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Design Project 488 (2019)

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	Section Description		Sizing and Costing			

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Document Submission Control				
Date	30/08/2019		Time	
Item 1	Evaluation of Options			
Item 2	Updated Process Description			
Item 3	Updated Process and Stream Highlights			
Item 4	Anaerobic Digester Design			
Item 5	Equipment Selection and Sizing			
Item 6	Capital Cost Estimation			
Item 7	Operating Cost Estimation			
Item 8	Economic Analysis			
Responsible	Student No.	Surname	Initials	Signature (Note 2)
Submitted				
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Note 2: The student who signs here was tasked by her/his group to submit the document-set on behalf of the group, but correct and timely submission of the relevant documents remains the responsibility of **the group and each of its members**. Penalties for late or incorrect submission will therefore be applicable to each group member.

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NOMENCLATURE

Symbols		
C_p°	Cost Estimation	\$ or R
D	Diameter	m
F	Equipment cost correction factor	-
I	Degree of Trough Filling	%
K_1,K_2,K_3	Costing Factors	-
K	Friction Coefficient	-
L	Length	m
$\dot{m}$	Mass flowrate	kg/s
Q	Energy	kJ
$\dot{Q}$	Power	kWh
S	Pitch	dm
Sg	Specific Gravity	g/cm^3
S_{yt}	Yield Strength	N/mm^2
t	Time	hr
V	Volume	m^3
$\dot{V}$	Volumetric flowrate	m^3/h
wt%	Weight percentage	%
x	Mass fraction	kg/kg
Subscripts & Superscripts		
L	Longitudinal	
M	Material factor	-
P	Pressure factor	
T	Temperature factor	-
Greek symbols		
η	Joint Efficiency	mm
ρ	Density	kg/m^3

ε	Degradation efficiency	-
δ	Solids recapture	-
ω	Polymer dose	mg/g
Acronyms		
CEPCI	Chemical Engineering Plant Cost Index	
CHP	Combined heat and power	
COD	Chemical oxygen demand	
CSTR	Continuously stirred tank reactor	
DAP	Diammonium Phosphate	
FSS	Fixed suspended solids	
MLVSS	Mixed Liquor Volatile Suspended Solids	
SA	Surface Area	
SCOD	Soluble chemical oxygen demand	
TCOD	Total chemical oxygen demand	
TSS	Total suspended solids	
VSR	Volatile solids reduction	
VSS	Volatile suspended solids	

1. BRIEF EVALUATION AND JUSTIFICATION OF OPTIONS

1.1 Feed Handling

The feed handling is dependent on the choice of anaerobic digester. A screen and grid chamber could be placed before the reactor to ensure no big solids and grid removal. It was decided to place a stationary bar screen in the wastewater inlet feed line to prevent any big solids or objects from entering the treatment process in the case of upstream errors. Also, a bar screen is a simple screen that would be sufficient for the purpose, without adding additional cost as would be the case if a vibrating screen was chosen. A grid chamber would not be required in the case where a CSTR is chosen, however, for UASB the case might be different.

A buffer tank is used to steady out any fluctuations in the influent flowrate. As 15% daily and seasonal fluctuations in the flowrate is expected, a buffer tank would be required. The steady out of the flowrate is also needed for when feedback control is applied.

A calamity tank would be required as an extra holding vessel in case of unexpected high influent or emergencies. This would be required to ensure a mitigation strategy and safe response plan.

1.2 Reactor considerations

A CSTR, or Continuously Stirred Reactor, consists of a vessel of with either a mechanical agitation mechanism (such as a paddle) or gas diffusers to ensure proper mixing. The vessel is typically constructed from steel, concrete or brick. The input and output flow to and from this reactor is intermittent, medium is pumped in while digested and excess waste is pumped out (S. M. Stronach, 1986). The biomass is present in the form of suspended flocculent (S. M. Stronach, 1986) which is kept in suspension by the agitator.

1.2.1.1 Advantages and Disadvantages

It has the advantage of being easier to design than its counterparts. (S. M. Stronach, 1986) The uniform distribution of nutrients, pH, substrate and temperature means that the operation is easily controlled. Sludge tends to float, resulting in sludge recycle difficulties (Gerardi, 2003). Vulnerable to shock loading.

1.2.1.2 Process Risks

The system is sensitive to mixing speed. Overmixing will result in the dispersion of the flocs necessary for operation while too little mixing will lead to compaction and resultant reactor failure (S. M. Stronach, 1986).

1.2.2 UASB

The Up-flow anaerobic sludge blanket reactor or UASB reactor consists of three distinct sections (IWA, 2018). At the bottom of the reactor lies the sludge bed, which consists of a high concentration of biomass; this is followed by the less dense sludge blanket which consists of small grains, flocs and gas bubbles (S. M. Stronach, 1986). Above the sludge blanket lies the biogas. The biomass in this reactor is in the form of granules of roughly 3-4mm. (IWA, 2018)

1.2.2.1 Advantages and Disadvantages

This reactor has a high efficiency rate and requires a low amount of energy and retention time compared to the CSTR contact process. Since the biomass is maintained at high concentrations, high biomass digestion rates are possible (S. M. Stronach, 1986), thus effective treatment of high COD strength waste is possible. (IWA, 2018) It is compact, has low operating cost (IWA, 2018). Since it is the most popular type of anaerobic digester (IWA, 2018) there is plenty of data available on it.

It has an extremely lengthy start-up period (IWA, 2018). It is more expensive than the CSTR (Gerardi, 2003). Tends to produce sulphurous compounds (S. M. Stronach, 1986). It releases bad odours (S. M. Stronach, 1986). It is not suited for TSS ratios higher than 500 $\frac{\mathrm{mg}}{\mathrm{L}}$ thus requires pre-treatment for these situations.

1.2.2.2 Process Risks

During the initial phases of operation it is prone to washout (S. M. Stronach, 1986). It is sensitive to feed interruptions (S. M. Stronach, 1986). Sludge will cake if agitation is not kept at an acceptable value.

1.2.3 Anaerobic Filter

1.2.3.1 Advantages

High biosolids retention (S. M. Stronach, 1986) can resist high shock loads (S. M. Stronach, 1986).

1.2.3.2 Disadvantages

Filter clogs often (S. M. Stronach, 1986). Solid accumulations tends to cause flow channelling (Gerardi, 2003). Lower loading capacities than the alternatives (Gerardi, 2003).

1.2.3.3 Process risk

Highly sensitive to pH (Gerardi, 2003). Filter blockages can halt digestion process (Gerardi, 2003).

1.3 Clarification and Effluent Treatment

1.3.1 Circular Sedimentation Tank

They can have either a central or peripheral feed. The feed flows out radially.

1.3.1.1 Advantages

It is economically compact, allows for easy sludge removal and has a high clarification efficiency (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012). It is more economical to build these than square tanks of the same capacity, but it takes up more ground space (Binnie, Kimber, & Smethurst, 2002). It is also easy to remove the sludge from the bottom of the tank (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012).

1.3.1.2 Disadvantage

However, these types of thickeners tend to require more piping for water and sludge transport to and from the tank than the rectangular configuration. It has a low tolerance to shock loads. Flocculation facilities, if needed, need to be added separately. It is largely intolerant of shock loads (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012). Peripheral loading eliminates the risk of short-circuiting (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012).

1.3.1.3 Process risks

The central feed configuration has a tendency to short circuit (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012). These arrangements are particularly susceptible to streaming (when thin influent does not mix with the main bulk of water).

1.3.2 Horizontal Flow Rectangular Sedimentation Basin

This thickener utilizes horizontal flow which distributes flocculated water over the basin at low velocity.

1.3.2.1 Advantages

This setup allows for two basins to be placed next to each other to save money by sharing a common wall (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012). These forms of thickeners are easy to build, operate and maintain. They are less susceptible to shock loads (Binnie, Kimber, & Smethurst, 2002). They have the ability to easily be upgraded by adding a plate settler module (Binnie, Kimber, & Smethurst, 2002). They scale-up easily and are best suited, economically, to large plants (Binnie, Kimber, & Smethurst, 2002).

1.3.2.2 Disadvantage

However, the sludge collection mechanism is not simple. There is a possibility of streaming occurring (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012). Density flows can result in the basin (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012). It usually requires a separate flocculation facility.

This setup has the risk of streaming (a particular form of short-circuiting) (Binnie, Kimber, & Smethurst, 2002).

1.3.2.3 Process risks

Settled solids in the corners are difficult to remove and may remain there where they increase dead volume.

1.3.3 Solids Contact Thickeners (Reactor thickeners)

This setup is usually used for plants treating raw water with a low solids concentration. This setup aims to combine the flocculation, mixing and sedimentation in one unit (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012).

1.3.3.1 Advantages

It has the advantage of being more compact, therefore more economical, than the alternatives. It outputs an effluent that is soft and has a low turbidity (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012).

1.3.3.2 Disadvantage

It is, however, sensitive to, not only shock loads, but changes in flowrates as well as temperature changes (Crittenden, Rhodes, Hand, Howe, & Tchobanoglous, 2012).

1.3.3.3 Process risks

Prone to short-circuiting and streaming

1.4 Sludge Dewatering

1.4.1 Plate and Frame Filter Press

In this setup two plates with a cloth in between plates. These plates press together and the sludge is pumped under high pressure between the fibers of the cloth. The cloth becomes caked with solids, thus becoming more efficient at retaining solids over time. About 30-50% solids content in the cake can be achieved (Binnie, Kimber, & Smethurst, 2002) (CRITERIA UNIFIED FACILITIES, 2003).

1.4.1.1 Advantages

High solids content of the cake (Binnie, Kimber, & Smethurst, 2002).

1.4.1.2 Disadvantages

High labour cost and frequent filter cloth replacement (Binnie, Kimber, & Smethurst, 2002). High capital and operation maintenance costs (Binnie, Kimber, & Smethurst, 2002). Large amounts of chemicals are required to operate effectively (Binnie, Kimber, & Smethurst, 2002).

1.4.1.3 Process Risks

Filter times gradually become longer as the filter cloth becomes more worn. This will result in a larger energy consumption. These cloths need to be replaced manually, thus cycle time becomes dependent on the speed of manual labour.

1.4.2 Centrifuges

Separation is achieved using high speed centrifugal motion. 35 – 50% solids concentration can be achieved with the correct dosing chemical (Binnie, Kimber, & Smethurst, 2002). The solid bowl type is the most efficient for wastewater purposes. Advantage: Highest potential solids concentration output (Binnie, Kimber, & Smethurst, 2002) Disadvantage: High capital, operation and maintenance costs (Binnie, Kimber, & Smethurst, 2002).

1.4.3 Drying Bed

This consists of a pit with sand on the bottom. Sludge is pumped to these pits and the water is allowed to evaporate into the atmosphere (Binnie, Kimber, & Smethurst, 2002). Advantage: Cheap, easy sludge removal and transportation (Binnie, Kimber, & Smethurst, 2002). Disadvantage: Labour intensive operation. Inefficient sludge dewatering. The performance is highly dependent on weather. Requires a large amount of land (Binnie, Kimber, & Smethurst, 2002). Risk: Rain or high humidity conditions will increase the drying time by a large factor.

1.4.4 Steam Boiler

Boilers are closed vessels used for the generation of steam from the combustion of natural gas (Jaya, 2011). Generally the energy released would be used to produce steam that will subsequently be used for the generation of electrical power. Minimum heat loss and maximum heat absorbed are characteristics of a good boiler performance, but these are normally influenced by process requirements, economics and safety. Advantage: For steam boilers the rate of steam generation is greater than that for fired tube boilers of the same furnace capacity (Jaya, 2011). Also, to procure a steam boiler is generally less expensive. Disadvantage: Relative to fired tube heaters greater skill is needed to operate a steam boiler. Due to its complexity (Jaya, 2011) also claimed that steam boilers present a higher risk of exploding while operating.

1.5 Biogas Handling

Within the anaerobic digester methanogens breaks downs biodegradable material to form biogas. The biogas contain various dissolved gases of which methane, carbon dioxide and hydrogen sulphide is significant. Due to it corrosive nature it is preferential to removes as much of the hydrogen sulphide to protect gas handling equipment (Erdirencelebi & Kucukhemek, 2018). To ensure that the concentration of hydrogen sulphide is sufficiently reduced, the addition of ferric chloride is considered. In theory the addition of ferric chloride should removes close to 100 % of the dissolved hydrogen sulphide such that it may be assumed to be negligible compared to methane and carbon dioxide production.

1.6 Dosing Station (Chemical Dosing)

1.6.1 Calcium Hydroxide [$Ca(OH)_2$] Dosing

For mesophilic bacteria to thrive within the anaerobic digester an optimum pH value of 7.2 is desired. Due to the occurrence of continuous acidification calcium hydroxide is dosed to bring about sufficient neutralization. Economically, hydrated lime is preferred over slaked lime because it costs less per kg. In addition to its price hydrated lime is also the preferred choice as it removes dissolved minerals and heavy metals much more effectively (Jackson, 2016). According to (Jackson, 2016) the addition of hydrated lime also generally improves the water quality which is ultimately the objective of this process. Assuming continuous operation hydrated lime should be dosed via D-1 at a mass flowrate of 4109.2 kg/d to achieve the desired effect. To ensure a rapid a response the hydrated lime solution is dosed in-line just before the anaerobic digester.

1.6.2 Urea (CH_4N_2O) Dosing

To ensure adequate microbial growth occurs the nitrogenous oxygen demand must be met such that methane is produced at a rate that is justifiable. Generally nitrogen would be fed as ammonia, nitrate or nitrate nitrogen (Qasim & Zhu, 2018). At a nitrogenous demand of 153.9 kg/d urea has to be fed at 263.4 kg/d via D-2. Like with the hydrated lime a portion of the wastewater influent stream is used to make up a suspended solids solution before it is dosed in-line before being fed into the anaerobic digester.

1.6.3 Diammonium Phosphate [$(NH_4)_2HPO_4$] Dosing

In addition to nitrogen microbial organisms also needs phosphorus to grow such that enough methane is produced from the breakdown of biodegradable material (Qasim & Zhu, 2018). By mass diammonium phosphate (DAP) has a nitrogen and phosphorus content of 21.2 and 23.5 %, respectively which makes DAP suitable for both nitrogen and phosphorus addition. Although mono-ammonium phosphate has a higher phosphorus content than DAP, DAP is still preferred for its higher nitrogen content to supplement the any nitrogen deficiencies should it deviate from specifications. To minimize the money spent on procurement of process water separate solutions of hydrated lime, urea and DAP is made using water mass flowrates of 43260, 904.3 and 138.3 kg/d, respectively.

1.6.4 Polymer Dosing

1.6.5 Ferric Chloride ($FeCl_3$) Dosing

The addition of ferric chloride to the anaerobic digester serves as a preventative measure to significantly reduce the formation of hydrogen sulphide such that it may be assumed to be negligible. Since the concentration of hydrogen sulphide is reduced within the biogas it eliminates the need for gas scrubbing which ultimately reduces the costs associated with equipment procurement, and in addition protects equipment from being exposed to a corrosive gas. With an average daily flowrate of 17870 m^3/d (Aslanidou, Lykakis-Simantiris, Kotsifaki, & Katsivela, 2018) managed to almost remove all the dissolved hydrogen sulphide whilst adding ferric chloride at 296.5 mg per L of wastewater. Using this as recommended value and downscaling to the specified flowrate 33 kg/d of $FeCl_3$ would be needed to sufficiently reduce the formation of hydrogen sulphide.

1.7 Pump Selection and Placement

1.7.1 Centrifugal Pumps

Centrifugal pumps refers to pumps that transport fluid by means of a rotating impeller (PumpScout, 2019). While centrifugal pumps are mostly used to transfer liquids, some centrifugal pumps have the ability to transfer sludge with a low viscosity and a small amount of suspended solids. (Sintech Pumps, 2018) These pumps can produce the highest flowrates compared to all pump types and are lower in cost compared to positive displacement pumps. Can transport clean and dirty liquids (PumpScout, 2019). A disadvantage of centrifugal pumps is that they have a bigger shear rate, which has a negative impact on the treatment process. The shearing generates a higher sludge surface area, causing the solid particles to be more distributed in the liquid. This results in a sludge that is more difficult to dewater. (Sintech Pumps, 2018). A risk to consider is that these pumps cannot pump liquids that contain air or vapours, this will lead to damage of the pump impeller (PumpScout, 2019).

1.7.2 Positive Displacement Pump – Peristaltic Pump

Peristaltic pumps fall under the category of rotary pumps. The pump consists of a roller that push the liquid through a tube as it squeezes the tube with the rotary movements (PumpScout, 2019).

Peristaltic pumps are commonly used for dosing chemicals in a process plant. These pumps can transport viscous liquid and liquids requiring a low flow rate. As the liquid flows through a tube in the pump, there would not likely be any leakages, thus the pump does not require a seal (PumpScout, 2019). A disadvantage of these pumps is that the inner tube degrades over a period of time and needs to be replaced periodically. A risk to consider is that most positive displacement pumps do not have a shutoff head, making it unable to continue operation when a valve on the discharge side is closed. (Positive Displacement Pump Basics, 2015)

1.7.3 Positive Displacement Pump – Diaphragm Pump

The diaphragm pump has a compressed air operated diaphragm that is used to transport the fluid. These pumps are mostly used to transport. This pump is known for its simple operation and for pumping liquid with high solids, creating a uniform, constant sludge blanket (Sintech Pumps, 2018). It is able to run dry without getting damaged (PumpScout, 2019). A disadvantage of these pumps is that in order to operate a diaphragm pump, compressed air needs to be available on the treatment site. The addition of an air compressor, in the case that it is not already available, would increase the pump operation significantly (Sintech Pumps, 2018). A risk to consider is that most positive displacement pumps do not have a shutoff head, making it unable to continue operation when a valve on the discharge side is closed. (Positive Displacement Pump Basics, 2015)

1.7.4 Positive Displacement Pump – Progressive Cavity Pump

A progressive cavity pump has a turning, single-threaded rotor inside a double-threaded elastic stator. This movement results on the formation of a cavity that is used to transport the liquid (PumpScout, 2019). These pumps are really useful in transporting sludge as it has a high efficiency and produce almost no shear on the sludge material (Sintech Pumps, 2018). A disadvantage of these pumps is that the repair and maintenance of these pumps can become really complicated, adding to additional costs. The overall footprint of these pumps are also bigger compared to other positive displacement pumps (Sintech Pumps, 2018). A risk to consider is that when replacing this pump with another type, changes to the piping system would be required due to its big footprint (Sintech Pumps, 2018).

