Production of Methane from Anaerobic Digestion of Molasses

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Certificate

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Dedication

We dedicate this project to our parents, who provide a carefree environment for us, so that we can concentrate on our study. Although they hardly understood the subject matter of our project, our parents were willing to stand behind every decision we made.

We are at the verge of graduating from this great institution, all because of the hard work our parents put into our bringing. We wouldn’t have been here without them. We thank for all the times they have supported us and took care of us. Once again, this project is for you all.

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Nomenclature and Terminology

Glossary

Acidogenic – acid producing.

Ambient - outside air temperature.

Anaerobic – in the absence of oxygen microbes breakdown organic material (i.e. molasses slurry).

Anaerobic bacteria – microbes whose metabolisms require the absence of oxygen to survive.

Anaerobic digestion – the breakdown of molasses slurry (organic material) in the absence of oxygen, (methane producing bacteria are most active in two temperature ranges, 35 to 40°C and 55 to 60°C, depending upon type)

Biogas – the gas produced from decomposition of molasses slurry in an anaerobic digester consisting of 60-80 percent methane, 30-40 percent carbon dioxide, and other trace gases such as hydrogen sulfide, ammonia and hydrogen.

Complete mix digester (*mixed flow reactor*) - a tank designed above or below ground as part of a molasses slurry management system to handle molasses slurry containing no solids. The digester is heated and mixed mechanically or with gas-mixing systems to keep the solids suspended. This maximizes biological activity for destruction of volatile solids, methane production and odor reduction.

Digester – a sealed container or tank, where the biological digestion can occur of molasses slurry and biogas formed.

Effluent – organic liquid and solid material (slurry) leaving a digester.

Feedstock – liquid and solid material fed to the digester, usually molasses slurry, also known as influent.

Hydraulic retention time (HRT) – the average length of time the liquid influent remains in the digester for treatment.

Influent – liquid and solid material fed to the digester, usually molasses slurry.

Loading rate – the total amount of solids and liquids fed to the digester daily.

Molasses slurry – consists of urine and feces, wasted feed and bedding collected to put into the digester as influent.

Mesophilic – the temperature range of 35 to 45°C in which hydrogenic microbes thrive.

Methane – a combustible gas produced by anaerobic digestion, also the principal component of natural gas.

Methanogenic – methane producing microbes.

Microturbine – a small-scale gas turbine generation system to combust gas and generate electricity.

Net metering – an agreement with the utility company to purchase the electricity produced by the digester system at a rate equal to the farm electricity purchase rate.

Psychrophilic – less than 20°C.

Settled solids – the separated molasses slurry solids which settle to the bottom of the digester.

Slurry – the mixture of molasses and water processed preceding the actual process.

Temperature-phased anaerobic digester (TPAD) – two tanks designed as part of a molasses slurry management system. The digesters are heated, the first digester in the mesophilic temperature range and the second digester in the thermophilic temperature range. This will maximize biological activity for the destruction of volatile solids, methane production and odor reduction.

Thermophilic – temperature range of 50 to 60 °C where certain methanogenic bacteria are most active, the greatest pathogen destruction occurs in this temperature range.

Upflow anaerobic sludge blanket reactor (UASB) – an anaerobic digester based on the principles of a three-phase separator. It allows for aggregation of bacterial colonies at the bottom, liquid slurry in the middle and gas at the top. UASBs are shown in several studies to give the highest amount of methanogenic conversion among all digester types.

Volatile acids – these are produced in the digester by acid-forming bacteria and then used by the methane-forming bacteria to produce methane.

Volatile solids – the organic matter in molasses slurry which can be converted to gas.

Volatile solids loading rate – the total amount of volatile solids fed to the digester daily (note: volatile solids are what the microbes use to make methane gas).

Abbreviations

AD – Anaerobic Digester

COD – Chemical Oxygen Demand

HRT – Hydraulic Retention Time

M_n – Material feeds

MFR – Mixed Flow Reactor

PKR – Pakistani Rupee

RAS – Return Activated Sludge

SRT – Solids Retention Time

UASB – Upflow Anaerobic Sludge Blanket Reactor

USD – US Dollar

VSS – Volatile Suspended Solids

Units

Btu – British thermal units (a unit of energy defined as the amount of heat required to raise the temperature of 1 pound of water 1 °F)

mcfd (millions ft^3/day) – million cubic feet per day

gpd – gallons per day

kW – kilowatt

kWh – kilowatt hours

tpd – tonnes per day

Abstract

Molasses has been identified as one of the most potent sources of fermentable sugar among agri-based industrial waste. Therefore, in the wake of the current energy crisis in the country it was decided to produce methane from a diluted feed based on molasses. The two-stage process consisted of two reactors, the first-stage hydrogenic reactor (MFR) operated at pH 5.5 and 35°C and the second-stage methanogenic reactor (UASB) functioned at pH 7.0 and 55°C. Condition specific flora of methanogenic and hydrogen-producing bacteria was utilized. The first reactor had a HRT of 13 hours during which it generates H_2 that is routed to the second reactor be used as substrate. In the second stage of the process, methane would also be produced from the effluent of the first reactor with a production rate of 3.5 t/d at the optimum HRT of 33 hours. The combined sucrose conversion efficiency of the two process was found to be 73.4%.

