.63	-$113.63
2024	2	-$7.99	0.798	-$6.37	-$120.00
2025	3	-$7.99	0.712	-$5.69	-$125.69
2026	4	-$7.99	0.636	-$5.08	-$130.77
2027	5	-$7.99	0.567	-$4.54	-$135.31
2028	6	-$7.99	0.507	-$4.05	-$139.36
2029	7	-$7.99	0.452	-$3.62	-$142.97
2030	8	-$7.99	0.404	-$3.23	-$146.20
2031	9	-$7.99	0.361	-$2.88	-$149.08
2032	10	-$7.99	0.322	-$2.57	-$151.66
2033	11	-$7.99	0.287	-$2.30	-$153.96
2034	12	-$7.99	0.257	-$2.05	-$156.01

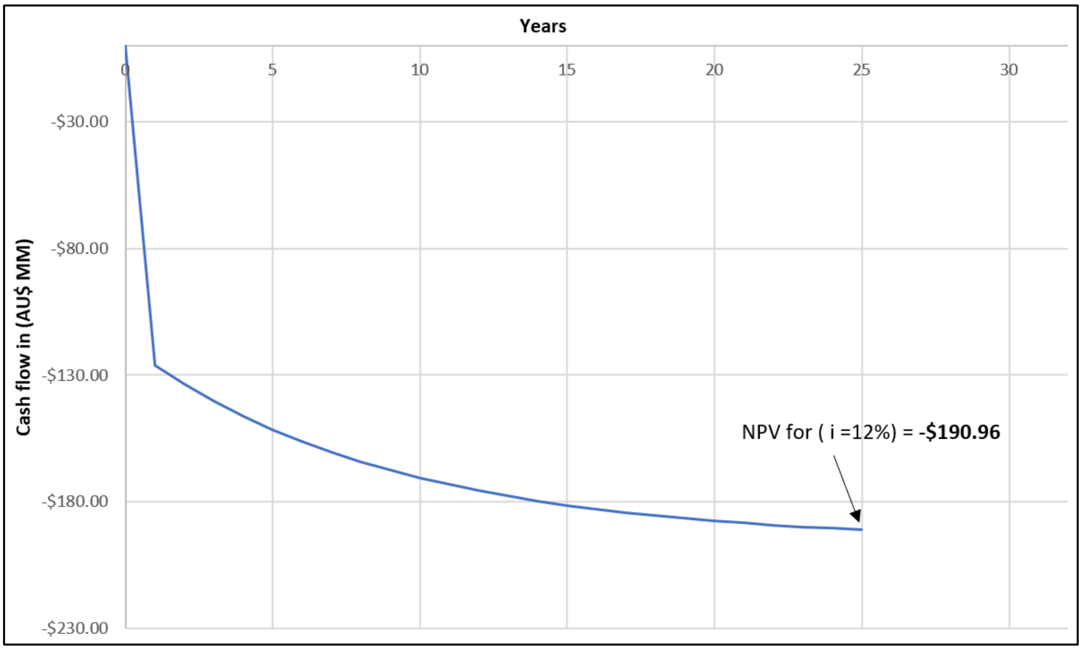
2035	13	-$7.99	0.229	-$1.83	-$157.84
2036	14	-$7.99	0.205	-$1.64	-$159.48
2037	15	-$7.99	0.183	-$1.46	-$160.94
2038	16	-$7.99	0.163	-$1.30	-$162.24
2039	17	-$7.99	0.146	-$1.16	-$163.40
2040	18	-$7.99	0.130	-$1.04	-$164.44
2041	19	-$7.99	0.116	-$0.93	-$165.37
2042	20	-$7.99	0.104	-$0.83	-$166.20
2043	21	-$7.99	0.093	-$0.74	-$166.94
2044	22	-$7.99	0.083	-$0.66	-$167.60
2045	23	-$7.99	0.074	-$0.59	-$168.19
2046	24	-$7.99	0.066	-$0.53	-$168.72
2047	25	-$6.09	0.059	-$0.36	-$169.08



Scenario 4: AD of WAS followed by 20% WAO of digestate

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D For i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1	$0.00	$0.00
2023	1	-$141.39	0.893	-$126.24	-$126.24
2024	2	-$9.33	0.798	-$7.44	-$133.68
2025	3	-$9.33	0.712	-$6.64	-$140.32
2026	4	-$9.33	0.636	-$5.93	-$146.25
2027	5	-$9.33	0.567	-$5.29	-$151.54
2028	6	-$9.33	0.507	-$4.73	-$156.27
2029	7	-$9.33	0.452	-$4.22	-$160.49