1.7.5 Positive Displacement Pump – Rotary Lobe Pump

Lobe pumps consist of two rotating lobes that mesh with another as it rotates. Timing gears are used to ensure that the lobes do not come into contact with each other. This is desired when viscous or fragile solids are transported (PumpScout, 2019). Rotary Lobe pumps are popular in the transportation of fluids with a high viscosity. The shear produced by this pump is also very low, making it ideal for sludge transportation. The overall footprint of this pump is small compared to other positive displacement pumps and maintenance is easily performed and not highly expensive (Sintech Pumps, 2018). A disadvantage to these pumps is that the initial cost of equipment can be expensive (Sintech Pumps, 2018). A risk to consider is that most positive displacement pumps do not have a shutoff head, making it unable to continue operation when a valve on the discharge side is closed. (Positive Displacement Pump Basics, 2015)

1.7.6 Final Selection

For the wastewater influent and effluent, it was decided to use a centrifugal pump. The solids content is very low in these streams, and the correct centrifugal pump would be able to pump this water with ease. These streams also require high flow rates, for which a centrifugal pump is desired.

For any stream transporting sludge, it was decided to use a positive displacement pump, as these streams are more viscous and abrasive. The recycled sludge also needs minimal shear and for this requirement, a rotary lobe pump was selected. This type of pump will pump the sludge with ease, without damage to the pump and would protect the biomass while being transported.

For the dosing station, it was decided to use peristaltic positive displacement pumps, as these pumps can handle viscous, low flow rates and is generally used when chemicals are dosed.

In terms of pump placement, a pump was required at the discharge of each vessel, except for stream 5 where the discharge of R-101 is transported to the thickener. The slightly pressurized reactor and the head produced from the liquid level in the reactor, provide enough pressure for the outlet stream to be transported naturally to V-111. In all other cases it was calculated that a pump would be needed to overcome the destination head.

1.8 Conveyor System Selection and Placement

1.8.1 Conveyor Belt

A conveyor belt is normally used when bigger solids are transported and when material are brought into the site from outside the battery limits or vice versa. These conveyors are usually open transport and thus cannot transport powers as it would be blown off by natural wind. A conveyor belt was selected for transport of the final sludge to the waste skips.

1.8.2 Screw Conveyor

Screw conveyors are normally used when powders and small granular material needs to be transported. These conveyors are popular for dosing systems as it can accurately transport small amounts of fine powders. Screw conveyors were selected for all powder and granule transport from the hoppers to the mixing tanks at the dosing station.

1.9 Heat Exchanger Placement

For the wastewater influent it was decided that two shell-and-tube heat exchangers be placed in series to facilitate some degree of heat integration. Since the wastewater leaves at about 30 °C it would be considered economical to use its energy in E-101 to heat the wastewater influent stream. However since typical river water is discharged at 22.8 °C as claimed by (Dallas, 2008) the wastewater influent undergoes a small temperature increase. The rest of the energy needed to get the wastewater influent to the operating temperature of the anaerobic digester would be provided by steam that passes through the tubes of E-102. Despite the fact that 1.3 t/d of methane is combusted not enough steam is produced therefore to supplement the energy requirements additional steam would have to be produced. E-103 is an electric powered heat exchanger that produces 5.7 t/d of steam in addition to the 26.4 t/d already produced by the combustion of methane formed.

2 UPDATED PROCESS DESCRIPTION

2.1 Screen

The process feed water containing 8880 mg TCOD/L and 1695 mg TSS/L, is received from the plant sump and transported to the wastewater treatment plant for processing. The feed water is pumped through stream 1 by means of a centrifugal pump, P-108 A/B, at 17 ^{o}C and 1 bar to the coarse screen, S-101, at a flowrate of 1410 m^3/d. This screen is used as a preventative measure to ensure that no big solids and objects are transported to the treatment plant, in case of any upstream errors. It is assumed that all particles within stream 1 is acceptably small such that no material losses occur over the separation device. The mass flowrate of water, TSS and TCOD are 1450.2, 2.4 t/d and 12521 kg/d. Other important characteristics of the influent stream is that it has a pH value of 6, the available alkalinity is 412 mg/L as $CaCO_3$ and the ammonia and phosphorus concentrations correspond to values of 22 and 7 mg/L.

2.2 Buffer Tank

Stream 2 enters TK-102 with the contents as stipulated above by means of a centrifugal pump, P-108. Stream 3 which is the effluent from TK-102 passes through E-101 where the contents is heated up to 20 ^{o}C by utilizing the heat of the final water effluent as the hot stream in E-101. After E-101 the heater water passes through another heat exchanger, E-102, where the contents is further heated to 37 ^{o}C in stream 4 before entering unit R-101, the anaerobic digester, by means of P-110. The high pressure steam produced in the biogas handling section is used as utility in E-102.

2.3 Calamity Tank

A calamity tank, TK-103, is provided for unusually high influent where some of the wastewater needs extra, temporary storage or any emergency in the plant where R-101 needs to be drained. Stream 2 can be directed to flow directly into TK-103 instead of TK-102. The discharge of TK-103 flows into stream 3 where it is recirculated into the treatment system by means of a positive displacement pump, P-109.

2.4 Anaerobic Digester

Within unit R-101 some of the biomass is converted into biogas when anaerobic bacteria is utilized to breakdown the degradable solids suspended within the solution. The biogas that forms leaves via stream 17 and consist of 65 % methane by volume with the remainder being carbon dioxide

(Tehobanoglous, Burton, & Stensel, 2003). A foam trap is installed on stream 17 to capture any unwanted formation of foam that leaves the reactor at the gas outlet. Accompanied by the reactor is an upstream chemical dosing station that is responsible for maintaining the ideal environment for biological growth of methanogenic bacteria. All chemicals are dosed in-line at stream 4 just before R-101, to ensure quick control response as the pH and nutrient measurements are made within R-101. Inside R-101 the dissociation of carbonic acid continuously takes place to such that a desired pH of 7.2 can be achieved and slaked lime is dosed at 47369 kg/d for this purpose. The desired pH was chosen to be 7.2 since (Poh, Gouwanda, Mohan, Gopolai, & Tan, 2016) claimed that values between 6.8-7.2 is optimal for biological growth of methanogens that converts biomass into methane. To ensure that there are enough macro-nutrients available both urea and diammonium phosphate need to be added in amounts equal to 263.4 and 78.94 kg/d, respectively. To prevent the formation of hydrogen sulphide, ferric chloride is also added to R-101. Micronutrient-solution is also available at the dosing station in the case that the micronutrient levels in R-101 drop and needs to be adjusted. The reactor effluent exit via stream 5 where it flows to V-111.

2.5 Thickener

Stream 5 which has a solids content of about 1.16 % by mass enters unit V-111 where a large portion of the solids is removed before discharging a clearer effluent, stream 6, which will undergo secondary treatment to remove some of the solids still in solution. Of the 21.7 t/d of solids that enters V-111, as much as 19.5 t/d reports to stream 8 which is the underflow, by means of P-113, while the remainder report to stream 6, the overflow by means of P-112. Both P-112 and P-113 are positive displacement pumps. Taking these numbers into account V-102 has an estimated solids removal efficiency of 90%. Polymer is dosed to stream 5 just before V-111 to ensure efficient settling of solids in V-111. Stream 8 which is split further into stream 9 and stream 11 has a solids flowrate of 1.9 and 17.6 t/d, respectively. The sludge contents of stream 11 is then recycled back into R-101 via stream 15 whereas the contents of stream 9 is further treated using unit C-101, a centrifuge.

2.6 Rotary Dum Filter

Unit F-101 is used to remove some of the solid remaining in the solution such that stream 7 and 18, the effluent, can be safely discharged to municipal sewer. With regards to solids a further amount is removed which corresponds to a mass flowrate of 0.3 t/d that reports to stream 12. The contents of this stream mixes with the downstream underflow from C-101, the centrifuge. Stream 7 which is also the final effluent has a mass flowrate of 1437 t/d of which solids contribute just 0.021% while water contribute the remainder of 99.8 %. The TSS and TCOD concentration within this stream is 200 and 1000 mg/L, respectively.

2.7 Centrifuge

Just like F-101 is used to further decrease the solids concentration of stream 6, C-101 is used to decrease the solids concentration of stream 9 coming from V-102. Stream 9 which has a mass flowrate of 39.7 t/d with a solids content of 4.8 % splits into underflow to overflow solids ratio of 0.188 which suggests that less solids is retained in stream 13 as opposed to that discarded within stream 10. Stream 13, the underflow, has a solids mass flowrate of 0.3 t/d as opposed to a much higher liquid mass flowrate of 32.9 t/d. The contents of stream 13, transported by a positive displacement pump P-115, combines with that of stream 12, transported by a positive displacement pump P-114, to yield stream 14 while the contents of stream 14 combines with that of stream 11 to yield stream 15.

2.8 Biogas Handling

At the biogas handling section, the biogas from stream 17 enters a gas boiler, H-101, where it is combusted to produce high pressure steam form a water circulating stream. To minimize costs, the produced steam is used as utility in E-102 to heat up the influent wastewater to 35 ^{o}C. However, the produced steam from H-101 is not enough for this purpose and hence additional steam needs to be produced. An electric heat exchanger E-103 is used to produce high pressure steam from process water. The combined steam is transported to E-102 by means of a blower, B-102. The CO_2 produced in the gas boiler is flared to the atmosphere.

3 UPDATED PROCESS AND STREAM HIGHLIGHTS=

The key performance parameters can be seen in Table 1 the majority of these performance parameters are subject to change with appropriate testing of a pilot plant in order to determine the various performance parameters on other performance parameters such as the actual efficiency of degradation of the effluent water. Therefore, previous work and literature was used to come to the following parameters.

Table 1: Process performance parameters

Performance Parameter	Value	Unit
COD Loading Rate	12520.8	kg COD/d
Anaerobic Reactor Volume (excluding headspace)	4598	m^3
Volumetric Loading Rate	2.7	kg COD/m^3.d
Biomass Activity (Food to Biomass Ratio)	0.25	kg COD/kg VSS.d
TSS Concentration in Reactor	11981	mg/L
VSS Concentration in Reactor	8000d	mg/L
Total Biomass in Reactor	36784	kg
Sludge Retention Time (SRT)	28.9	days
Hydraulic Retention Time (HRT)	3	days

The COD loading rate is the total amount of COD that enters the process per volume of the reactor as such is used as a basis in order to size the reactor volume. According to Qasim (Qasim & Zhu, 2018) the appropriate volumetric loading rate should be between 2-5 kg ,This can be seen in table 8 in Appendix C, COD/m^3d leaving the appropriate volume of the reactor to lie within the range of 2500-6200 m^3. The reactor was sized to deal with this load of COD was 5076 m^3 so as to achieve a Solid retention time and Hydraulic retention time within the ranges set out by Qasim (Qasim & Zhu, 2018). The current VSS Concentration within our reactor, which is just below the concentration bonds in which an anaerobic.

The general inputs and outputs can be seen in Table 2, here one can see that the content within the waste stream of the distillery is reduced by the separation of the undesirable compounds into a useful biogas stream final effluent stream below the battery limits as well as a stream high in solids content.

Table 2: Process input/output streams

Inlet		**Outlet**	
Stream number	**Mass flow (t/d)**	**Stream number**	**Mass flow (t/d)**
1	1453	10	7
		17	9
		18	1437
Total mass flow in:	1453	**Total mass flow out:**	1453

Table 3: Process input/output stream detail

Stream number	**1**	**10**	**17**	**18**
Total Volumetric Flowrate (m3/d)	1410	6.1	2789	1395
Liquid Volumetric Flowrate (m3/d)	1408	4.7		1394.8
Solid Volumetric flowrate (m3/d)	2	1.4		0.2
Total Mass Flowrate (t/d)	1452.6	6.5	9.2	1437
Liquid Mass Flowrate (t/d)	1450.2	4.9		1436.7
Solid Mass Flowrate (t/d)	2.4	1.6		0.3
Concentrations				
TCOD (mg/L)	8880	414395		1000
SCOD (mg/L)	6246	1863		760
TSS (mg/L)	1695	266954		200
FSS (mg/L)	339	88709		66
VSS (mg/L)	1356	178245		134
Loadings				
TCOD (kg/d)	12521	1961	9164	335
SCOD (kg/d)	8807	11	7735	1060
TSS (kg/d)	2390	1625		279
FSS (kg/d)	478	540		93
VSS (kg/d)	1912	1085		186

4 ANAEROBIC DIGESTER DESIGN CONSIDERATIONS

Two important parameters when it comes to the overall efficiency of the reactor is the solids retention time as well as the hydraulic retention time. The importance of these parameters are due to the degradation of the COD that they contain, the hydraulic retention time is set to ensure degradation of the soluble COD and the Solid retention time is set in order to degrade a sufficient portion of the particulate COD, this can be seen in appendix C.

The reactor within the mass balance was designed in order to have operate at the limit of Table 7 in appendix C, where the concentration in the reactor was set at 8 kg/m^3 this will ensure a high Solid retention time whilst minimizing the size of the reactor, which would help to ensure minimizing the cost of the plant. The typical operating conditions of anaerobic digesters can be seen in table 7 and the reactor was thus sized whilst keeping these parameters within the recommended or typical operating ranges.

The reactor was sized within the parameters set by Qasim (Syed Qasim, 2018) wherein the reactor should be within the following specs displayed in table XX. The volume of the reactor was sized from the calculations in appendix C wherein the Volatile solids reduction can be used to calculate the Solids retention time, the volatile solids reduction was calculated from the Particulate removal efficiency from section 1.

The reactor was then sized geometrically to ensure adequate mixing in dimensions set out by tank 1 in Maroney's paper (Meroney, 2008). This was sized by keeping the same side wall height to diameter ratio as well as cone height to diameter ratio as tank 1 in Maroney's paper. By having the same dimensions as this tank the reactor would have similar mixing characteristics as well as duties for the draft tubes. Tank 1 was selected as this configuration utilizes external draft tubes, this would allow for maintenance of tubes to be completed quickly to ensure minimal distribution to the process. As well as being within the geometrically set parameters set by Qasim (Syed Qasim, 2018). These parameters were that the design of anaerobic digesters were sized between 6 and 40m in diameter, the conical slope was between 4 and 6 meters horizontal to 1 meter vertical and the side wall height is between 7.5 to 15m. (Syed Qasim, 2018). A summary of the reactor dimensions can be seen in appendix C along with the tank to which the digester was based.
Maroney's paper displays that these external draft tubes provide sufficient mixing within the reactor by being with in the ranges of the mixing parameters set by Qasim (Syed Qasim, 2018) these parameters are that the pumps have a velocity gradient of mora than 50 s^{-1} and a digester volume turnover rate of 20-30min. Assuming that the fluid within the reactor has the same viscosity and that the pumps have the same efficiency the reactor will operate with the a velocity gradient of 94 s^{-1} and a DVTT of 30 minutes, these values are calculated in Appendix C. As the scum formation within the reactor is difficult to determine the over specification of the mixers should allow for the problems relating to scum formation to be mitigated. This scum formation was then assumed to be removed from the top of the thicker to prevent build up (S. M. Stronach, 1986).

5 EQUIPMENT SELECTION AND SIZING

5.1 Feed Handling

It was decided to place a stationary bar screen in the wastewater inlet feed line to prevent any big solids or objects from entering the treatment process in the case of upstream errors. The material of construction for the bas screen is carbon steel as this has resistance to wear.

A buffer tank is used to steady out any fluctuations in the influent flowrate. As 15% daily and seasonal fluctuations in the flowrate is expected, a buffer tank would be required. The steady out of the flowrate is also needed for when feedback control is applied. The material of construction for the buffer tank is carbon steel as this provides some resistance to corrosion but is also resistant to abrasion as the liquid containing solids flow in and out at a high rate and would be continuously stirred in the buffer tank.

A calamity tank would be required as an extra holding vessel in case of unexpected high influent or emergencies. The material of construction for the calamity tank is also carbon steel.

5.2 Reactor

In order to minimize the space the effluent treatment plant would occupy due to the high rate in which the influent is being introduced, only high rate anaerobic digesters were considered for the main piece of equipment within the process. As there are relatively small fluctuations in the influent to our proposed plant the reactor and subsequent plant was designed to operate at the average follow rate of the liquid as well as the solids in the feed. In selecting the type of reactor the characteristics of the effluent must be analyzed. Of particular interest in our feed is that the Total Suspended solids content of 1.69 g/l which can fluctuate daily by 15%.

There are several high rate anaerobic reactors used in industry, the three considered were a continuously stirred tank reactor (CSTR) within an anaerobic contact process, an Up-flow anaerobic sludge blanket reactor (UASB) and an Anaerobic Filter. Molleta (Moletta, 2005)suggests that the use of both UASB and Anaerobic filters is only recommended when there is no or very low concentration of total suspended solids within the feed. The influent TSS concentration is more than 3 times the recommended TSS concentrations for the aerobic filter and UASB (500 mg/L) (S. M. Stronach, 1986), thus pre-treatment is necessary for both those reactors. These additional processes would need to lower the solids concentration to acceptable operation levels in order to maintain smooth operation. This would lead to additional units being utilized for pre-treatment and lead to less of the influent to the process being anaerobically digested. This would thus lead to more sludge production along with additional units within the process.

The solution to this would be to use a CSTR reactor within an anaerobic contact process as this type of reactor is preferred when dealing with high concentrations of suspended solids (Moletta, 2005). This type of reactor deals with high concentrations of suspended solids which allows for all of the influent to travel through the reactor without pre-treatment. This gives the opportunity for all of the biodegradable COD to be degraded which will allow for less total sludge to be produced with the same separation efficiency over the overall process. As a result more biogas is available.

It is cheaper than the alternatives, while being suited for the influent to be treated. The anaerobic filter is prone to blockages, thus requires more frequent maintenance (S. M. Stronach, 1986). The anaerobic filter would be justified if the effluent had strict discharge values, however in this design it is not the case.

5.3 Thickener

A thickener is used in order to increase the concentration of VSS recycled back to the reactor. A thickener is used in order to decrease the amount of water recycled and thus a smaller reactor would be required in order to generate the same biomass within the reactor. The Thickener was sized from the settling characteristic tests presented in Shea (T.G. Shea, 1974) carbon steel would be used as the material of construction as it is a cheaper alternative to stainless steel. The use of carbon steel can be justified as it was assumed that all the biogas left in the reactor and this material can be left unprotected if it is constantly submerged in a slurry according to Penn state University (University, 2019).

The circular thickener was selected due to the cheap building cost compared to its rectangular counterpart. It is easier to recover the sludge than in the horizontal tanks. The rectangular tank tends to collect sludge in the corners of the reactor leading to dead volume and lost sludge. However, both these tanks require separate flocculation facilities increasing the capital required to run the plant. The circular reactor thickener has the advantages of the circular thickener while removing the need of a separate dosing tank.

The circular thickener with a horizontal inlet port was selected due to the easy sludge removal and high efficiency.

5.4 Vacuum drum filter

The vacuum drum filter is needed in due to the primary objective of the thickener being to recycle the VSS back to the reactor. Therefore, the thickener shouldn't be sufficient in order to lower the TSS concentration down to within the battery limits. A Vacuum drum filter would then be an appropriate piece of technology in order to reduce the TSS concentration in the effluent.