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Introduction

Pakistan is an energy deficient country. Our gas reserves in particular are seeing drastic decline. It has been estimated that our current production capacity of 4 mcfd would decline to half its size; to 2 mcfd by 2020. The demand on the other hand would surge to 8 mcfd. This would mean a shortfall of an astounding 6 mcfd. Imagine the scenario in winters then.

While, steps must be taken to diversify our energy mix for the long term. In the meantime, however, attention must be focused on bridging the gap between demand and supply. The government has recently initiated the import of LNG to curtail the natural gas shortfall in the country. Nevertheless, this step falls heavily on the national exchequer. Other avenues must be explored to bolster the plummeting natural gas supply.

In this project we have looked at one such way to mitigate this difference between demand and supply. We have exploited the fact that Pakistan is an agri-based economy with a considerable portion of the arable land dedicated towards sugarcane cultivation. Consequently, Pakistan is the 15th largest producer of sugar in the world and as a result produces around 25 million tons of molasses annually. We have diverted a portion of this molasses to produce cost-effective methane from its anaerobic digestion. This methane would, in turn, be purified to a degree at which it can be released into the national supply network, used as fuel for cars or even a source of electricity generation.

In the following pages, you will see a brief overview of the process, followed by designing of all the critical equipment of the process. Succeeding plant design are costing and conclusion.

1.1. Historical developments of anaerobic digestion technology

The appearance of flickering lights emerging from below the surface of swamps was noted by Plinius (van Brakel 1980) and Van Helmont recorded the emanation of an inflammable gas from decaying organic matter in the 17th Century. Volta is generally recognized as putting methane digestion on a scientific footing. He concluded as early as 1776 that the amount of gas that evolves is a function of the amount of decaying vegetation in the sediments from which the gas emerges, and that in certain proportions, the gas obtained forms an explosive mixture with air.

In 1804 - 1810 Dalton, Henry and Davy established the chemical composition of methane, confirmed that coal gas was very similar to Volta's marsh gas and showed that methane was produced from decomposing cattle manure. France is credited with having made one of the first significant contributions towards the anaerobic treatment of the solids suspended in waste water. In 1884 Gayon, a student of Pasteur, fermented manure at 35°C, obtaining 100 liters of methane per m of manure. It was concluded that fermentation could be a source of gas for heating and lighting. It was not until towards the-end of the 19th Century that methanogenesis was found to be connected to microbial activity. In 1868, Bechamp named the "organism" responsible for methane production from ethanol. This organism was apparently a mixed population, since Bechamp was able to show that, depending on the substrate, different fermentation products were formed. In 1876, Herter reported that acetate in sewage sludge was converted stoichiometrically to equal amounts of methane and carbon dioxide (Zehnder 1978, 1982).

As early as 1896, gas from sewage was used for lighting streets in Exeter, England, while gas from human wastes in the Matinga Leper Asylum in Bombay, India, was used to provide lighting in 1897. Then, in 1904, Travis put into operation a new, two-stage process, in which the suspended material was separated from the wastewater, and allowed to pass into a separate "hydrolyzing" chamber. In 1906, Sohngen was able to enrich two distinct acetate utilizing bacteria, and he found that formate and hydrogen, plus carbon dioxide, could act as precursors for methane.

On the applied side, Buswell began studies of anaerobic digestion in the late 1920s and explained such issues as the fate of nitrogen in anaerobic digestion, the stoichiometry of reaction, the production of energy from farm wastes and the use of the process for industrial wastes (Buswell and Heave 1930; Buswell and Hatfield 1936).

Barker's studies contributed significantly to our knowledge of methane bacteria, and his enriched cultures enabled him to perform basic biochemical studies (Barker 1956). Schnellen (1947) isolate two methane bacteria: Methanosarcina barker) and Methanobacterium formicicum which are still studied.

Heating digestion tanks made practical use of the methane produced by the anaerobic process. It is of interest to note that methane gas was collected in Germany in 1914-1923 and used to generate power for biological treatment of plants, as well as for the cooling water from the motors being used to heat the digestion tanks.

Numerous additional studies led to a better understanding of the importance of seeding and pH control in the operation of anaerobic digestion systems. Much of this work is still relevant today, and those who are developing biogas as an energy source would gain much from review of this earlier work.

1.2. Present interest in anaerobic digestion

There is an increased recognition, in both developing and industrial countries, of the need for technical and economic efficiency in the allocation and exploitation of resources. Systems for the recovery and utilization of household and community wastes are gaining a more prominent place in the world community. During the last years, anaerobic fermentation has developed from a comparatively simple technique of biomass conversion, with the main purpose of energy production, into a multi-functional system (Prawit, Thang and Angelidaki):

:

a) Treatment of organic wastes and wastewaters in a broad range of organic loads and substrate concentrations;

b) Energy production and utilization;

c) Improvement of sanitation; reduction of odors;

d) Production of high quality fertilizer.