2030	8	-$9.33	0.404	-$3.77	-$164.25
2031	9	-$9.33	0.361	-$3.36	-$167.62
2032	10	-$9.33	0.322	-$3.00	-$170.62
2033	11	-$9.33	0.287	-$2.68	-$173.30
2034	12	-$9.33	0.257	-$2.39	-$175.70
2035	13	-$9.33	0.229	-$2.14	-$177.84
2036	14	-$9.33	0.205	-$1.91	-$179.75
2037	15	-$9.33	0.183	-$1.70	-$181.45
2038	16	-$9.33	0.163	-$1.52	-$182.97
2039	17	-$9.33	0.146	-$1.36	-$184.33
2040	18	-$9.33	0.130	-$1.21	-$185.55
2041	19	-$9.33	0.116	-$1.08	-$186.63
2042	20	-$9.33	0.104	-$0.97	-$187.60
2043	21	-$9.33	0.093	-$0.86	-$188.46
2044	22	-$9.33	0.083	-$0.77	-$189.23
2045	23	-$9.33	0.074	-$0.69	-$189.92
2046	24	-$9.33	0.066	-$0.61	-$190.53
2047	25	-$7.17	0.059	-$0.42	-$190.95



Appendix F5: NPV and IRR Calculations (Contd)

Scenario 1: Business as usual of anaerobic digestion of WAS

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D For i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1	$0.00	$0.00
2023	1	-$50.19	0.893	-$44.81	-$44.81
2024	2	-$4.86	0.798	-$3.87	-$48.68
2025	3	-$4.86	0.712	-$3.46	-$52.14
2026	4	-$4.86	0.636	-$3.09	-$55.23
2027	5	-$4.86	0.567	-$2.76	-$57.99
2028	6	-$4.86	0.507	-$2.46	-$60.45
2029	7	-$4.86	0.452	-$2.20	-$62.65
2030	8	-$4.86	0.404	-$1.96	-$64.61
2031	9	-$4.86	0.361	-$1.75	-$66.37
2032	10	-$4.86	0.322	-$1.56	-$67.93
2033	11	-$4.86	0.287	-$1.40	-$69.33
2034	12	-$4.86	0.257	-$1.25	-$70.58
2035	13	-$4.86	0.229	-$1.11	-$71.69
2036	14	-$4.86	0.205	-$0.99	-$72.68
2037	15	-$4.86	0.183	-$0.89	-$73.57
2038	16	-$4.86	0.163	-$0.79	-$74.36
2039	17	-$4.86	0.146	-$0.71	-$75.07
2040	18	-$4.86	0.130	-$0.63	-$75.70
2041	19	-$4.86	0.116	-$0.56	-$76.27
2042	20	-$4.86	0.104	-$0.50	-$76.77
2043	21	-$4.86	0.093	-$0.45	-$77.22
2044	22	-$4.86	0.083	-$0.40	-$77.62
2045	23	-$4.86	0.074	-$0.36	-$77.98
2046	24	-$4.86	0.066	-$0.32	-$78.30
2047	25	-$3.81	0.059	-$0.22	-$78.53

Appendix F5: NPV and IRR Calculations (Contd)

Scenario 2: TH of WAS followed by AD

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D For i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1.000	$0.00	$0.00
2023	1	-$86.10	0.518	-$44.61	-$44.61
2024	2	-$12.62	0.268	-$3.39	-$48.00
2025	3	-$12.62	0.139	-$1.76	-$49.75
2026	4	-$12.62	0.072	-$0.91	-$50.66
2027	5	-$12.62	0.037	-$0.47	-$51.13
2028	6	-$12.62	0.019	-$0.24	-$51.38
2029	7	-$12.62	0.010	-$0.13	-$51.50
2030	8	-$12.62	0.005	-$0.07	-$51.57
2031	9	-$12.62	0.003	-$0.03	-$51.60
2032	10	-$12.62	0.001	-$0.02	-$51.62
2033	11	-$12.62	0.001	-$0.01	-$51.63
2034	12	-$12.62	<0.001	$0.00	-$51.64
2035	13	-$12.62	<0.001	$0.00	-$51.64
2036	14	-$12.62	<0.001	$0.00	-$51.64
2037	15	-$12.62	<0.001	$0.00	-$51.64
2038	16	-$12.62	<0.001	$0.00	-$51.64
2039	17	-$12.62	<0.001	$0.00	-$51.64
2040	18	-$12.62	<0.001	$0.00	-$51.64
2041	19	-$12.62	<0.001	$0.00	-$51.64
2042	20	-$12.62	<0.001	$0.00	-$51.64
2043	21	-$12.62	<0.001	$0.00	-$51.64
2044	22	-$12.62	<0.001	$0.00	-$51.64
2045	23	-$12.62	<0.001	$0.00	-$51.64
2046	24	-$12.62	<0.001	$0.00	-$51.64
2047	25	-$10.55	<0.001	$0.00	-$51.64