5.5 Centrifuge

The centrifuge (C-101) was selected due to the low operating cost compared to the filter press and the significantly higher efficiency compared to the drying bed, despite the drying bed being cheaper. It is also significantly more reliable than the drying bed as its performance it is not dependent on the weather.

5.6 Dosing station

5.6.1 Hoppers/ Tanks

V-101, V-102, V-103 and V-105 are storage tanks to keep an inventory of slaked lime, urea, DAP and cationic polymer, respectively. Slaked lime, urea, DAP are powdered chemical compounds used to regulate the conditions for optimal biological growth within R-101 whereas polymer is used to facilitate flocculation. Since of these compounds are available in powdered form the use of hoppers as storage tanks is justified. Based on hydrostatic pressures of 2.5, 1.4, 1.3 and 1.1 bar(a) for V-101, V-102, V-103 and V-105, stainless steel should be used as the material of construction as suggested by (Van Wyk, 2014). Considering that TK-101, TK-102 and TK-103 are pressurized vessel, the same applied to them. The dosing station is elevated 5m above ground level. This helps to overcome the high destination head of dosing to the reactor.

5.6.2 Mixers

The impeller were all selected due to the cost effectiveness compared to draft tubes which are more suited to larger tank diameters (Gerardi, 2003). The specific impeller type selected for each vessel (V-106 to V-110),

5.7 Heat Exchangers

Two separate shell-and-tube heat exchangers are placed in series to facilitate heat integration. E-101 uses the energy from the wastewater effluent to heat up the wastewater influent to 20.4°C. Although the temperature of the wastewater influent only increase by 3.4°C it is justifiable considering that the flowrate of the wastewater influent is much greater than the flowrate of the wastewater effluent. Due to the abrasive nature of the wastewater influent the wastewater influent should be transported through the shell made of stainless steel whereas the cleaner effluent would be transported through the tubes.

E-102 is a secondary heat exchanger that is used to get the wastewater influent to the operating temperature of R-101. To ensure that the wastewater influent absorbed enough energy to be heated from 20.4 to 35 °C a steam flowrate of 32 207.9 kg/d is required at 400 °C. Again, since the wastewater influent contains dissolved solids and heavy metals of which many may be corrosive the shell-side should be fabricated from stainless steel. It would acceptable to send the high pressure steam through the tube of the heat exchanger.

5.8 Transport

5.8.1 Centrifugal Pumps

(P-101 A/B, P-108 A/B, P-112 A/B, P-116 A/B, P-117 A/B)

P-101 is used to transport a fraction of the wastewater influent to the dosing station where it is used to make up the dosing chemical solutions. P-108 is used to transport the wastewater influent from the sump to the anaerobic treatment system and P-112 is used to transport the thickener overflow effluent to the drum filter. The final water effluent is circulated through E-101 by means of P-116 and the condensate from E-102 is circulated to H-101 by means of P-117.

The process water that circulates through E-102 is transported by a centrifugal pump as it is a good choice for streams with a high flow rate and no solids content. As the solids concentration in the wastewater influent and effluent is really low (below 2%), a centrifugal pump can be used and was selected since it would have a lower cost compared to a positive displacement pump. Some companies, like Sintech Pumps, manufacture specific centrifugal pumps that can tolerate a low solid concentration.

The material of construction for the centrifugal pumps, including casing and impeller, is carbon steel. The chosen impeller is type K. This material have some corrosion resistance but also have good resistance to wear due to abrasion, which is required when pumping liquid with a solids content (Material selection for wastewater pumps, 2013).

5.8.2 *Peristaltic Positive Displacement Pumps*

(P-102 A/B, P-103 A/B, P-104 A/B, P-105 A/B, P-106 A/B, P-107 A/B)

These pumps are used for dosing the required chemicals. P-102 transports the slaked lime to R-101, P-103 transports the urea solution to R-101, P-104 transports the diammonium phosphate to R-101, P-105 transports the micronutrients to R-101, P-106 transports the polymer to V-111 and P-107 transports the ferric chloride to R-101.

As these streams have a slightly higher density, a positive displacement pump would be a good consideration. A peristaltic pump was selected as it is a good choice for viscous, low flow dosing streams and is resistant to acid and caustic chemicals. It also ensures no leakages, which is desired when dosing chemicals. The material of construction for the peristaltic pumps is carbon steel for the shell and rotating parts and long-lasting rubber for the inner tubing. The rubber tubing provides resistance to corrosive and abrasive chemicals (Treutel, 2014).

5.8.3 *Rotary Lobe Positive Displacement Pumps*

(P-109 A/B, P-110 A/B, P-111 A/B, P-113 A/B, P-114 A/B, P-115 A/B)

P-109 recirculate the calamity tank, TK-103, contents back to the system and P-110 transports the discharge from TK-102 and TK-103 to R-101. P-111 is used to drain R-101 to TK-103 in the case of an emergency. P-113, P-114 and P-115 recirculates the solid underflow from V-111, F-101 and C-101, respectively, to R-101. P-117

These streams transport water containing sludge. As the sludge needs to be transported with minimum shear, rotary lobe pumps were selected. The material of construction for the rotary lobe pumps is carbon steel. This material have some corrosion resistance but also have good resistance to wear due to abrasion (Material selection for wastewater pumps, 2013).

5.8.4 *Conveyors*

(**CB-101, SC-101, SC-102, SC-103, SC-104, SC-105)**

Conveyor belt CB-101 was selected as the sludge is thick and heavy. This means that a high amount of power is required to transport it. The sludge is not high value, so losing sludge to the wind is not a major problem.

Screw Conveyor SC-101 was selected to transport the slaked lime powder as there is a reduced risk of losing the substance to the surrounding areas because of airflow. It allows for a higher degree of accuracy than the conveyor belt.

5.8.5 Blowers

In both cases the use of B-101 and B-102 is justified by the fact that the vapour steam being transported is at much higher pressure than the destination pressure. Air is available at 5.5 bar(a) but it needs to be transported to the heater which is at an operating pressure of 1.01 bara. Similar the steam is discharged at 44.3 but E-102 only operates at 3 bar(a). Naturally gaseous streams would move from a high-pressure to a low-pressure destination.

6 CAPITAL COST ESTIMATION

The capital cost estimation was performed using factors presented in Peters (Max S. Peters, 2003), whereby the capital cost are estimated based on the total delivered equipment costs. A summary of these factors to determine this capital requirement can be seen in Appendix G.

6.1 Main equipment cost

Table 4: Main equipment capital costs

Equipment Tag	Description	Parameter		Cost (R)
		Value	Unit	
R-101	Anaerobic Digester	5058	m^3	9 669 147
E-101	Primary Heat Exchanger	21.1	m^2	428 066
E-102	Secondary Heat Exchanger	1.84	m^2	363 048.1
E-103	Electric Heater	181.9	kW	403 795
H-101	Steam Heater	832.9	kW	9 414 984.6
V-101	Hydrated Lime Storage Tank (Hopper)	24.69	m^3	500 085.3
V-102	Urea Storage Tank (Hopper)	2.76	m^3	121 580.5
V-103	DAP Storage Tank (Hopper)	0.69	m^3	62 625.4
V-104	Micronutrient Storage Tank (Hopper)	-	m^3	-
V-105	Polymer Storage Tank (Hopper)	0.04	m^3	11 548.1
V-106	Slaked Lime Solution Vessel			4 583.3
V-107	Urea Solution Vessel			1 129.86
V-108	DAP Solution Vessel			3341.0
V-109	Micronutrient Solution Vessel	-	-	-
V-110	Polymer Solution Vessel			10 039.8
V-111	Thickener			2 001 447.5
F-101	Drum Filter	6	m^2	2 685 127.3
C-101	Centrifuge			47 865.08
TK-101	Ferric Chloride Solution Storage Tank	0.37	m^3	49 272.1
TK-102	Buffer Tank			467 044.5

6.2 Overview

The capital cost breakdown as well as the cost per m^3 treated (based on a plant life of 25Years running 350days per year) of the processing plant can be seen in Table 5

Table 5: Total capital cost breakdown

	Cost (R)	Cost (R/m3 treated)
Direct cost		
Purchased equipment delivered	34 898 455	2.83
Equipment installation	13 610 398	1.10
Piping (Installed)	10 818 521	0.88
Instrumentation and controls	4 536 799	0.37
Electrical (Installed)	3 489 846	0.28
Utilities	12 214 459	0.99
Off-sites	6 979 691	0.57
Building (Including Services)	10 120 552	0.82
Site preparation	2 093 907	0.17
Total Direct Cost	98 762 628	8.01
Indirect Cost		
Design, engineering and supervision	11 167 506	0.91
Construction expenses	11 865 475	0.96
Total Indirect Cost	23 032 980	1.87
Contractors fee	7 916 715	0.64
Contingency	39 583 573	3.21
Total Fixed Capital Cost	158 334 291	12.83
Working capital	27 953 663	2.27
Total Capital Cost	186 357 751	15.10
Total Capital Cost	186 357 751	15.10

The raw material and utility requirements for each unit were determined for each unit and summed in order to come to a total utility and raw material cost of the plant. The waste treatment and operating labour were both determined from calculations outlined in Appendix H. From these costs and the fixed capital investment cost in section 6 multiplication factors can be used to estimate the total operating

7 OPERATING COST ESTIMATION

The operating cost breakdown as well as the cost per m^3 treated (based on a plant life of 25Years running 350 days per year) of the processing plant can be seen in Table 6

7.1 Overview

Table 6: Operating cost breakdown

Item	**Cost (R)(year)**	**Cost (R/m3 treated)**
Direct Operating Cost	8 864 975	17.96
Raw materials	3 318 849	6.73

Waste treatment	95 550	0.19
Utilities	3 074 366	6.23
Operating labour	1 786 624	3.62
Direct supervisory and clerical labour	321 592	0.65
Maintenance and supplies	9 500 057	19.25
Laboratory charges	267 994	0.54
Patents and royalties	1 715 768	3.48
Total Direct Operating Cost	20 080 801	40.69
Fixed Operating Cost		
Depreciation	15 833 429	32.08
Local taxes and insurance	5 066 697	10.27
Plant overhead costs	6 964 964	14.11
Total Fixed Operating Cost	27 865 091	56.46
General Operating Cost		
Administration costs	1 741 241	3.53
Distribution and selling costs	6 291 149	12.75
Research and development	2 859 613	5.79
Total General Operating Cost	10 892 004	22.07

8 ECONOMIC ANALYSIS AND OVERALL RECOMMENDATION

8.1 Overview

Economic In order to evaluate the economic implications of the plant it was assumed that the cost of land was insignificant compared to the price of the plant.

It was Also assumed that the plant would take 2 years to build with the capital cost being spread equally between these two years, Therefore the first year of successful operation of the plant was assumed to be year 3 where the plant ran treated 100% of the influent to the process.

It was assumed that the penalty for effluent discharge was incorporated into the expenses of the industrial plant. Therefore the fee mitigated from treating the effluent of this industrial process was viewed as a sale and the cost of sales was viewed as the expenses of the plant.

8.2 Recommendation

As the Net Present Value (NPV) of the plant is R-164 351 000 the plant is currently not economically feasible. In order to increase the profitability of the plant one of two things would need to change. Either the penalties for the effluent discharge would have to increase or the operating costs of the Plant would have to decrease.

For the current processing plant to have a NPV of zero. The minimum penalty for the industrial effluent to be emitted would need to be 142.3 R/m^3 of discharge current trends in the escalation of polluting penalties should be analyzed into in order to determine whether the penalties for emitting the effluent would rise. If the effluent discharge penalty were to increase relative to the operating cost this increase would need to be 4.74% year on year to obtain a NPV of zero.

A factor one can’t consider in the financials of the feasibility of the processing plant is the potential of the effluent discharge to impede the water quality flowing to the upstream industrial plant. The effect this discharge would have on the industrial company’s stakeholders is of importance as such effects could impede the sales, productivity of workers and governmental leniency. Ultimately the Profitability of the upstream industry should be evaluated in ord er to determine whether the processing plant is feasible due to these important stakeholder interactions.

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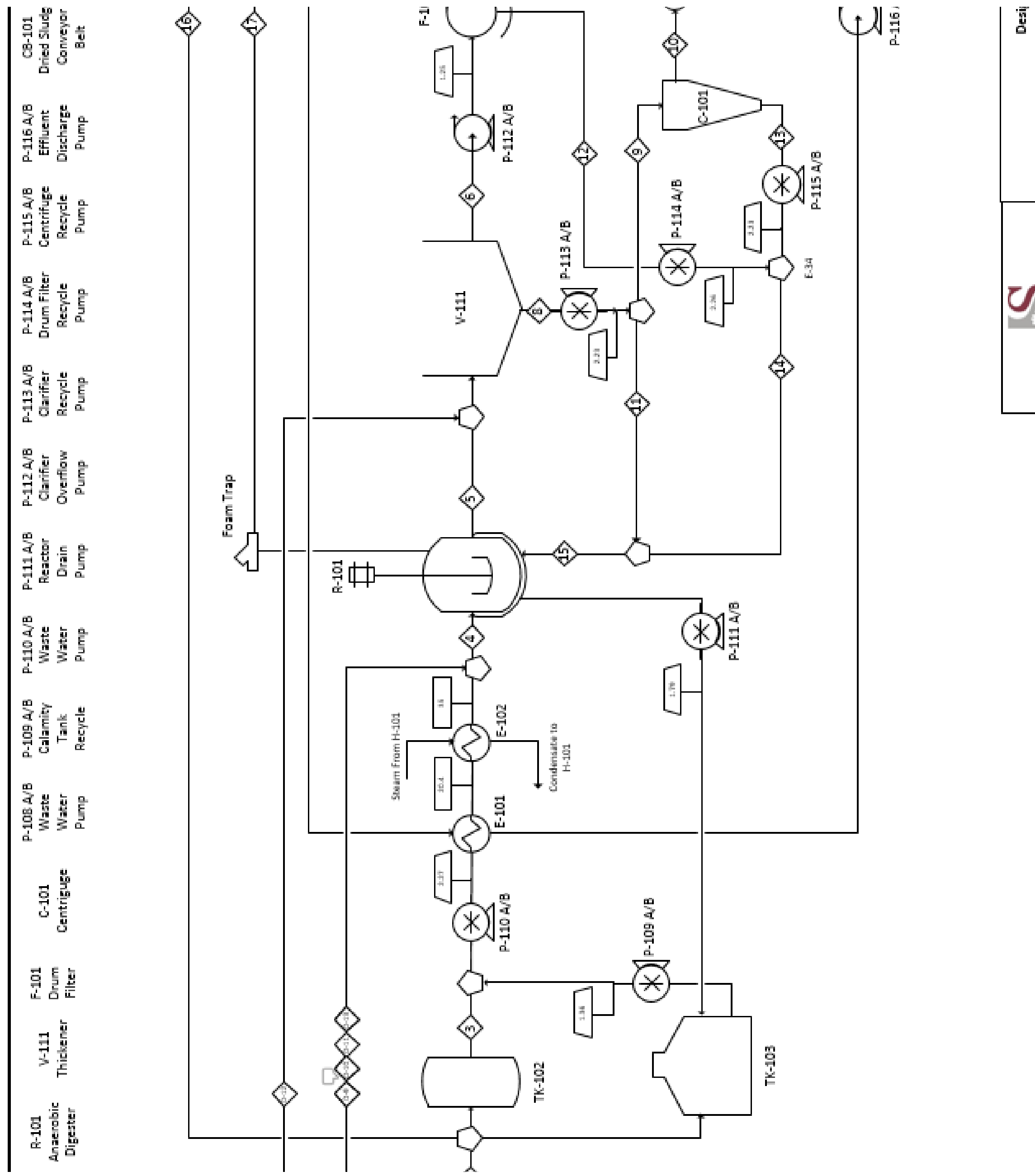
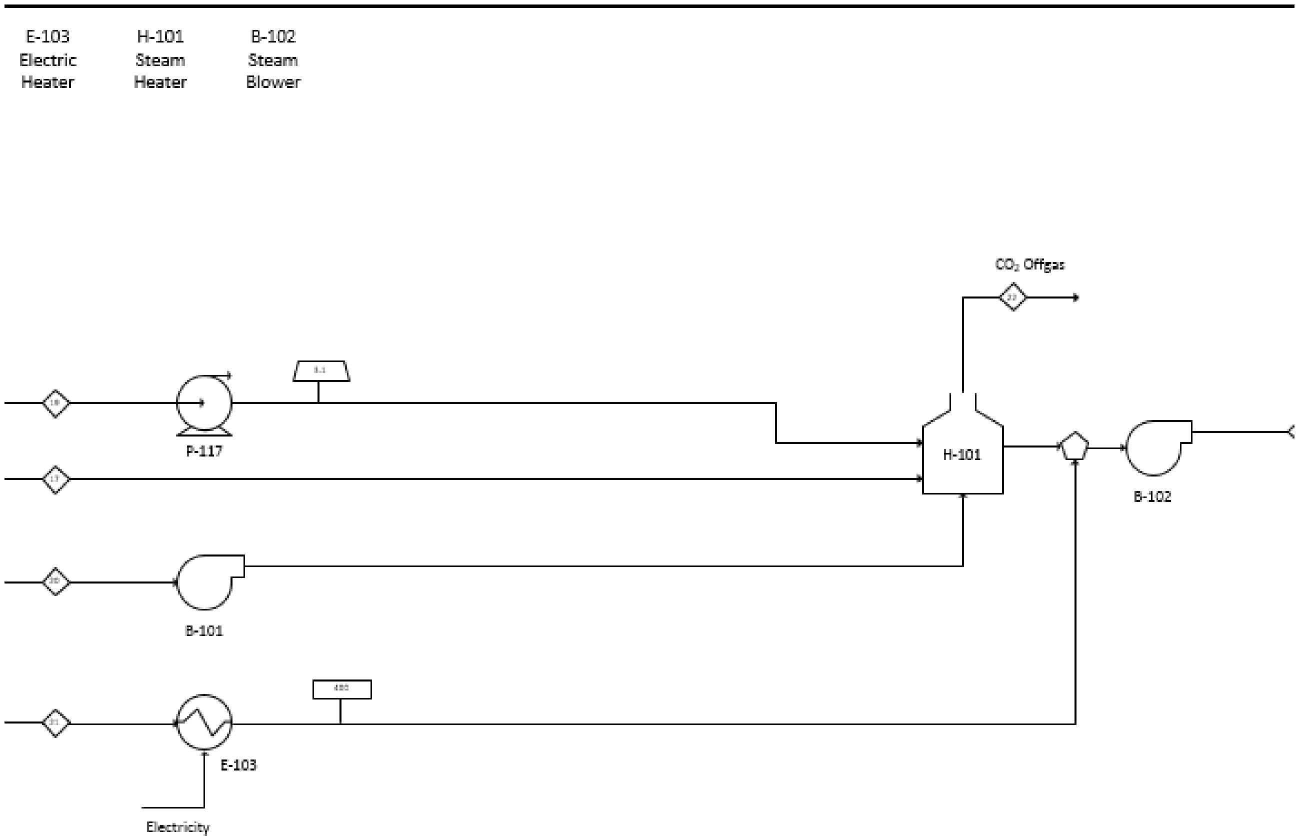
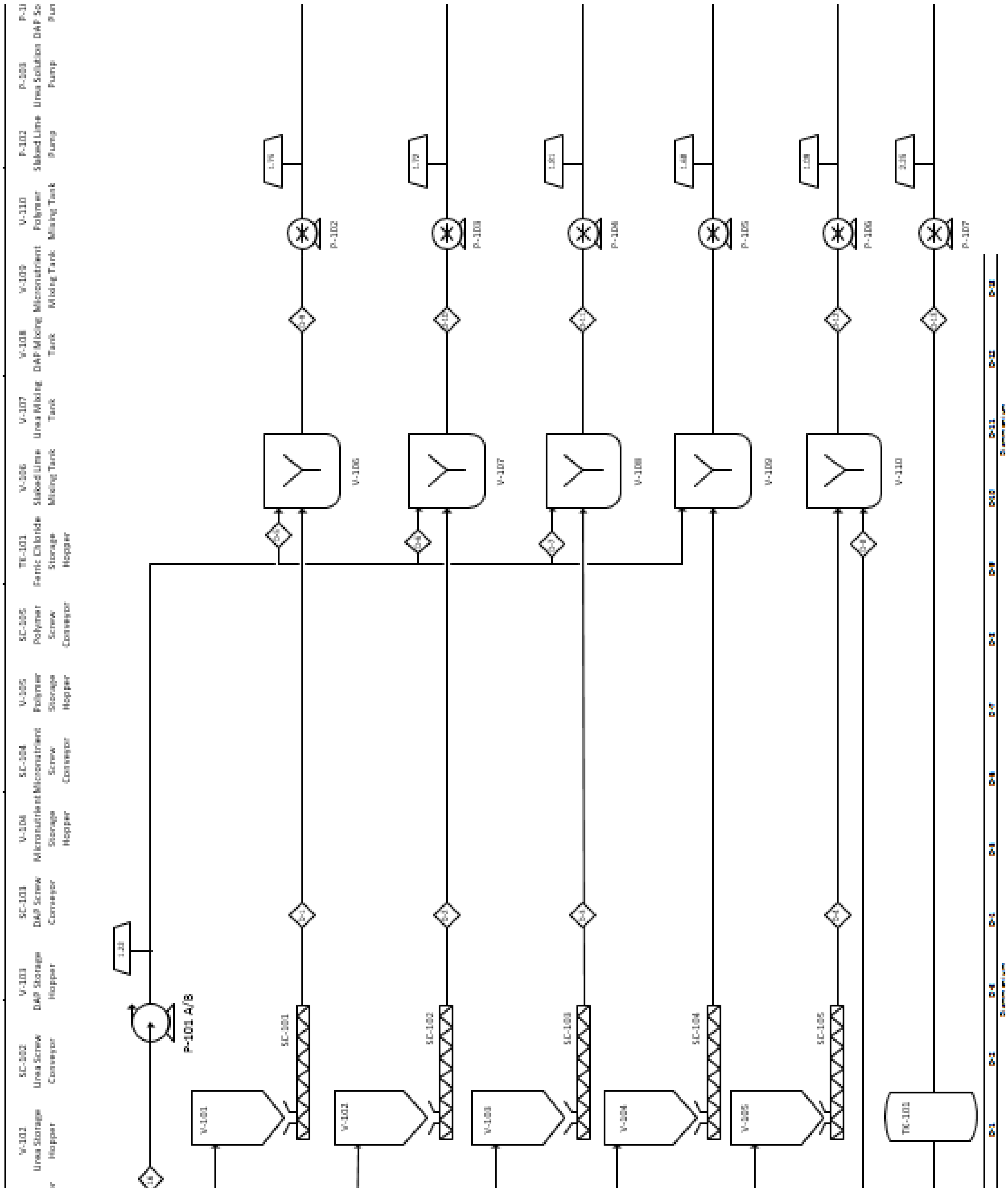
10 APPENDIX A – PROCESS FLOW DIAGRAM AND STREAM TABLE



	4	5	6	7	8	9	10	1
3.3	1410	1807.5	1417.7	1395	389.8	38.2	6.1	35
3.2	1408	1789.5	1415.9	1394.8	373.6	36.6	4.7	33
1	2	18	1.8	0.2	16.2	1.6	1.4	14
9.4	1452.6	1864.9	1460.5	1437	404.3	39.7	6.5	36
9.3	1450.2	1843.2	1458.4	1436.7	384.9	37.7	4.9	34
1	2.4	21.7	2.2	0.3	19.5	1.9	1.6	17
30	8880	15174	2590	1000	60931	60931	414395	60
46	6246	796	759	760	931	931	1863	93
95	1695	11981	1526	200	50000	50000	266954	50
9	339	3981	507	66	16615	16615	88709	16
56	1356	8000	1019	134	33385	33385	178245	33
48	12521	27428	3672	335	23756	2330	1961	21
45	8807	1439	1076	1060	363	36	11	32
19	2390	21657	2163	279	19494	1912	1625	17
4	478	7197	719	93	6478	635	540	58
5	1912	14461	1444	186	13016	1276	1085	11



	17	19	20	21	22	23
	Biogas	Water Condensate	Instrument Air	Process Water	Offgas	HP Steam
°C	35	115	25	115	1564.1	400
bar(a)	1	3	5.5	3	1	44.3
m3/d	3139.3	31.3	3042.5	5.6	169164.7	2160.5
t/d	3.2	32.2	26.6	5.8	33.1	32.2



11 APPENDIX B – PROCESS HIGHLIGHTS SAMPLE CALCULATIONS

COD Loading rate:

$$\dot{m}_{TCOD,1} = \dot{V}_{17} \times [TCOD]_{17}$$

$$\dot{m}_{TCOD,1} = 1410\,\frac{m^3}{d} \times 8.880\,\frac{g}{l}$$

$$\dot{m}_{TCOD,1} = 12520.8\,\frac{kgCOD}{d}$$

Volumetric Organic Loading:

$$VOL = \frac{\text{COD loading rate}}{V_{reactor}}$$

$$VOL = \frac{12520.8\,\frac{kgCOD}{d}}{4598\ m^3}$$

$$VOL = 2.7\,\frac{kgCOD}{m^3 d}$$

Biomass Activity:

$$\text{Biomass activity} = \frac{Q_{in}(COD_{in} - COD_{out})}{VSS_{reactor}}$$

(Cruz, 2016)

$$\therefore \text{Biomass activity} = \frac{TCOD_{22}}{VSS_{reactor}}$$

$$\therefore \text{Biomass activity} = \frac{TCOD_{22}}{V_{reactor} \times [VSS]_{reactor}}$$

$$\therefore \text{Biomass activity} = \frac{9164\,\frac{kg\ COD}{d}}{4598\ m^3 \times 8\,\frac{kgVSS}{m^3}}$$

$$\therefore \text{Biomass activity} = 0.25\,\frac{kgCOD}{kgVSSd}$$

Solids retention time:

$$SRT = \theta_c = \frac{TSS_{reactor}}{\dot{m}_{TSS,12} + \dot{m}_{TSS,16}}$$

$$\theta_c = \frac{V_{reactor} \times [TSS]_{reactor}}{\dot{m}_{TSS,25} + \dot{m}_{TSS,30}}$$

$$\theta_c = \frac{4598\ m^3 \times 11981\,\frac{g}{m^3}}{279\,\frac{kg}{d} + 1625\,\frac{kg}{d}}$$

$$\theta_c = 28.9\ d$$

Hydraulic retention time:

$$\mathrm{HRT} = \frac{(\mathrm{V_{reactor}} - \frac{\mathrm{TSS_{reactor}}}{\mathrm{\rho_{TSS}}}) \times \mathrm{\rho_{water}}}{\mathrm{\dot{m}_{water,25}} + \mathrm{\dot{m}_{water,30}}}$$

$$\mathrm{HRT} = \frac{(4598\ \mathrm{m^3} - \frac{4598\ \mathrm{m^3} \times 11981 \frac{\mathrm{g}}{\mathrm{m^3}}}{1200 \frac{\mathrm{kg}}{\mathrm{m^3}}}) \times 1030 \frac{\mathrm{kg}}{\mathrm{m^3}}}{1436695 \frac{\mathrm{kg}}{\mathrm{d}} + 4875 \frac{\mathrm{kg}}{\mathrm{d}}}$$

$$\mathrm{HRT} = 3.3\ \mathrm{d}$$

12 APPENDIX C – ANAEROBIC DIGESTOR DESIGN CALCULATIONS AND SPECIFICATION SHEET

12.1 Sizing

The reactor was sized within the parameters set by Qasim (Syed Qasim, 2018) wherein the reactor should be within the following specs displayed in table XX. The volume of the reactor was sized from the following equation wherein the Volatile solids reduction can be used to calculate the Solids retention time required to get the particulate COD destruction:

$$\mathrm{VSR} = 13.7\ln(\theta_c) + 18.9$$

$$\mathrm{VSR} = \frac{\varepsilon_{\mathrm{PCOD}} \times \dot{m}_{\mathrm{BioVSS,5}}}{\dot{m}_{\mathrm{VSS,5}}}$$

$$\mathrm{VSR} = \frac{0.87 \times 1259\frac{\mathrm{kg}}{\mathrm{d}}}{1911.96\frac{\mathrm{kg}}{\mathrm{d}}}$$

$$\mathrm{VSR} = 65 = 13.7\ln(\theta_c) + 18.9$$

$$\theta_c = 28.9 \text{ days}$$

$$\theta_c = \frac{\mathrm{TSS_{reactor}}}{\dot{m}_{\mathrm{TSS,12}} + \dot{m}_{\mathrm{TSS,16}}}$$

$$\mathrm{TSS_{reactor}} = [\mathrm{TSS}]_{\mathrm{reactor}} \times \mathrm{V_{reactor}}$$

$$\mathrm{V_{reactor}} = \frac{\theta_c \times (\dot{m}_{\mathrm{TSS,12}} + \dot{m}_{\mathrm{TSS,16}})}{[\mathrm{TSS}]_{\mathrm{reactor}}}$$

$$\mathrm{V_{reactor}} = \frac{28.9 \times (279.02 + 1625)}{11.981}$$

$$\mathrm{V_{reactor}} = 4598\ \mathrm{m}^3$$

The reactor was then sized geometrically to ensure adequate mixing in dimensions set out by tank 1 in Maroney's paper (Meroney, 2008). This was sized by keeping the same side wall height to diameter ratio as well as cone height to diameter ratio as tank 1 in Maroney's paper. By having the same dimensions as the tank within Maroney's paper the mixing within the reactor can be modelled to have similar mixing characteristics with these characteristics the duties for the draft tubes from the study can be used. Tank 1 was selected as this configuration has external draft tubes only which will allow for maintenance of tubes used for mixing to be completed quickly to ensure minimal distribution to the process. As well as being within the geometrically set parameters set by Qasim (Syed Qasim, 2018). These parameters where that the design of anaerobic digesters were sized between 6 and 40m in diameter, the conical slope was between 4 and 6 meters horizontal to 1 meter vertical and the side wall height is between 7.5 to 15m. (Syed Qasim, 2018)

Table 7: Reactor dimensions

	Tank 1	*Reactor*
Diameter (m)	30.5	25

Side water depth (m)	10.1	8.23
Cone depth (m)	3.8	3.1
Volume (m³)	8304	4598

12.1.1 Thickness

12.1.1.1 Cylindrical Portion

The thickness calculation of the anaerobic digester was calculated using the method detailed in (Nikunj & Bhabhor, 2013).

The maximum working pressure is the pressure at the base of the anaerobic digester. Assuming the gas in the headspace has a negligible effect on the force on the cylindrical portion of the digester the pressure exerted at the bottom of the cylindrical section is equal to:

$$t_{cyl} = \frac{P_i * D_i}{2 * \sigma_{all}\eta_L - P_i} + C$$

Where Pi is equal to the liquid head of the cylindrical portion of the reactor.

$$P_i = \rho_{mix} * g * h_{cylinder}$$
$$P_i = 1100 * 9.81 * 8.29$$
$$P_{i,gauge} = 89494.9\ Pa$$
$$P_{i,abs} = 190819.94 Pa$$

The thickness was calculated using the cylindrical portion as the conical portion at the bottom will be supported by the foundation, therefore not liable to break.

A safety factor was calculated. If the anaerobic digester was to overflow (i.e. the headspace were to fill with gas) the reactor should be able to withstand the extra force exerted by the liquid. This extra force is quantified as:

$$P_{headspace\ (gauge)} = \frac{\rho_{mix} V_{half\,Sphere}\, g}{\pi r^2}$$
$$P_{headspace\,(gauge)} = \frac{1100 * \left[\frac{4}{3} * \pi * (6.03)^3\right] * 9.81}{2 * \pi(6.03)^2}$$
$$P_{headspace\,(gauge)} = 86328\ Pa \cong\ P_{i,abs}$$

This value is roughly equal to the pressure exerted by the liquid. Therefore a safety factor of 2 is used.

$$P_{design} = 2 * P_{i,abs}$$
$$P_{design} = 2 * \ 190819.94$$
$$P_{design} = 381639.89\ kPa$$

The σ_{all} value needs to be worked out. (Nikunj & Bhabhor, 2013) suggests using:

$$\sigma_{all} = \frac{S_{yt}}{1.5}$$

The yield tensile strength was found to be 215 MPa (ASM Aerospace specification metals, 2019).

$$\sigma_{all} = \frac{215MPa * \frac{\frac{1N}{1\,mm^2}}{MPa}}{1.5}$$
$$\sigma_{all} = 143\frac{N}{mm^2}$$

The longitudinal joint efficiency and corrosion resistance is taken as 2mm as per (Nikunj & Bhabhor, 2013).

Therefore the thickness of the cylindrical section (after making the necessary unit conversions):

$$t_{cyl} = \frac{0.382\ \left[\frac{N}{mm^2}\right] * 25044.67\ [mm]}{2 * 143\left[\frac{N}{mm^2}\right] * 2[mm] - 0.382\left[\frac{N}{mm^2}\right]} + 2[mm]$$
$$t = 16.68\ mm$$

12.1.1.2 Dome

The top of the reactor is a hemisphere. The pressure is as follows (Nikunj & Bhabhor, 2013):

$$t_{head} = \frac{2 + K_1^2}{6} * \frac{P_i d_i}{2\sigma_{all}\eta_L - 0.2P_i} + C$$

As the top is half a sphere K_1 is equal to 1 (Nikunj & Bhabhor, 2013). Assuming ideal gas as the pressure is low.

$$Q_{biogas} = 2138.36\frac{m^3}{d}$$
$$\dot{n}_{biogas} = 124.16\frac{mol}{d}$$
$$P_{i,gauge} = \frac{\dot{n}\ R\ T}{Q}$$
$$P_{i,gauge} = \frac{124.16 * 8.314 * 308}{2138.36}$$
$$P_{i,gauge} = 148.69\ Pa$$

$$P_{i,abs} = 101473.69\ Pa$$

Assuming it takes 1 full day to fix any biogas blockage which will result in gas buildup in the headspace. The safety factor can be calculated using the ideal gas law.

$$P_{i,gauge} = \frac{n\ R\ T}{V}$$
$$P_{i,gauge} = \frac{124.16 * 8.314 * 308}{2138.36}$$
$$P_{safety,gauge} = 148.69\ Pa = P_{i,gauge}$$

Therefore, the safety factor is essentially 2.

$$P_{design,abs} = 0.203\frac{N}{mm^2}$$

The volume of the head of the digester is 10% of the whole volume (Samer).
The σ_{all}, corrosion factor and longitudinal joint efficiency are the same as previously.

$$V = 459.8\ m^3$$
$$D = 12.07\ m = 1207mm$$
$$t_{head} = \frac{2+1}{6} * \frac{0.203 * 1207}{2 * 143.33 * 2 - 0.2(0.203)} + 2$$
$$t_{head} = 2.21\ mm$$

12.2 Mixing

$$\mathrm{G_{tank\ 1}} = \sqrt{\frac{\mathrm{P_{Mixers}}}{\mathrm{V_{tank\ 1}} \times \mu}} = 71\frac{1}{\mathrm{s}}$$

$$\mathrm{G_{reactor}} = \sqrt{\frac{\mathrm{P_{Mixers}}}{\mathrm{V_{tank\ 1}} \times \frac{\mathrm{V_{reactor}}}{\mathrm{V_{tank\ 1}}} \times \mu}}$$

$$\therefore \mathrm{G_{reactor}} = \mathrm{G_{tank\ 1}} \sqrt{\frac{1}{\frac{\mathrm{V_{reactor}}}{\mathrm{V_{tank\ 1}}}}}$$

$$\therefore \mathrm{G_{reactor}} = 71\frac{1}{\mathrm{s}} \sqrt{\frac{1}{\frac{4978\ \mathrm{m^3}}{8301\ \mathrm{m^3}}}}$$

$$\therefore \mathrm{G_{reactor}} = 94\frac{1}{\mathrm{s}}$$

$$\mathrm{DVTT_{Tank\ 1}} = \frac{\mathrm{V_{Tank\ 1}}}{\mathrm{Q_{mixers}}}$$

$$\mathrm{DVTT_{Reactor}} = \frac{\mathrm{V_{Tank\ 1}}}{\mathrm{Q_{mixers}}} \times \frac{\mathrm{V_{reactor}}}{\mathrm{V_{tank\ 1}}}$$

$$\mathrm{DVTT_{Reactor}} = 54\ \mathrm{min} \times \frac{4978\ \mathrm{m^3}}{8301\ \mathrm{m^3}}$$

$$\mathrm{DVTT_{Reactor}} = 30\mathrm{min}$$

With this volume value other processing parameters can be calculated. These must then also fit within the parameters given by Qasim (Syed Qasim, 2018) a summary of these operating conditions can be seen in Table XX.