NPV	$0.00
IRR	93%

Appendix F5: NPV and IRR Calculations (Contd)

Scenario 3: AD of WAS followed by 100% WAO of digestate

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D for i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1.000	$0.00	$0.00
2023	1	-$127.26	0.541	-$68.79	-$68.79
2024	2	-$7.99	0.292	-$2.34	-$71.13
2025	3	-$7.99	0.158	-$1.26	-$72.39
2026	4	-$7.99	0.085	-$0.68	-$73.07
2027	5	-$7.99	0.046	-$0.37	-$73.44
2028	6	-$7.99	0.025	-$0.20	-$73.64
2029	7	-$7.99	0.013	-$0.11	-$73.75
2030	8	-$7.99	0.007	-$0.06	-$73.81
2031	9	-$7.99	0.004	-$0.03	-$73.84
2032	10	-$7.99	0.002	-$0.02	-$73.86
2033	11	-$7.99	0.001	-$0.01	-$73.86
2034	12	-$7.99	0.001	$0.00	-$73.87
2035	13	-$7.99	<0.001	$0.00	-$73.87
2036	14	-$7.99	<0.001	$0.00	-$73.87
2037	15	-$7.99	<0.001	$0.00	-$73.87
2038	16	-$7.99	<0.001	$0.00	-$73.87
2039	17	-$7.99	<0.001	$0.00	-$73.88
2040	18	-$7.99	<0.001	$0.00	-$73.88
2041	19	-$7.99	<0.001	$0.00	-$73.88
2042	20	-$7.99	<0.001	$0.00	-$73.88
2043	21	-$7.99	<0.001	$0.00	-$73.88
2044	22	-$7.99	<0.001	$0.00	-$73.88
2045	23	-$7.99	<0.001	$0.00	-$73.88
2046	24	-$7.99	<0.001	$0.00	-$73.88
2050	28	-$6.09	<0.001	$0.00	-$73.88

NPV	$0.00
IRR	85%

Appendix F5: NPV and IRR Calculations (Contd)

Scenario 4: AD of WAS followed by 20% WAO of digestate

Year	Year Number	Annual Cash Flow $A_{CF}=A_{NCI}-A_{TC}$	Discount Factor, f_D For i=12%	Annual Discounted cash flow for i= 12% $A_{DCF}=A_{CF} \times f_D$	Cumulative discounted cash flow $\sum A_{DCF}$
2022	0	$0.00	1.000	$0.00	$0.00
2023	1	-$141.39	0.532	-$75.21	-$75.21
2024	2	-$9.33	0.283	-$2.64	-$77.85
2025	3	-$9.33	0.150	-$1.40	-$79.25
2026	4	-$9.33	0.080	-$0.75	-$80.00
2027	5	-$9.33	0.043	-$0.40	-$80.39
2028	6	-$9.33	0.023	-$0.21	-$80.60
2029	7	-$9.33	0.012	-$0.11	-$80.72
2030	8	-$9.33	0.006	-$0.06	-$80.78
2031	9	-$9.33	0.003	-$0.03	-$80.81
2032	10	-$9.33	0.002	-$0.02	-$80.83
2033	11	-$9.33	0.001	-$0.01	-$80.83
2034	12	-$9.33	0.001	$0.00	-$80.84
2035	13	-$9.33	<0.001	$0.00	-$80.84
2036	14	-$9.33	<0.001	$0.00	-$80.84
2037	15	-$9.33	<0.001	$0.00	-$80.84
2038	16	-$9.33	<0.001	$0.00	-$80.84
2039	17	-$9.33	<0.001	$0.00	-$80.84
2040	18	-$9.33	<0.001	$0.00	-$80.84
2041	19	-$9.33	<0.001	$0.00	-$80.84
2042	20	-$9.33	<0.001	$0.00	-$80.84
2043	21	-$9.33	<0.001	$0.00	-$80.84
2044	22	-$9.33	<0.001	$0.00	-$80.84
2045	23	-$9.33	<0.001	$0.00	-$80.84
2046	24	-$9.33	<0.001	$0.00	-$80.84
2047	25	-$7.17	<0.001	$0.00	-$80.84

NPV	$0.00
IRR	88%

RACE for 2030

www.racefor2030.com.au

Australian Government
Department of Industry, Science, Energy and Resources

AusIndustry
Cooperative Research Centres Program