Table 8: Typical process parameters for the anaerobic contact process

Parameter	Units	Low Value	High Value
Volumetric Organic loading	kg COD/m^3.d	2	5

COD Removal efficiency	%	75	90
Solids retention time	days	15	30
Hydraulic retention time	h	12	120
MLVSS concentration in reactor	kg/m^3	4	8
Surface overflow rate in thickener	$m^3/m^2.h$	0.5	1

The Volumetric organic loading can be calculated by the following equation:

$$\mathrm{VOL} = \frac{\dot{m}_{\mathrm{TCOD}}}{V_{\mathrm{reactor}}}$$

$$\mathrm{VOL} = \frac{12520.8\,\frac{\mathrm{kg}}{\mathrm{day}}}{4598\ \mathrm{m}^3}$$

$$\mathrm{VOL} = 2.7\,\frac{\mathrm{kgCOD}}{\mathrm{m}^3\mathrm{d}}$$

The Hydraulic retention time (HRT) can be calculated in the same way as SRT according to the following equation:

$$\mathrm{HRT} = \frac{m_{\mathrm{reactor}}}{\dot{m}_{\mathrm{water,12}} + \dot{m}_{\mathrm{water,16}}}$$

$$\mathrm{HRT} = \frac{\left(V_{\mathrm{reactor}} - \frac{\mathrm{TSS}_{\mathrm{reactor}}}{1200}\right) \times 1030}{\dot{m}_{\ 12} - \dot{m}_{\mathrm{12,TSS}} + \dot{m}_{\ 16} - \dot{m}_{\mathrm{16,TSS}}}$$

$$\mathrm{HRT} = \frac{\left(4598\ \mathrm{m}^3 - \frac{55087\mathrm{kg}}{1200\frac{\mathrm{kg}}{\mathrm{m}^3}}\right) \times 1030\,\frac{\mathrm{kg}}{\mathrm{m}^3}}{1436974\,\frac{\mathrm{kg}}{\mathrm{d}} - 279\,\frac{\mathrm{kg}}{\mathrm{d}} + 6500\,\frac{\mathrm{kg}}{\mathrm{d}} - 1625\,\frac{\mathrm{kg}}{\mathrm{d}}}$$

$$\mathrm{HRT} = 3.25\ \mathrm{d}$$

$$\mathrm{HRT} = 78\ \mathrm{hr}$$

12.3 Materials

The material of construction of the reactor was designed based on a publication in which it was described that for a 25 year plant life the cost of a stainless steel reactor compared to other materials is cheaper (The Steel Construction Institute, 2016). As metal is a good conductor of electricity the heat transfer implications of the reactor must be considered. The heat loss from the rector can be calculated by the following equations: Assuming the worst case scenario in the reactor is exposed to ambient air at -3°C and assuming a high heat transfer coefficient value for free convection, the worst case scenario of heat flux leaving the reactor can be calculated can be calculated.

Assuming the inside of the reactor is 37°C and the reactor is made of ASI 304 Stainless steel which is 8mm thick (Vertical 5000 m3 Cylindrical steel tank, 2019). The Heat flux out of the reactor can be calculated based on the stainless steel being exposed to ambient air with natural convection. As the size of the reactor has already been calculated the area at which the tank losses heat can be calculated according to the following equation.

$$\mathrm{Cone_{SA}} = \pi \times \mathrm{r}\sqrt{\mathrm{r}^2 + \mathrm{h}_{cone}^2}$$

$$\mathrm{Cone_{SA}} = \pi \times 12.5\sqrt{12.5^2 + 3.1^2}$$

$$\mathrm{Cone_{SA}} = 507.6\ \mathrm{m}^2$$

$$\mathrm{Roof_{SA}} = \pi \times \mathrm{r}^2$$

$$\mathrm{Roof_{SA}} = \pi \times 12.5^2$$

$$\mathrm{Roof_{SA}} = 492.6\ \mathrm{m}^2$$

$$\mathrm{Side\ walls_{SA}} = 2 \times \pi \times \mathrm{r} \times \mathrm{h_{side\ walls}}$$

$$\mathrm{Side\ walls_{SA}} = 2 \times \pi \times 12.5 \times 10.1$$

$$\mathrm{Side\ walls_{SA}} = 652.5\ \mathrm{m}^2$$

$$\mathrm{SA_{Tot}} = \mathrm{Side\ walls_{SA}} + \mathrm{Roof_{SA}} + \mathrm{Cone_{SA}}$$

$$\mathrm{SA_{Tot}} = 1653\ \mathrm{m}^2$$

This total surface are of the reactor can then be used to calculate the amount of heat loss form a reactor just constructed of steel. Assuming a high value for natural convection as 25 W/m^{2}K (Yunus A. Cengel, 2015) and that the thermal diffusivity of AISI 304 Steel remains constant at 14.9 W/mK.

$$\mathrm{R_{tot}} = \frac{\mathrm{L}}{\mathrm{k\,A}} + \frac{1}{\mathrm{h\,A}}$$

$$\mathrm{R_{tot}} = \frac{0.01668}{14.9 \times 1652.8} + \frac{1}{25 \times 1652.8}$$

$$\mathrm{R_{tot}} = 2.49 \times 10^{-5}$$

$$\mathrm{q} = \frac{(\mathrm{T_1} - \mathrm{T_2})}{\mathrm{R_{tot}}}$$

$$\mathrm{q} = \frac{(40)}{2.49 \times 10^{-5}}$$

$$\mathrm{q} = 1607.9\ \mathrm{kW}$$

$$\Delta\mathrm{T} = \frac{\mathrm{q}}{\mathrm{c_p}\,\dot{\mathrm{m}}}$$

$$\Delta\mathrm{T} = \frac{1607.9\ \mathrm{kW}}{4.178\frac{\mathrm{kj}}{\mathrm{kg\,K}}\ \ 21.3\frac{\mathrm{kg}}{\mathrm{s}}}$$

$$\Delta\mathrm{T} = 17.8\ \mathrm{K}$$

As just having stainless steel as the material of construction would result in an 18K temperature change leaving the reactor. Therefore insulation should be applied to the reactor in order to maintain the temperature of the streams in the recycle as well as the streams that lead to the primary heat exachanger.

$$R_{tot} = \frac{L}{k\,A} + \frac{L}{k\,A} + \frac{1}{h\,A}$$

$$R_{tot} = \frac{0.01668}{14.9 \times 1652.8} + \frac{0.2}{0.046 \times 1652.8} + \frac{1}{25 \times 1652.8}$$

$$R_{tot} = 0.0027$$

$$q = \frac{(T_1 - T_2)}{R_{tot}}$$

$$q = \frac{(40)}{0.0027}$$

$$q = 15.1\ \mathrm{kW}$$

$$\Delta T = \frac{q}{c_p\,\dot{m}}$$

$$\Delta T = \frac{15.1\mathrm{kW}}{4.178\frac{\mathrm{kJ}}{\mathrm{kgK}} \times 21.3\frac{\mathrm{kg}}{\mathrm{s}}}$$

$$\Delta T = 0.17\ \mathrm{K}$$

Estimating the heat loss from the Thickener assuming that the top fluid is the major contributor to the loss of temperature of the streams. This loss in temperature can be calculated as follows:

$$R_{tot} = \frac{1}{h\,A}$$

$$R_{tot} = \frac{1}{25\frac{\mathrm{W}}{\mathrm{m^2K}}\ 80\ \mathrm{m^2}}$$

$$R_{tot} = 0.0005$$

$$q = \frac{(T_1 - T_2)}{R_{tot}}$$

$$q = \frac{(40)}{0.0005}$$

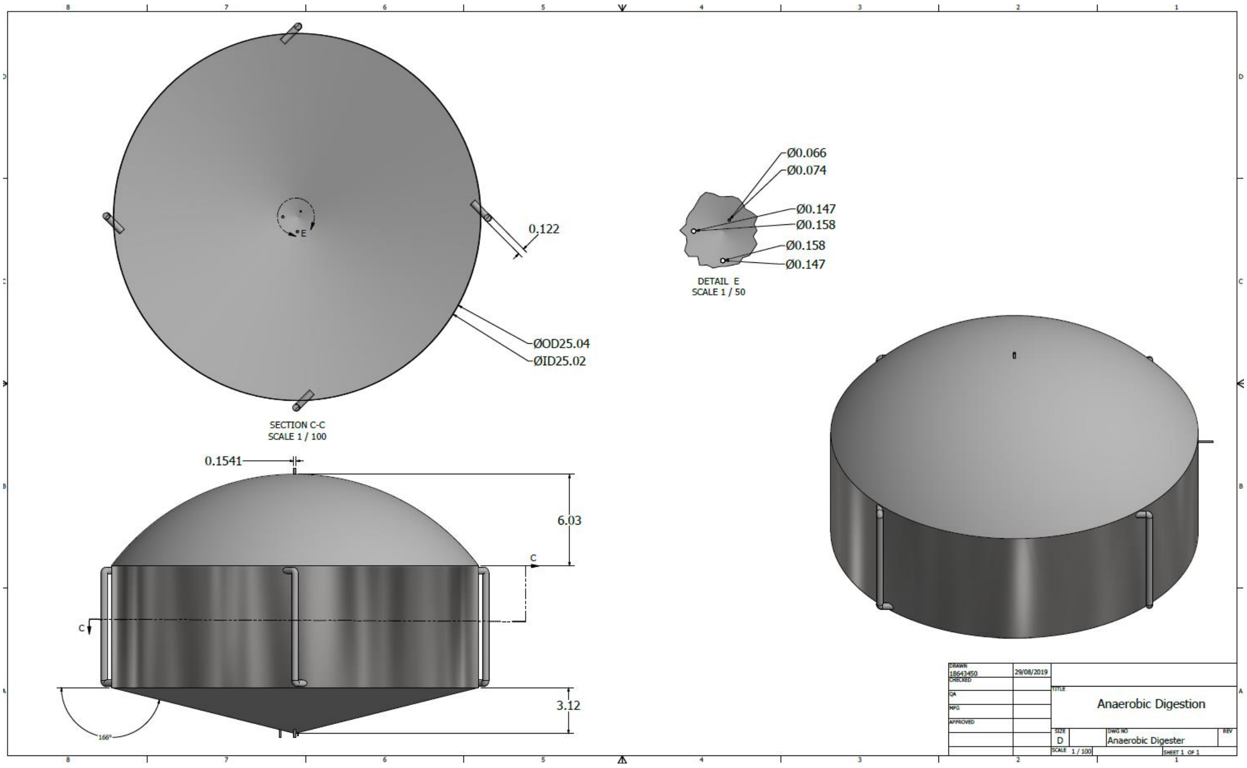
$$q = 80116\ \mathrm{W}$$

$$\Delta T = \frac{q}{c_p\,\dot{m}}$$

$$\Delta T = \frac{80.1\mathrm{kW}}{4.178\frac{\mathrm{kJ}}{\mathrm{kgK}}\ 21.3\frac{\mathrm{kg}}{\mathrm{s}}}$$

$$\Delta T = 0.9\ \mathrm{K}$$

13 APPENDIX D – ANAEROBIC DIGESTOR DRAWING



14 APPENDIX E – EQUIPMENT SIZING

14.1 Thickener:

The thickener or thickener was sized in order to allow for operation in the lower bounds of VSS concentration within the reactor. This is due to the limiting settling flux within the thickener would be controlled in part due to the concentration of the suspended solids concentration leading to the thickener (T.G. Shea, 1974). Data obtained from settling tests conducted on the effluent of anaerobic reactor waste performed by (T.G. Shea, 1974) showed an increase in limiting layer flux as the concentration of the MLVSS was increased. Therefore in sizing the Thickener/ thickener using the minimum TSS concentration at which the reactor would operate is used. A line of best fit was used to estimate the degree to which the flux varied with concentration of the TSS concentration entering the thickener.

When the reactor is operated at a VSS concentration of 4000mg/l (low for an anaerobic digester according to (Syed Qasim, 2018) and if one assumes the ratio of VSS/TSS remains equal at this new operating concentration to the mass balance presented in Appendix B. it follows that:

$$[\mathrm{TSS}] = [\mathrm{VSS}] \times \frac{[\mathrm{TSS}]}{[\mathrm{VSS}]}$$

$$[\mathrm{TSS}] = 4000 \times \frac{11981}{8000}$$

$$[\mathrm{TSS}] = 5990.7 \frac{\mathrm{mg}}{\mathrm{l}}$$

The trend line plotted from the data in from the settling tests conducted by Shea (T.G. Shea, 1974) yielded the following equation:

$$y = 0.0316x - 56.026$$

Where:

Y= Solids flux in limiting layer

X= Initial TSS concentration

This equation was then used in order to calculate the settling flux in the limiting layer.

$$\text{Solids flux in limiting layer} = 0.0316 \times (5590.7) - 56.026$$

$$\text{Solids flux in limiting layer} = 133.28 \frac{\text{kg}}{\text{m}^2\text{day}}$$

$$\text{Solids flux in limiting layer} = 5.55 \frac{\text{kg}}{\text{m}^2\text{hr}}$$

From this solids limiting flux the area of the thickener can be calculated:

$$A = \frac{Q \times X}{SF_L}$$

$$A = \frac{75.3116 \times 5.9907}{5.55 \frac{\text{kg}}{\text{m}^2\text{hr}}}$$

$$A = 81.2\ m^2$$

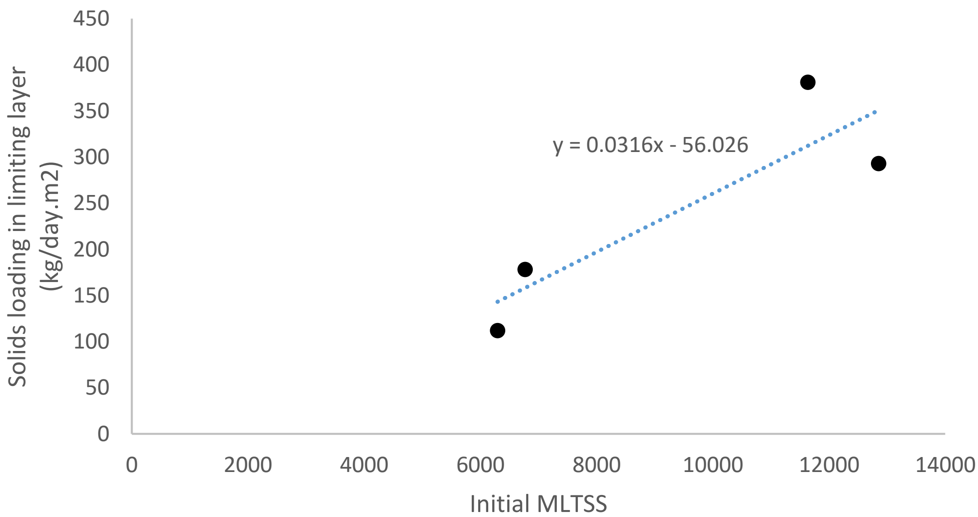


Figure 1: Batch settling tests on Anaerobic reactor effluent (T.G. Shea, 1974)

14.2 Vacuum Drum Filter

The Vacuum drum filter was sized in accordance with the processing parameters the JX Filtration drum filter can operate at. The Filtration capacity of their units specify the ranges at which they operate in terms of mass flowrate. This allowed the area of filtration to be sized based the the mass flowrate leading to the reactor.

It was found that their ZG1 would be able to filter the stream 6 as its mass flowrate is within its specification.

$$A = 6\mathrm{m}^2$$

14.3 Tanks

14.3.1 Influent Buffer Tank

The buffer tank was sized in terms of fluid detention time. A half-full detention time of 10 minutes is generally used for buffer tanks (Faanes, 2000).

The flowrate into the buffer tank was given as 1410 m^3/d from the mass balance. A daily flow disturbance of 15% can be expected and thus the tank was designed with a maximum holding capacity that included the upper limit of the flow deviation.

The tank volume was calculated as follows.

$$Q = 1410\frac{\mathrm{m}^3}{\mathrm{d}}$$

$$Q = 1410 \times \left(\frac{1}{24 \times 60}\right) = 0.98\frac{\mathrm{m}^3}{\mathrm{min}}$$

$$Q_{\mathrm{max}} = 0.98 \times 1.15 = 1.13\frac{\mathrm{m}^3}{\mathrm{min}}$$

$$V = Q \times t$$

$$V = 1.13 \times (2 \times 10)$$

$$V = 22.52\ \mathrm{m}^3$$

The tank diameter and length was then calculated. The length to diameter (L/D) ratio for a vertical buffer tank was taken as 3 (ref).

$$L = 1.2D$$

$$V = \frac{\pi D^2 L}{4}$$

$$\sqrt[3]{D^3} = \sqrt[3]{\frac{4V}{1.2\pi}}$$

$$D = 2.88\ \mathrm{m}$$

$$L = 1.2D = 1.2 \times 2.88$$

$$L = 3.46 \text{ m}$$

14.3.2 Calamity Tank

The calamity tank was sized in terms of fluid detention time. A detention time of 3 days was used for the calamity tank, as was for the anaerobic reactor, to allow the calamity tank to hold the entire volume of the reactor (ref).

Assuming all influent is directed to flow into the calamity tank, the flowrate was given as 1410 m^3/d from the mass balance.

The tank volume was calculated as follows.

$$Q = 1410 \frac{m^3}{d}$$

$$V = Q \times t$$

$$V = 1410 \times (3)$$

$$V = 4230 \text{ m}^3$$

The tank diameter and length was then calculated. The length to diameter (L/D) ratio for a vertical calamity tank was taken as 1/3 (ref).

$$D = 3L$$

$$V = \frac{\pi D^2 L}{4}$$

$$\sqrt[3]{L^3} = \sqrt[3]{\frac{4V}{3^2 \pi}}$$

$$L = 8.43 \text{ m}$$

$$D = 3L = 3 \times 8.43$$

$$D = 25.28 \text{ m}$$

14.3.3 Hopper storage tanks

For each of the hopper storage tanks it is imperative that the chosen inventory is large enough should anything deviate from specification that negatively impacts on the performance of the process occurring within the anaerobic digester. The sizing of the storage tanks was done so that when it occupies 50 % of the total volume, there is enough material to last an entire week. The methodology followed to obtain the required volume together with the costing procedure is shown below, but as a demonstration only the sample calculations for unit V-101 will be shown:

$$t_{res} = 168 \text{ hr}$$

According to a report distributed by (Engineer's Aide Reference Guide, 2007) the length to diameter ratio is generally assumed to be equal to 3.

$$\mathrm{L/D} = 3$$

$$\mathrm{L} = 3\mathrm{D}$$

To ensure that the desired pH of 7.2 is maintained within the anaerobic digester the mass flowrate of calcium hydroxide [$Ca(OH)_2$] was determined to be 4109.2 kg/d. The mass flowrate together with the assumed residence time can be used to determine the material holdup from which the volume can be established, but to correctly determine the volume of the hopper storage tank the fact that it is only 50 % full must be accounted by incorporating a factor of 2 as shown:

$$\mathrm{Holdup} = \dot{\mathrm{m}}_{\mathrm{D-1}}\mathrm{t}_{\mathrm{res}}$$

$$\mathrm{Holdup} = \left(\frac{4109.2\ \mathrm{kg}}{\mathrm{d}}\right)\left(\frac{\mathrm{d}}{24\ \mathrm{hr}}\right)(168\ \mathrm{hr})$$

$$\mathrm{Holdup} = 28764.4\ \mathrm{kg}$$

$$\mathrm{V}_{\mathrm{V-101}} = 2\frac{\mathrm{Holdup}}{\rho_{\mathrm{Ca(OH)_2}}}$$

$$\mathrm{V}_{\mathrm{V-101}} = 2\left(\frac{28764.4\ \mathrm{kg}}{2329.1\ \mathrm{kg/m^3}}\right)$$

$$\mathrm{V}_{\mathrm{V-101}} = 24.69\ \mathrm{m^3}$$

Since the volume is known the dimensions of the hopper storage tank can be determined by expressing volume in terms of both the length and the diameter:

$$\mathrm{V}_{\mathrm{V-101}} = \frac{\pi}{4}D^2L$$

$$\mathrm{V}_{\mathrm{V-101}} = \frac{\pi}{4}D^2(3D)$$

$$24.69 = \frac{3\pi}{4}D^3$$

$$D = 2.2\ m$$

$$\mathrm{L} = 3\mathrm{D} = 6.6\ \mathrm{m}$$

Table 9: Summary table for the sizing of the various storage tanks (hoppers)

Parameter	V-101	V-102	V-103	V-105	TK-106
Flowrate (kg/d)	4109.2	263.4	78.9	2.9	76.7
Density (kg/m^3)	2329.61	1335.04	1611.98	1030	2891.96
Volume (m^3)	24.69	2.76	0.69	0.04	0.37
Diameter (m)	2.2	1.1	0.7	0.3	0.5
Length (m)	6.6	3.3	2.1	0.9	1.5

Configuration	Vertical	Vertical	Vertical	Vertical	Vertical

14.4 Transport

14.4.1 Pumps

The pump shaft power is the main parameter that determines the size and cost of a specific pump. The procedure followed to calculate the pump shaft power is shown below. All pumps were sized in the same manner and sample calculations for P-108 is shown. A summary for all calculated pump values is shown in table at the end.

Firstly, the system head has to be calculated and can be divided into the static head and the dynamic head. The static head refers to any pressurized systems and vertical fluid columns attached to the pump, while the dynamic head refers to the straight run head-loss and fittings head-loss. A common superficial velocity for systems pumping water-like fluids with settleable solids is generally between 1 and 1.5 m/s. A starting superficial velocity of 1.3 m/s was assumed for the pump sizing calculations. In order to get the real superficial velocity, the following procedure was followed:

$$Pipe\ cross-sectional\ area = \frac{Q}{v}$$

$$Pipe\ cross-sectional\ area = \frac{0.0163}{1.3} = 0.126\ m$$

From Turton the correlating pipe internal diameter was found to be

$$Std\ cross\ sectional\ area = 129.1\ \times 10^{-4} m^2$$

$$ID = 128.2\ mm$$

$$v_{real} = \left(\frac{0.0163}{129.1}\right) \times 10^4$$

$$v_{real} = 1.26\ m/s$$

The process inputs for P-108 is shown below.

Process inputs	**Symbol**	**Unit**	**Value**
Volumetric flowrate	Q	m^3/h	58.75
Liquid density	ρ	kg/m^3	1030.3
Acceleration due to gravity	g	m/s^2	9.81
Superficial velocity	v	m/s	1.26
Pump efficiency	η	-	0.7
Source Information			
Pressure at source (abs)	P_0	Pa	100000
Pressure at source (abs)	P_0	m	9.894
Height of source (Static head)	H_0	m	0
Pipe length	L	m	2
Destination Information			

Pressure at destination (abs)	P	Pa	100000
Pressure at destination (abs)	P	m	9.894
Height of destination	h	m	7
Pipe length	L	m	5

The bulk density was calculated as follows:

$$\rho_{bulk} = \rho_{solid} \times \frac{m_{solid}}{m_{total}} + \rho_{liquid} \times \frac{m_{liquid}}{m_{total}}$$

$$\rho_{bulk} = 1030\left(\frac{1450}{1452}\right) + 1200\left(\frac{2.4}{1452}\right)$$

$$\rho_{bulk} = 1030.3\frac{kg}{m^3}$$

Wherever units of Pascal were converted to units of meter, the following equation was used:

$$\mathrm{m} = \frac{\mathrm{Pa}}{\rho \mathrm{g}}$$

The head-loss due to fittings were calculated as follows:

$$\mathrm{h_f} = \frac{\mathrm{k_{suction}} \mathrm{v}^2}{2\mathrm{g}}$$

At pump suction side:

$$\mathrm{h_f} = \frac{2.3 \times (1.26)^2}{2 \times 9.81}$$

$$\mathrm{h_f} = 0.187$$

At pump discharge side:

$$\mathrm{h_f} = \frac{63.3 \times (1.26)^2}{2 \times 9.81}$$

$$\mathrm{h_f} = 0.284\ \mathrm{m}$$

Where the k-value was determined by using the following values found in (Moran, 2016).

Fitting type	k-value
long-radius bends	0.1
Open isolation valve	0.4
Tee (Flow from side branch)	1.2
Tee (Flow straight-through)	0.1
Swing check non-return valve	1
Sharp entry	0.5

The straight-run head-loss was determined from the pipe heuristics in Turton Table 11.8 (Richard Turton, 2018).

For liquid pump suction:

$$\Delta P = \frac{0.4 \text{ psi}}{100 \text{ ft}} = 90.48\frac{\text{Pa}}{\text{m}}$$

$$h_{sr} = \text{pressure drop } \times \text{pipe length}$$

$$h_{sr} = \frac{90.48}{9.81 \times 1000} \times 2$$

$$h_{sr} = 0.018 \text{ m}$$

For liquid pump discharge:

$$\Delta P = \frac{2.0 \text{ psi}}{100 \text{ ft}} = 452.4\frac{\text{Pa}}{\text{m}}$$

$$h_{sr} = \text{pressure drop } \times \text{pipe length}$$

$$h_{sr} = \frac{452.4}{9.81 \times 1000} \times 11$$

$$h_{sr} = 0.507 \text{ m}$$

From this, the total dynamic head was calculated:

$$\text{Dynamic head} = \text{ fitting headloss} + \text{straight run headloss}$$

At pump suction side:

$$\text{Dynamic head} = 0.187 + 0.018$$

$$\text{Dynamic head} = 0.205 \text{ m}$$

At pump discharge side:

$$\text{Dynamic head} = 0.28 + 0.22$$

$$\text{Dynamic head} = 0.51 \text{ m}$$

The total pump head was then calculated as follows:

$$H = \left(P + h + \text{dynamic head}_{\text{discharge}}\right) - (P_0 + h_0 - \text{dynamic head}_{\text{source}})$$

$$H = (9.89 + 7 + 0.51) - (9.89 + 0 - 0.187)$$

$$H = 7.71 \text{ m}$$

Then lastly, the pump shaft power was calculated:

$$P = \frac{Q\rho g H}{3.6 \times 10^{6}\eta}$$

$$P = \frac{58.76 \times 1030.3 \times 9.81 \times 7.71}{3.6 \times 10^{-6} \times 0.7}$$

$$P = 1.82 \text{ kW}$$

	Symbol	Unit	P-101	P-102	P-103	P-104	P-105	P-106	P-107	P-108	P-109	P-110	P-111	P-112	P-113	P-114	P-115	P-116	P-117
Process inputs																			
Volumetric flowrate	Q	m3/h	1.79	1.82	0.04	0.01	0.01	0.02	0.003	58.75	58.75	57.01	57.01	59.07	16.24	0.94	1.34	58.13	89.08
Liquid density	ρ	kg/m3	1030	1142.73902	1098.80837	1241.40986	1030	1030	1831.10404	1030.28	1030.28	1030.01	1030.01	1030.33	1038.45	1048.13	1031.54	1030.04	1030.04
Pump efficiency	η	-	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Pipe cross-sectional area	m		0.0004	0.0004	9.5726E-06	1.6239E-06	1.83761E-06	5.10684E-06	6.6239E-07	0.013	0.013	0.012	0.012	0.013	0.003	0.000	0.000	0.012	0.019
Pipe cross-sectional area	10^4m2		3.8295	3.8964	0.10	0.02	0.02	0.05	0.01	125.53	125.53	121.82	121.82	126.22	34.70	2.01	2.86	124.20	190.35
Pipe cross-sectional area (std)	10^4m2		3.441	3.441	0.3664	0.3664	0.3664	0.3664	0.3664	129.1	129.1	129.1	129.1	129.1	30.89	2.791	2.791	129.1	186.5
Pipe ID (Standard)	mm		20.93	20.93	6.83	6.83	6.83	6.83	6.83	128.2	128.2	128.2	128.2	128.2	62.71	18.85	18.85	128.2	154.1
Pipe velocity	m/s		1.44676935	1.47203655	0.33964095	0.05761766	0.065198933	0.181192382	0.02350194	1.2640933	1.264093295	1.226708409	1.226708	1.2709965	1.460528998	0.937205568	1.33116366	1.25064549	1.32682951
Source Information																			
Pressure at source (abs)	P0	Pa	100000	112331.297	103664.965	102070.299	102020.86	104142.763	126944.696	100000	135374.6939	130313.255	159118.7	100000	150936.1745	100000	100000	100000	300000
Pressure at source (abs)	P0	m	9.90	10.02	9.62	8.38	10.10	10.31	7.07	9.89	13.39	12.90	15.75	9.89	14.82	9.73	9.88	9.90	29.69
Height of source (Static head)	H0	m	0.5	0	0	0	0	0	0	0	0	0	1	2	0	1	0	0	0.5
Headloss due to fittings	hf	m	0.160	0.099	0.005	0.00015	0.00019	0.00151	0.00003	0.187	0.073	0.161	0.161	0.091	0.120	0.094	0.081	0.167	0.179
Straight run headloss	hsr	m	0.179	0.016	0.017	0.015	0.018	0.018	0.010	0.018	0.018	0.018	0.018	0.036	0.018	0.035	0.018	0.027	0.090
Headloss (Dynamic head)	hd	m	0.339	0.116	0.022	0.015	0.018	0.019	0.010	0.205	0.091	0.179	0.179	0.126	0.137	0.129	0.099	0.194	0.269
Suction pressure	Ps0	m	10.058	9.905	9.595	8.366	10.079	10.287	7.057	9.689	13.303	12.718	16.568	11.767	14.679	10.596	9.783	9.702	29.920
k-value	k	-	1.5	0.9	0.9	0.9	0.9	0.9	0.9	2.3	0.9	2.1	2.1	1.1	1.1	2.1	0.9	2.1	2
Straight run pressure drop	delP	m/m pipe	0.009	0.008	0.008	0.007	0.009	0.009	0.005	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.009
Pipe length	L	m	20	2	2	2	2	2	2	2	2	2	2	4	2	4	2	3	10
Destination Information																			
Pressure at destination (abs)	P	Pa	101473	101473	101473	101473	101473	101473	101473	100000	100000	101473	100000	100000	101473	101473	101473	180000	300000
Pressure at destination (abs)	P	m	10.04	9.05	9.41	8.33	10.04	10.04	5.65	9.89	9.89	10.04	9.90	9.89	9.96	9.87	10.03	17.81	29.69
Height of destination	H	m	1	6.41	6.41	6.41	6.41	0	6.41	7	3	11.41	7.00	2	11.41	11.41	11.41	0	0.5
Headloss due to fittings	hf	m	0.47	0.08	0.08	0.08	0.08	0.28	0.28	0.28	0.34	0.56	0.24	0.24	0.24	0.34	0.34	0.18	0.24
Straight run headloss	hsr	m	0.58	0.08	0.08	0.07	0.09	0.45	0.20	0.22	0.22	0.49	0.54	0.27	0.36	0.35	0.27	0.67	0.22
Headloss (Dynamic head)	hd	m	1.05	0.16	0.16	0.15	0.17	0.73	0.49	0.51	0.57	1.05	0.78	0.51	0.60	0.70	0.61	0.85	0.47
Discharge pressure	Ps0	m	12.09	15.62	15.99	14.89	16.62	10.77	12.54	17.40	13.46	22.50	17.68	12.40	21.97	21.98	22.05	18.67	30.65
			122158	175105	172309	181384	167930	108869	225342	175884	136066	227398	178597	125367	223786	225954	223136	188614	309751
k-value	k	-	5.4	0.9	0.9	0.9	0.9	3.3	3.3	3.3	4	6.5	2.8	2.8	2.8	4	4	2.1	2.8
Straight run pressure drop	delP	m/m pipe	0.045	0.040	0.042	0.037	0.045	0.045	0.025	0.045	0.045	0.045	0.045	0.045	0.044	0.044	0.045	0.045	0.045
Pipe length	L	m	13	2	2	2	2	10	8	5	5	11	12	6	8	8	6	15	5
Pump Power																			
Pump head	PH	m	2.032	5.715	6.390	6.528	6.541	0.487	5.488	7.713	0.160	9.787	1.107	0.636	7.288	11.379	12.268	8.964	0.734
Pump power	P	kW	0.015	0.046	0.001	0.00024	0.00023	0.00005	0.00012	1.817	0.038	2.237	0.253	0.151	0.479	0.044	0.066	2.089	0.262

14.4.2 Screw Conveyor SC-101 Slaked Lime

Needs to transport 43.8 m^3/d = 64.40 ft^2/h.

(KWS, 2018) was used to select the appropriate outer diameter (9 inches).

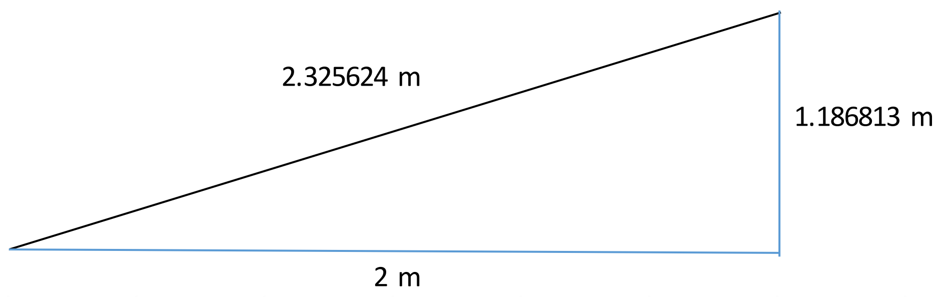
To calculate the power required to operate the spiral the capacity needs to be calculated using (Bechtel, 2018). The inner diameter was provided by the reference (Bechtel, 2018). The pitch was selected as half the outer diameter as the conveyer will be at a 30° incline and the efficiency for an incline is higher for a lower pitch (Bechtel, 2018).

$$Q = \frac{3.14 * (D - D_i)^2}{4} * s * N * sg * i * 60$$

$$Q = \frac{3.14 * (\ 2.286 - 0.5529)^2}{4} * 2.5 * 25 * 0.7 * \ 60$$

$$Q = \ 34.47 \frac{t}{h}$$

The length of transportation was calculated using the horizontal length of transport as well as how high it needs to be lifted. With Pythagoras's theorem the actual conveyor length was calculated.



From here the power can be calculated through (Bechtel, 2018) for a mechanical efficiency of 70%:

$$P = \frac{Q * L * K}{3600 * 102} * \frac{1}{\eta}$$

$$P = \frac{34.47 * \ 2.33 * 0.5}{3600 * 102} * \frac{1}{70\%}$$

$$P = 0.156 \ W$$

The friction coefficient (K) was obtained from (Bechtel, 2018) for stainless steel.

14.5 Centrifuge

The required flowrate is 1.66 $\frac{m^3}{h}$. A centrifuge was found with an acceptable capacity (F.N.S, 2019). The bowl diameter is 300mm. It has dimensions of (L x W x H) of 2610 x 800 x 1080 in mm.

14.7 Impellers

14.7.1 Vessel V-101 – Slaked Lime Agitator

As calcium carbonate is not very soluble in water it forms a suspension (Chem Libretexts, 2019). Thus, an axial flow impeller should be used to maintain the undissolved solids in suspension while still providing a maximum area of influence (Fogler & Gurmen, 2008). The radial flow impeller was not selected as it only has a great influence around its impeller not over the entire tank as the axial flow impeller does (James, Selecting Wastewater Mixers for Wastewater Treatment – Part 1, 2013). The mixture is not shear sensitive, therefo1re an aggressive option can be used (Rayneri, 2018). SEVIN with inlets was selected as it allows for some shear while still ensuring the entire vessel is mixed (Rayneri, 2018).

9. Sevin with inlets

(Rayneri, 2018)

Thus from (Rayneri, 2018) a Np and Nq value can be estimated (this is later validated using the Reynolds number)

$$N_p = 1.37$$
$$N_Q = 0.68$$

The d/D ratio needs to be between 0.3 and 0.5. The higher ratio influences the amount of shear

A d/D ratio was selected as 0.4 as, although the mixture is not particularly viscous, it requires that the calcium hydroxide be kept in suspension else settling will occur. This will result in dead volume in the mixer and loss of calcium hydroxide. The extra power required to operate the mixer is will reduce the likelihood of this occurring. The shear will ensure that the calcium hydroxide does not flocculate. This ratio is still small enough to ensure that the entire mixture is mixed as a too large impeller diameter will inhibit flow around the reactor (Chem Libretexts, 2019).

The density of the mixture was calculated:

$$\rho_{CaOH} = 2329.61 \frac{kg}{m^3}$$
$$\rho_{H_2O} = 998 \frac{kg}{m^3}$$

The volumetric flowrate of calcium hydroxide is calculated from the mass of calcium hydroxide divided by the density of calcium hydroxide.

$$Q_{CaOH} = \frac{m_{Ca(OH)_2}}{\rho_{CaOH}}$$

$$Q_{CaOH} = \frac{4109.2 \frac{kg_{Ca(OH)_2}}{d}}{2329.61 \frac{kg}{m^3}}$$

$$Q_{Ca(OH)_3} = 1.7639 \frac{m^3}{d}$$

$$Q_{Water} = 42 \frac{m^3}{d}$$

Total volume:

$$Q_{total} = Q_{Ca(OH)_3} + Q_{Water}$$

$$Q_{total} = 1.7639 + 42$$

$$Q_{total} = 43.76 \frac{m^3}{d}$$

Mixture density:

$$\rho_{mix} = \frac{\rho_{Ca(OH)_3} Q_{Ca(OH)_3} + \rho_{water} Q_{water}}{Q_{total}}$$

$$\rho_{mix} = \frac{2329.61(1.7639) + 998(43.76)}{43.76}$$

$$\rho_{mix} = 1051{,}67 \frac{kg}{m^3}$$

Concentration of calcium hydroxide:

$$C_{Ca(OH)_2} = \frac{m_{Ca(OH)_3}}{Q_{total}}$$

$$C_{Ca(OH)_2} = \frac{4109.20}{43.76}$$

$$C_{Ca(OH)_2} = 97.8 \frac{g}{L}$$

(James, Selecting Wastewater Mixers for Wastewater Treatment – Part 1, 2013) Suggests keeping the concentration of calcium hydroxide in-between 50 and 120 g/L to prevent carbonation or settling. For the mixing tank we are assuming uniform mixing, thus the whole tank will have a concentration below 120g/L. The aggressive agitation will ensure that this stays true so that the lowest amount of water can be used while safely avoiding settling.

If the amount of water drops by 15% the concentration becomes:

$$C_{Ca(OH)_3} = \frac{4109.20}{43.76 - 15\%(43.76)}$$

$$C_{Ca(OH)_3} = 110.5 \frac{g}{L}$$

Even if the influent fluctuations works its way to the dosing system in spite of the control algorithm, the concentration will still be within acceptable limits.

The scale of agitation factor (V_c) was then selected based on the relative viscosity (which was obtained to be below 20 (Senapati, Panda, & Parida, 2009)). Thus the scale of agitation factor should be selected as 1, but for ensuring that the mixture will not settle even under a shock load of calcium hydroxide the factor was selected as 3. This is unproblematic due to the high shearing capacity of the mixture.

The holding time was selected as 30 minutes as per the recommendation in (Suez, 2019).

Therefore, the volume is (Rayneri, 2018):

$$V_{mixing\ tank} = \frac{Q}{t_h}$$
$$V_{mixing\ tank} = 43.76 * 0.02083$$
$$V_{mixing\ tank} = 0.911747925$$

The height of the mixing tank can be estimated using the rules of thumb available in (James, Mixing 101: Optimal Tank Design, 2015). An L/D ratio between 0.6 and 1.4 is acceptable for a mixing tank. In the interest of saving land space, but still maintaining acceptable mixing conditions a 1.2 ratio was selected.

The specifications of the tank are as follows (Rayneri, 2018):

$$D = \left(4 * \frac{V_{mixing\ tank}}{\pi * 1.2}\right)^{\frac{1}{3}}$$
$$D = \left(4 * \frac{0.912}{\pi * 1.2}\right)^{\frac{1}{3}}$$
$$D = 0.99\ m$$

$$L = 1.2 * D$$
$$L = 1.2 * 0.99$$
$$L = 1.188\ m$$

The d/D ratio was previously selected as 0.4.

$$V_c = 3 * 6 \frac{\text{ft}}{\text{min}} * \frac{1}{196.85} \frac{m}{s}$$
$$V_c = 0.091 \frac{m}{s}$$

The pumping rate

$$Q = \frac{V_c \pi D^2}{4}$$
$$Q = \frac{0.091(\pi)(0.99)^2}{4}$$
$$Q = 0.07 \frac{m^3}{s}$$

The RPM (Rayneri, 2018):

$$N = \frac{Q}{N_Q d}$$
$$N = \frac{0.07}{0.68 * 0.396}$$

$$N = 1.67\frac{1}{s}$$

The power required to power the motor with an 80% efficiency (INDCO, 2013)

$$P = \frac{N_p \rho N^3 d^5}{0.8}$$

$$P = \frac{1.37 * 1051.7 * 1.67^3 * 0.396^5}{4}$$

$$P = 0.08\ kW$$

The Reynolds number is calculated and the N_p and N_Q are read off a chart (Rayneri, 2018). The calculation procedure is repeated for these new Np and N_Q values.

$$Re = \frac{d^2 N \rho}{\mu}$$

$$Re = \frac{0.396^2 * 1.67 * 1051.7}{0.9}$$

$$Re = 305.14$$

No baffles are recommended for suspensions (Brodkey & Hershey, 2003).

TK-102 Mixer

A low shear mixing setup is required to avoid shearing the microorganisms to death. The hydrofoil impeller is best suited for this purpose. No baffles are necessary as the aim is to maintain the biomass in suspension, not mix it (Brodkey & Hershey, 2003).

V-107 Urea

The PSVB blade was selected. This impeller is specifically suited to dissolving solutes into solvents. Shear is not a problem so aggressive options are available. Vc is 3 and the d/D ratio is 0.3.
The rest of the calculations are as above.

V-108 Diammonium Phosphate

The airfoil blade was selected using the same method as above.

V-109 Micronutrients

The PA four blade axial impeller was calculated to be most suitable.

V-110 Polymer

The PA four blade axial impeller was calculated to be most suitable.

The power values calculated for all the impellers are available in the costing section.

14.8 Belt Conveyor

The solids flowrate is 6.5 tonnes/day which is 0.27 tonnes/hour. Using the costing table in (Engineering ToolBox, 2019). For a 5 meter long conveyor the power required is 0.31 kW.

15 APPENDIX F – EQUIPMENT LIST AND SUMMARIES

Table 10: Equipment Summary for Critical Equipment Parameters

Heat Exchangers	**E-101**	**E-102**	**E-103**	**Blower**	**B-101**	**B-102**
Type	Shell-and-tube	Shell-and-Tube	Electric	**Type**	Air	Steam
Area (m2)	21.1	1.84		**Flow (m³/h)**	126.8	90.0
Duty (kW)	235.1	1014.8	181.9	-	-	-
Shell						
Temperature (°C)	30			-	-	-
Pressure (bar)	1			-		-
Phase	Liquid			-	-	-
MOC				-	-	-
Tube						
Temperature (°C)	17			-	-	
Pressure (bar)	1			-	-	-
Phase	Liquid			-	-	-
MOC				-	-	-
Reactors/Main Vessels	**R-101**	**V-111**	**F-101**	**C-101**	**Screen**	**S-101**
Temperature (°C)	37				**Spacing (mm)**	38.1
Pressure (bar)	1.02				**Area (m²)**	2
Orientation	Vertical				-	-
MOC	SS 304				-	-
Size						
Height/Length (m)					-	-
Diameter (m)					-	-
Conveyors	**CB-101**	**SC-101**	**SC-102**	**SC-103**	**SC-104**	**SC-105**
Length (m)	5.0	2.3	2.0	2.0	2.0	2.0
MOC	PVC	Stainless Steel	Stainless Steel	Stainless Steel	Stainless Steel	Stainless Steel
Tanks/Vessels	**TK-101**	**TK-102**	**TK-103**	**V-101**	**V-102**	**V-103**
Temperature (°C)	25	25	25	25	25	25
Pressure (bar)	1.4	1.3	1.35	2.5	1.4	1.3
Orientation	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical
MOC	CS	CS	CS	CS	CS	CS
Size						
Height/Length	3	3.46	8.43	3	3	3
Diameter (m)	0.5	2.88	25.28	2.2	1.1	0.66
Volume (m³)	0.37	22.52	4230	24.69	2.76	0.69
Pumps	**P-101**	**P-102**	**P-103**	**P-104**	**P-105**	**P-106**
Flow (t/d)	44.30	47.37	1.17	0.22	0.22	0.59
Fluid Density (kg/m³)	1030	1143	1099	1241	1030	1030

Power (shaft) (kW)	0.015	0.046	0.001	0.001	0.001	0.001
Type	C	PD				
Efficiency	0.7	0.7	0.7	0.7	0.7	0.7
MOC	CS	CS	CS	CS	CS	CS
Suction Pressure (m)	10.06	9.91	9.60	8.37	10.08	10.29
Discharge Pressure (m)	12.09	15.62	15.99	14.89	16.62	10.77

Tanks/Vessels	**V-104**	**V-105**	**V-106**	**V-107**	**V-108**	**V-109**	**V-110**	**Heater**	**H-101**		
Temperature (°C)	-	25	25	25	25	25	25	**Type**	Steam		
Pressure (bar)	-	1.1	1.1	1.1	1.1	1.1	1.1	**Duty (kW)**	832.9		
Orientation	-	Vertical	Vertical	Vertical	Vertical	Vertical	Vertical	**MOC**	SS		
MOC	-	CS	CS	CS	CS	CS	CS	-	-		
Size											
Height/Length		3						-	-		
Diameter (m)		0.3						-	-		
Volume (m^3)		0.04						-	-		
Pumps	**P-107**	**P-108**	**P-109**	**P-110**	**P-111**	**P-112**	**P-113**	**P-114**	**P-115**	**P-116**	**P-117**
Flow (kg/h)	0.08	1452.6	1452.6	1409.4	1409.4	1460.5	404.3	23.5	33.2	1437	32.2
Fluid Density (kg/m3)	1831	1030	1030	1030	1030	1030	1038	1048	1032	1030	1030
Power (shaft) (kW)	0.001	1.82	0.038	2.24	0.25	0.15	0.48	0.044	0.066	2.09	0.0003
Type											
Efficiency	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
MOC	CS	CS	CS	CS	CS	CS	CS	CS	CS	CS	CS
Suction Pressure (m)	7.06	9.69	13.30	12.72	16.57	11.77	14.68	10.60	9.78	9.70	30.10
Discharge Pressure (m)	12.54	17.40	13.46	22.50	17.68	12.40	21.97	21.98	22.05	18.67	30.65

Key:

C: Centrifugal Pump

PD: Positive displacement pump

CS: Carbon Steel

16 APPENDIX G – CAPITAL COST INFORMATION

Table 11: Equipment base case cost and correction factors used for equipment costing

Process Unit	Base Case Costs					Correction Factors			Scaling Factor (n)
	Size	Unit	Cost(US$)	Year	Reference	F_m	F_T	F_P	
Clarifier	300	m^2	460 000	CEPCI=1000	(Woods, 2007)			1	0.58
						1			
						1.3			
Pumps	1.82	kW	2 591	2001		1	1	1	0.34
Impeller	5	kW	21 921.06	2001	(Richard Turton, 2018)	1.8	1	1	0.6
Centrifuge	2.0	t/h	25 000	1982	(Ulrich, 1984)	1	1	1	-
Screw Conveyor	0.5	m^2	3 500	2001	(Richard Turton, 2018)	1	1	1	0.6

Table 12: Equipment cost and K-factors used for equipment costing

Process Unit	Base Case Costs								
	Size	Unit	K_1	K_2	K_3	$Cp°_{2001}$ (US$)	CEPCI	$Cp°_{2018}$ (US$)	Reference
Centrifugal Pump	1.82	kW	3.3892	0.0536	0.1538	2591	397	3966	(Richard Turton, 2018)
Positive Displacement Pump	2.24	kW	3.4771	0.135	0.1438	3483	397	5331	(Richard Turton, 2018)
Reactor mixers	7.5	kW	3.4092	0.3776	0.0030	6917.2	397	10508.14	(Richard Turton, 2018)
Process Vessels (Vertical)	22.52	m^3	3.4974	0.4485	0.1074	19977	397	30578	(Richard Turton, 2018)
Tanks (Fixed Roof)	4230	m^3	4.8509	-0.3973	0.1445	204352	397	310440	(Richard Turton, 2018)
Impeller	5	kW	3.4974	0.4485	0.1074	19976.56	397	30347.27	(Richard Turton, 2018)
Centrifuge	Graphical Method								(Ulrich, 1984)

Vacuum drum filter	6	M^2	4.8123	0.2858	0.042	114849	397	174472	
Screw Conveyor	0.5	M^2	3.6062	0.2659	0.1982				

16.1 Reactor

The reactor cost was determined using the cost of an anaerobic digestion tank available to purchase in industry and scaling it as a tank using the cost equations presented in Turton (Richard Turton, 2018). The Cost for a 4030 m^3 stainless steel reactor was found in literature including insulation, foundation and agitation (silo, 2019) in Danish Krone. The conversion rate to US dollars was found in 2012 to be 1USD:5.6586DKK (Exchange-Rates.org, 2019).

$$C_{reactor,2012,4030} = \frac{\text{DKK } 3300000}{5.6586\,\frac{\text{DKK}}{\text{USD}}}$$

$$C_{reactor,2012,4030} = \text{USD } 583183$$

$$C_{tank,2001} = 10^{(4.8509-0.3973\log V+0.1445(\log V)^2}$$

$$C_{tank,2001} = 10^{(4.8509-0.3973\log 4030+0.1445(\log 4030)^2}$$

$$C_{tank,2001} = \text{USD } 198041$$

$$C_{tank,2012} = \text{USD } 198041 \times \frac{CEPCI_{2012}}{CEPCI_{2001}}$$

$$C_{tank,2012} = \text{USD } 198041 \times \frac{585}{397}$$

$$C_{tank,2012} = \text{USD } 291824$$

$$C_{reactor,2012} - C_{tank,2012} = \text{USD } 291359$$

$$C_{reactor,2012,5058} = 10^{(4.8509-0.3973\log V+0.1445(\log V)^2} + C_{reactor,2012} - C_{tank,2012}$$

$$C_{reactor,2012,5058} = 10^{(4.8509-0.3973\log 5057.8+0.1445(\log 5057.8)^2} + \text{USD } 291359$$

$$C_{reactor,2018,5058} = \text{USD } 630301 \times \frac{CEPCI_{2018}}{CEPCI_{2012}}$$

$$C_{reactor,2018,5058} = \text{USD } 630301 \times \frac{603.1}{585}$$

$$C_{reactor,2018,5058} = \text{USD } 649802$$

16.2 Thickener

From the Area calculated in Appendix E one can then calculate the unit cost of the thickener from Rules of thumb in engineering practice (Woods, 2007). The values presented in this book attain to the CEPCI value of 1000 therefore the cost will adjusted to the current CEPCI value.

$$\frac{P_1}{P_2} = \left(\frac{A_1}{A_2}\right)^n$$

$$\frac{P_1}{\$460000} = \left(\frac{81.2}{300}\right)^{0.58}$$

$$P_1 = \$215633$$

In 2018 the average CEPCI Value was 603.1. Therefore in 2019 the unit cost of a thickener will be:

$$P_1 = \$215633 \times \frac{603.1}{1000}$$

$$P_1 = \$130049$$

16.3 Pumps

From the determined kW required per pump in section 5, equation A.1 from Turton (Richard Turton, 2018) was used to calculate the cost of each pump. The corresponding correction factors used are shown in Table 13 and K-factors used are shown in Table 13. The cost for each pump was obtained in the same manner. Sample calculations for P-101 (wastewater influent to dosing station pump) is shown below. Table 13 provides a summary of all pump costing values.

Cost of equipment in 2001 (base case cost):

$$\log_{10} C_p^0(2001) = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$

The minimum value for the capacity variable for pumps to be used in this equation, is 1 kW. In the case that the pump power was smaller than 1 kW, this minimum value of 1 kW was used in the equation and the calculated cost was scaled down by using a capacity ratio.

$$\log_{10} C_p^0(2001, 1\ kW) = 3.3892 \times 0.0536(\log_{10}(1)) + 0.1538((\log_{10}(1))^2$$

$$C_p^0(2001, 1\ kW) = US\$\ 2450$$

In order to scale the obtained cost to the correct pump size, a scaling factor from Timmerhause (Max S. Peters, 2003) of 0.34 was used.

$$C_p^0(2001) = C_p^0(2001, 1\ kW) \times \left(\frac{A}{A_{1kW}}\right)^{0.34}$$

$$C_p^0(2001) = 2450 \times \left(\frac{0.0146}{1}\right)^{0.34}$$

$$C_p^0(2001) = US\$\ 582.2$$

The unit cost was then scaled from 2001 to 2018 by using the obtained CEPCI values. The correction factors are multiplied by the cost to account for material of construction, pressure and temperature. In accordance with (Van Wyk, 2014) the material, temperature and pressure factor should be 1, 1 and 1, respectively based on the material of construction as well as the operating temperature and pressure.

The pumps are constructed from carbon steel and operates between 1 and 3 Bar(abs) and temperatures between 17 and 37 ^{o}C.

$$C_p^0(2018) = C_p^0(2001) \times \left(\frac{I_{CEPCI,2018}}{I_{CEPCI,2001}}\right) F_M F_T F_P$$

$$C_p^0(2018) = 582.2 \times \left(\frac{603.1}{397}\right) \times 1 \times 1 \times 1$$

$$C_p^0(2018) = US\$\,891$$

Table 13: All pump costing values

Pump	$	kW
P-101	891.3	0.0146
P-102	1616.2	0.0464
P-103	469.8	0.0012
P-104	269.8	0.0002
P-105	264.3	0.0002
P-106	154.7	0.00005
P-107	214.0	0.0001
P-108	3966.0	1.8175
P-109	1505.0	0.0376
P-110	5330.9	2.2374
P-111	2877.7	0.2530
P-112	1970.9	0.1507
P-113	3574.1	0.4785
P-114	1584.3	0.0437
P-115	1821.3	0.0659
P-116	3395.4	2.0892
P-117	2379.2	0.2622

16.4 Tanks and Hoppers

16.4.1 Hoppers

For the storage tank in question an estimated equipment cost can be determined using the formula below as obtained from (Richard Turton, 2018).

$$\log \mathrm{C_{p\,2001}^{0}} = \mathrm{K_1} + \mathrm{K_2} \log \mathrm{V_{V-101}} + \mathrm{K_3}[\log \mathrm{V_{V-101}}]^2$$

$\mathrm{C_{p\,2001}^{0}}$: Cost of equipment in 2001 in USD$

$\mathrm{K_1, K_2, K_3}$: Cost correlation constants

$\mathrm{V_{V-101}}$: Storage capacity (m^3)

From (Richard Turton, 2018) the cost correlations factors are 3.4974, 0.4485 and 0.1074 for K_1, K_2 and K_3, respectively. Since the required capacity of 24.68 m^3 is within the range of 0.3-520 m^3 there is no need to scale for capacity. By substituting the values into the given equation above the estimated equipment cost can be determined as shown below:

$$\log \mathrm{C}\mathring{\mathrm{p}}_{2001} = \mathrm{K}_1 + \mathrm{K}_2 \log \mathrm{V}_{\mathrm{V}-101} + \mathrm{K}_3[\log \mathrm{V}_{\mathrm{V}-101}]^2$$

$$\log \mathrm{C}\mathring{\mathrm{p}}_{2001} = 3.4974 + 0.4485 \log(24.69) + 0.1074[\log(24.69)]^2$$

$$\log \mathrm{C}\mathring{\mathrm{p}}_{2001} = 4.33$$

$$\mathrm{C}\mathring{\mathrm{p}}_{2001} = \$\ 21\ 389.8$$

To account for operating conditions that deviate from standard temperature and pressure (STP) conditions material-, temperature and pressure factors must be incorporated. In accordance with (Van Wyk, 2014) the material-, temperature and pressure factor should be 1, 1 and 1, respectively based on the material of construction as well as the operating temperature and pressure. The storage tank used to store the inventory of the calcium hydroxide can possibly be made of carbon steel whereas the operating temperature and pressure is 25°C and 2.5 bar(a). Additionally the cost must also be account for time by utilizing CEPCI values in the following manner:

$$\mathrm{C}\mathring{\mathrm{p}}_{2018} = \mathrm{C}\mathring{\mathrm{p}}_{2001}\left(\frac{\mathrm{I}_{\mathrm{CEPCI},2018}}{\mathrm{I}_{\mathrm{CEPCI},2001}}\right)\mathrm{F_M F_T F_P}$$

$$\mathrm{C}\mathring{\mathrm{p}}_{2018} = (21\ 389.8)\left(\frac{603.1}{397}\right)(1)(1)(1)$$

$$\mathrm{C}\mathring{\mathrm{p}}_{2018} = \$\ 32\ 494.2$$

Lastly the cost must be converted in Rand so that the estimated equipment cost is more applicable to a South African market:

$$\mathrm{C}\mathring{\mathrm{p}}_{2018} = (\$\ 32\ 494.2)\left(\frac{\mathrm{R}\ 15.39}{\$}\right)$$

$$\mathrm{C}\mathring{\mathrm{p}}_{2018} = \mathrm{R}\ 500\ 085.3$$

Following the same methodology sizing and costing was done for units V-102, V-103, V-105 and TK-101. Given below in Table 14Table 9 is a summary of the important parameters applicable to the unit in question

Table 14: Summary of the equipment cost for the various storage tanks (hoppers)

Parameter	V-101	V-102	V-103	V-105	TK-106
Volume (m^3)	24.69	2.76	0.69	0.04	0.37
$\mathrm{C}\mathring{\mathrm{p}}_{2001}$ (\$)	21 289.8	5200.3	4069.2	-	2107.5
F_M	1	1	1	1	1
F_T	1	1	1	1	1

F_P	1	1	1	1	1
Cp^0_{2018} (R)	500 000.3	121 580.5	62 625.4	488	48 823.8

16.4.2 Buffer Tank

16.4.3 Calamity Tank

From the volume calculated in section 5, equation A.1 from Turton (Richard Turton, 2018)was used to calculate the cost for the calamity tank.

Cost of equipment in 2001 (base case cost):

$$\log_{10} C_p^0(2001) = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$

$$\log_{10} C_p^0(2001) = 4.8509 - 0.3973(\log_{10}(4230)) + 0.1445(\log_{10}(4230))^2$$

$$C_p^0(2001) = US\$\ 204352$$

The unit cost was then scaled from 2001 to 2018 by using the obtained CEPCI values. The correction factors are multiplied by the cost to account for material of construction, pressure and temperature. In accordance with (Van Wyk, 2014) the material, temperature and pressure factor should be 1, 1 and 1, respectively based on the material of construction as well as the operating temperature and pressure. The calamity tank is constructed from carbon steel and operates between 1 and 1.4 Bar(abs) and temperatures between 17 and 37 ^{o}C.

$$C_p^0(2018) = C_p^0(2001) \times \left(\frac{I_{CEPCI,2018}}{I_{CEPCI,2001}}\right) F_M F_T F_P$$

$$C_p^0(2018) = 204352 \times \left(\frac{603.1}{397}\right) \times 1 \times 1 \times 1$$

$$C_p^0(2018) = US\$\ 310440.6$$

16.5 Bar Screen

From the volume calculated in section 5, equation A.1 from Turton (Richard Turton, 2018) was used to calculate the cost for the calamity tank.

$$\log_{10} C_p^0(2001) = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$

$$\log_{10} C_p^0(2001) = 3.8219 - 1.0368(\log_{10}(2)) - 0.605(\log_{10}(2))^2$$

$$C_p^0(2001) = US\$\ 12000$$

$$C_p^0(2018) = C_p^0(2001) \times \left(\frac{I_{CEPCI,2018}}{I_{CEPCI,2001}}\right)$$

$$C_p^0(2018) = 12000.1 \times \left(\frac{603.1}{397}\right)$$

$$C_p^0(2018) = US\$\ 18368.6$$

16.6 Mixers

TK-110 Mixer

The mixer costs were estimated using the method in (Richard Turton, 2018). The mixer in this case is of the impeller variety. The minimum power sizable is 5 kW, but the mixer power in this case is below that (0.09 kW).

So get a cost estimate for 5kW:

$$\log_{10} C^0_{p\,(2001)} = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$
$$\log_{10} C^0_{p\,(2001)} = 3.8511 + 0.7009(\log_{10}(5)) \; - 0.0003(\log_{10}(5))^2$$
$$C^0_{p\,(2001)} = US\$\, 21921.06 \; for\; 5kW$$

$$C_{p_{2001}} = C^o_{p_{2001}} F_M F_T F_p$$
$$C_{p_{2001}} = 21921.06 * 1.8 * 1 * 1$$
$$C_{p_{2001,\; 5kW}} = US\$\, 39457.90$$

In order to scale the obtained cost to the correct impeller size, a scaling factor (0.6 which is the rule of 16ths) from Turton of was used.

$$C_{p_{(2001),0.09kW}} = C_{p_{(2001,5\,kW)}} \times \left(\frac{A}{A_{5kW}}\right)^{0.6}$$
$$C_{p_{(2001),0.09kW}} = 39457.90 * \left(\frac{0.09}{5}\right)^{0.6}$$
$$C_{p_{(2001),0.09kW}} = US\$\; 3497.23$$

$$C_p(2018) = C_p(2001) \times \left(\frac{I_{CEPCI,2018}}{I_{CEPCI,2001}}\right)$$
$$C_{p_{(2018)}} = \; 3497.23\left(\frac{603.1}{397}\right)$$
$$C_{p_{(2018)}} = US\$\, 5312.79$$

16.7 Centrifuge

The centrifuge was sized graphically in (Ulrich, 1984). The graphical method is for a stainless steel centrifuge. This method is not ideal due to its graphical nature.

$$C^o_{p_{1984}} = US\$\, 25\,000$$

$$C^0_{p\,(2018)} = C^0_{p_{1984}} \times \left(\frac{I_{CEPCI,2018}}{I_{CEPCI,1984}}\right) F_M F_T F_P$$

$$C_{p_{(2018)}} = 25000\left(\frac{603.1}{315}\right)(1*1*1)$$
$$C_{p_{(2018)}} = US\$\,47865.08$$

16.8 Screw Conveyor

SC-101

The conveyor costs were estimated using the method in (Richard Turton, 2018). The conveyor in this case is of the screw variety. The minimum area is 0.5 m^2, but the screw conveyor area in this case is below that (0.04 m^2). Therefore size for the 0.5m^2 and scale-down.

$$\log_{10} C^0_{p_{(2001)}} = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$
$$\log_{10} C^0_{p_{(2001)}} = 3.6062 + 0.2659(\log_{10} 0.5) + 0.1982(\log_{10} 0.5)^2$$
$$C^0_{p_{(2001)}} = US\$\,3500.39$$

$$C_{p_{2001}} = C^o_{p_{2001}} F_M F_T F_p$$
$$C_{p_{2001}} = 3500.39(1*1*1)$$
$$C_{p_{2001}} = US\$\,3500.39$$

Using the rule of 6/10

$$C_{p_{(2001),}} = C_{p_{(2001)}} \times \left(\frac{0.0410}{0.5}\right)^{0.6}$$
$$C_{p_{(2001),}} = 3500.39 * \left(\frac{0.0410}{0.5}\right)^{0.6}$$
$$C_{p_{(2001),}} = US\$\,1\,186.53$$

16.9 Belt Conveyor

$$\log_{10} C^0_{p_{(2001)}} = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$

$$\log_{10} C^0_{p_{(2001)}} = 3.6062 + 0.2659(\log_{10} 4) + 0.1982(\log_{10} 4)^2$$
$$C^0_{p_{(2001)}} = 18856.28 = C_{p_{(2001)}}$$

$$C^0_{p_{(2018)}} = C^0_{p_{2001}} \times \left(\frac{I_{CEPCI,2018}}{I_{CEPCI,2001}}\right)$$
$$C_{p_{(2018)}} = 1\,186.53\left(\frac{603.1}{397}\right)$$
$$C_{p_{(2018)}} = US\$\;28\,645.40$$

16.10 Foam Trap

Actual A value below minimum so the equipment needs to be scaled up.

$$\log_{10} C^0_{p\,(2001)} = K_1 + K_2(\log_{10} A) + K_3(\log_{10} A)^2$$

$$\log_{10} C^0_{p\,(2001)} = 4.1 + 0.4999(\log_{10} 0.9) + 0.0001(\log_{10} 0.9)^2$$

$$C^0_{p\,(2001)} = US\$\ 11\ 943.35$$

$$C_{p\,(2001),} = C^0_{p\,(2001)} F_M F_T F_p$$

$$C_{p\,(2001),} = US\$\ 11\ 943.35\ (1.8 * 1 * 1)$$

$$C_{p\,(2001),} = US\$\ \ 21\ 498.02$$

Now scale down for size

$$C_{p\,(2001),actual\ size} = C_{p\,(2001)} \times \left(\frac{0.071}{0.9}\right)^{0.6}$$

$$C_{p\,(2001),} = 21\ 498.02 * \left(\frac{0.071}{0.9}\right)^{0.6}$$

$$C_{p\,(2001),} = US\$\ 7096.62$$

16.11 Determination of fixed capital investment

Various capital requirements of a plant can be estimated for the processing plant using the factors presented in Table 15. These factors are based on the price of the total price of delivered equipment within the plant.

Table 15: Estimation of Total Capital Investment

Cost Component	Factor of Delivered Equipment Cost
Direct Costs	
Purchased equipment delivered	1
Equipment installation	0.39
Piping (Installed)	0.31
Instrumentation and controls	0.13
Electrical (Installed)	0.1
Utilities	0.35
Off-sites	0.2
Building (Including Services)	0.29
Site preparation	0.06
Total Direct Cost	2.83
Indirect Cost	
Design, engineering and supervision	0.32
Construction expenses	0.34
Total Indirect Cost	0.66
Contractors fee	*5%

Contingency	*25%
Total Fixed Capital Cost	4.537
Working capital	**15%
Total Capital Cost	5.34
***% Total fixed capital cost**	
****%Total capital cost**	

Adapted from Peters and Timmerhaus (Max S. Peters, 2003)

Table 16:

Item	Multiplying Factors
Direct Operating Cost	
Raw materials	C_{RM}
Waste treatment	C_{WT}
Utilities	C_{UT}
Operating labour	C_{OL}
Direct supervisory and clerical labour	0.18 C_{OL}
Maintenance and supplies	0.06FCI
Laboratory charges	0.009FCI
Patents and royalties	0.03COM
Total Direct Operating Cost	C_{RM}+ C_{WT}+ C_{UT}+1.33 C_{OL}+0.03COM+0.069FCI
Fixed Operating Cost	
Local taxes and insurance	0.032FCI
Plant overhead costs	0.708 C_{OL}+0.036FCI
Total Fixed Operating Cost	0.708 C_{OL}+0.036FCI
General Operating Cost	0.708 C_{OL}+0.068FC
Administration costs	0.177 C_{OL}+0.009FCI
Distribution and selling costs	0.11COM
Research and development	0.05COM
Total General Operating Cost	0.177 C_{OL}+0.009FCI+0.16COM
Total Operating Cost	C_{RM}+ C_{WT}+ C_{UT}+2.215 C_{OL}+0.19COM+0.146FCI

17 APPENDIX H – OPERATING COST INFORMATION

17.1 Operating Labor

The operating labor cost for the effluent treatment plant was determined from the following equation: (Richard Turton, 2018)

$$N_{OL} = \left(6.29 + 31.7P^2 + 0.23N_{np}\right)^{0.5}$$

Where:

N_{OL}= number of operators per shift

P=number of solid handling units

N_{np}= number of processing steps

It was assumed that the dosing was equivalent to one unit due to the small proximity of the system, this then put the number of processing steps to be 11.

Table 17: Number of processing steps table

Equipment Type	Number of Equipment	N_{np}
Centrifuge	1	1
Conveyor Belt	1	1
Exchangers	2	2
Heater	2	2
Dosing Station	1	1
Drum Filter	1	1
Reactor	1	1
Thickener	1	1
	Total	10

$$N_{OL} = (6.29 + 31.7(0)^2 + 0.23 \times 10)^{0.5}$$

$$N_{OL} = 2.93$$

To ensure that there are always 2.97 people per shift during the plants operation.

$$\text{Operating labor} = 2.93 \times 4.5$$

Rounding up to nearest number of employees yields:

$$\text{Operating labor} = 14$$

According to (Payscale, 2019) the average operator salary per annum is R127 616.

$$C_{LB} = 14 \times \text{R}127616$$

$$C_{LB} = \text{R}1786624$$

17.2 Waste Treatment

The effluent treatment process produces a sludge of 0.25% by weight solids. This equates to 6.5t/d of waste from the process. It was assumed that the sludge handing company has 2 trailers for the trunk in use therefore continuous operation can be assumed between dumping trips. So that one trailer may be filled whilst the other is being transported and emptied. It was also assumed that the truck was charged at R3.5

$$\text{Disposals per day} = \frac{\frac{6.5\text{t}}{\text{d}}}{5\text{t}} = \frac{1.3}{\text{d}}$$

$$C_{WT} = 1.3 \times 60\text{km} \times \frac{\text{R}3.5}{\text{km}} \times 343\text{d}$$

$$C_{WT} = \text{R}93639$$

17.3 Raw Materials

Table 18: Summary table of the raw materials to be purchased

Compound	Mass Flowrate (kg/d)	Cost Index (R/kg)	Cost (R/yr)
$Ca(OH)_2$	4109.2	1.9	2 735 741
CH_4N_2O	263.4	3.7	341 016.6
$(NH_4)_2HPO_4$	78.9	4.8	133 432.0
$FeCl_3$	43.7	5.4	62 416.0
Polymer	2.86	45	45 096.5

17.4 Utilities

Table 19: Summary of duties of the various electrical powered units

Unit	Power(kW)	Unit	Power(kW)	Unit	Power(kW)
Draft Tube mixers	30.00	P-102	1.10	P-113	0.48
Electric heater	181.91	P-103	0.01	P-114	0.04
TK-102 Mixer	0.09	P-104	0.05	P-115	0.07
V-106 Mixer	0.08	P-105	0.00	P-116	2.09
V-107 Mixer	0.02	P-106	0.00	P-117	0.26
V-108 Mixer	0.00	P-107	0.00	SC-101	0.00
V-109 Mixer	0.01	P-108	0.00	SC-102	0.00
V-110 Mixer	0.00	P-109	0.00	SC-103	0.00
Centrifuge	5.80	P-110	1.82	SC-104	0.00
Drum filter	1.10	P-111	0.04	SC-105	0.00
P-101	0.01	P-112	2.24	Total electricity requirement's	226.52

$$C_{UT} = C_{Ele} + C_{processwater} + C_{Air}$$

$$C_{Ele} = P_{Ele} \times 24h \times \frac{1.3R}{kWh} \times 350d$$

Table 20: Summary of the utilities used to supplement process requirements

Compound	Flowrate (m^3/d)	Cost Index (R/m^3)	Cost (R/yr)
Process water	6.18	7	15 158.3
Instrument air	3042.5	0.55	586 350.6
[redacted]			
	Heat duty (kW)	Electricity cost (R/kWh)	Cost (R/yr)
Energy	181.9	1.30	1 988 618.1

17.4.1 Thickener Polymer

From Shea the optimal polymer dose is 1.5 mg/g of TSS in the thickener, therefore the amount of cationic polymer needed for the stream would be, assuming the polymer associates itself with the solids in the recycle.

Polymer required:

$$\dot{m}_{10,\text{nonrecycledTSS}} = \dot{m}_{10,\text{TSS}} - \dot{m}_{15,\text{TSS}} - \dot{m}_{20,\text{TSS}}$$

$$\dot{m}_{10,\text{nonrecycledTSS}} = 1904 \frac{\text{kg}}{\text{d}}$$

$$\dot{m}_{\text{cationic polymer}} = \omega \times \dot{m}_{10,\text{nonrecycledTSS}}$$

$$\dot{m}_{\text{cationic polymer}} = 1.5 \frac{\text{mg}}{\text{gTSS}} \times 1904 \frac{\text{kg}}{\text{d}}$$

$$\dot{m}_{\text{cationic polymer}} = 2.9 \frac{\text{kg}}{\text{d}}$$

$$\dot{m}_{\text{cationic polymer,90\%}} = \frac{2.9 \frac{\text{kg}}{\text{d}}}{90\%}$$

$$\dot{m}_{\text{cationic polymer,90\%}} = 3.2 \frac{\text{kg}}{\text{d}}$$

$$R_{\text{cationic polymer,90\%}} = 3.2 \frac{\text{kg}}{\text{d}} \times \text{R}45.00$$

$$R_{\text{cationic polymer,90\%}} = \frac{\text{R}143}{\text{day}}$$

18 APPENDIX I – 8 ECONOMIC ANALYSIS INFORMATION

All values are in xR1000

				Sales		Cash Flow	Depreciation	Tax		After Tax	Cum	Disc	Cum
Year	Plant	Working	Total	Sales	Cost		Plant	Profit	Tax	Cash Flow	Cash Flow	Cash Flow	DCF
0										0	0	0	0
1	-79 167		-79 167			-79 167				-79 167	-79 167	-71 970	-71 970
2	-79 167	-27 954	-107 121			-107 121				-107 121	-186 288	-88 530	-160 500
3				41 948	-58 838	-16 890	-17 813			-16 890	-203 178	-12 690	-173 190
4				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-220 069	-11 536	-184 726
5				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-236 959	-10 488	-195 214
6				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-253 850	-9 534	-204 748
7				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-270 740	-8 667	-213 415
8				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-287 630	-7 879	-221 295
9				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-304 521	-7 163	-228 458
10				41 948	-58 838	-16 890	-17 813	-34 703		-16 890	-321 411	-6 512	-234 970
11				41 948	-58 838	-16 890		-34 703		-16 890	-338 302	-5 920	-240 890
12				41 948	-58 838	-16 890		-16 890		-16 890	-355 192	-5 382	-246 272
13				41 948	-58 838	-16 890		-16 890		-16 890	-372 082	-4 893	-251 164
14				41 948	-58 838	-16 890		-16 890		-16 890	-388 973	-4 448	-255 612
15				41 948	-58 838	-16 890		-16 890		-16 890	-405 863	-4 043	-259 656
16				41 948	-58 838	-16 890		-16 890		-16 890	-422 753	-3 676	-263 331
17				41 948	-58 838	-16 890		-16 890		-16 890	-439 644	-3 342	-266 673
18				41 948	-58 838	-16 890		-16 890		-16 890	-456 534	-3 038	-269 711
19				41 948	-58 838	-16 890		-16 890		-16 890	-473 425	-2 762	-272 473
20				41 948	-58 838	-16 890		-16 890		-16 890	-490 315	-2 511	-274 983
21				41 948	-58 838	-16 890		-16 890		-16 890	-507 205	-2 282	-277 266
22				41 948	-58 838	-16 890		-16 890		-16 890	-524 096	-2 075	-279 341
23				41 948	-58 838	-16 890		-16 890		-16 890	-540 986	-1 886	-281 227
24				41 948	-58 838	-16 890		-16 890		-16 890	-557 877	-1 715	-282 942
25		27 954	27 954	41 948	-58 838	11 063		-16 890		11 063	-546 813	1 021	-281 921
26	15 833		15 833			15 833				15 833	-530 980	1 329	-280 592
											NPV	-280 592	