R & D has shifted from basic studies on anaerobic fermentation of quasi-homogeneous substrates, with contents of organic solids in the range of about 5 - 10%, to the digestion of more complex materials that need modified digester designs. The main fields of R & D activities are:

a) Fermentation at high organic loadings;

b) High rate digestion of diluted waste waters of agro-industries including substrate separation during fermentation; immobilization of the microorganisms;

c) Fermentation and re-use of specific materials in integrative farming systems;

d) Biogas purification;

e) Simple but effective digested design/construction of standardized fermenters;

f) Domestic waste water treatment.

Anaerobic digestion with high organic load can be performed when the concentration of methanogenic bacteria is kept at a high level. Pilot experiments, with mixtures of slaughterhouse waste water and cattle manure, succeeded in reducing retention times from about 20 to 8 - 10 days, (Prawit, Thang and Angelidaki) by a specific mixing technique, which allowed the mixture from time to time to separate. By this method, the liquid phase is enriched with dissolved organic matter, which is brought into contact with solid material, containing a relatively high concentrations of active bacteria.

Dissolved organic compounds can normally be degraded much faster than solid materials in suspension. If the retention times for dissolved and suspended components can be adjusted separately, the overall process can be performed at higher rates. Similar techniques are under investigation or implementation into large scale application in most of the countries which perform biogas R & D activities and biogas promotion programs. For example, in the Netherlands intense work is done to reduce the concentration of organic matter in the digested materials, and to reduce the volume of liquid effluents of agricultural activities. In that country, with its intensive animal production, problems of soil and groundwater pollution become more and more severe - a situation similar to other countries with intensive agricultural production.

A two-step system is being developed and tested for agricultural solid wastes (greenhouse waste, organic fractions of municipal refuse, cannery waste, grass clippings). The first step is a batch type hydrolytic/acidic unit, in which percolation water is circulated. The percolation water is anaerobically treated in the second step, and recycled to the percolation unit. The retention time of the waste in the first step depends on the digestibility of the raw material, and can take several months.

Another system is being tested for treatment of the organic fraction of municipal refuse. After mixing with recirculated water and subsequent maceration, the waste is pumped into the first step reactor. Here the conditions (temperature 37C, hydraulic retention time 12 - 24 h) are such that a very efficient microbial population develops, that degrades cellulose. This population, in which ciliated play an important role, resembles the population in the paunch of ruminants. After passing the first step reactor, the mixture is mechanically drained, the liquid fraction is anaerobically treated (e.g. in an UASB-type reactor) and recirculated to the mixture tank. The solid fraction is partly forced back into the first step reactor, the remainder being discharged. To cope with water shortage and water pollution in the medium/long-term, a 6 year R & D project for water re-use and energy recovery by biogas production has been implemented in Japan since 1985, under the sponsorship of the Ministry of International Trade and Industry ("Aqua Renaissance 90"). The object of this project is to establish the technology to ensure a low cost treatment of industrial waste water, sewage etc. to enable re-use of treated water by utilizing a big-reactor - working at high concentrations - coupled with membrane-separation techniques, and which allows the efficient production of methane and other useful resources as well.

The main areas of this project are:

a) Microorganisms and big-reactors (fixed beds, fluidized beds, two phase binary tank, UASB-processes);

b) Membrane techniques: materials (organic polymers and ceramics) and modules;

c) Control and sensor system: a direct measurement of activity of microorganisms to control and optimize methane production;

d) Total water treatment system: a technology to integrate all the above methods.

By studying the structure of the hierarchy that promotes biogas digestion system in some of the Developing Countries, which is the key to the efficient and wide distribution of biogas plants in those countries, officials can profit for their own country. This review summarizes the latest developments in anaerobic digestion applicable to Developing Countries, as reported in English language publications up to the year 1990, and the lessons from newly developed systems can be applied in other countries. Sharing the new ideas and their economic benefits, especially for the uses of digested slurry, can be beneficial to most Developing Countries.

Although the problems of stratospheric ozone depletion, (the Greenhouse Effect) and climatic changes, resulting from deforestation and wrong treatment of the environment, have not yet reached the same level of public recognition as toxic waste treatment, more and more people are becoming aware of and concerned about them. These problems are dramatic new reminders that we live on a valuable planet, and we have to think and act in consort and deal in a global, integrated, way with all our organic wastes as well as the woods and the forests.

1.3. Context of Usage of Methane

Among fossil fuels, currently methane seems to be the one that will be exploited more and more in the near future, thanks to its relative abundance and thanks to the fact that it is relatively ‘clean’. Its molecule is made up of 4 atoms of hydrogen and one of carbon ($CH4$): on burning, it is the hydrocarbon that releases the smallest amount of carbon and it is for this reason that it is less harmful for the environment. Its $CO2$ emissions are 25% lower than petrol, 16% lower than Liquid Propane Gas, 30% lower than diesel and 75% lower than carbon. Its capacity to form ozone is 80% less than petrol and 50% less than diesel and Liquid Propane Gas. Moreover, the combustion emissions do not contain carbonaceous residues, benzene and microscopic dusts (PM10), contrary to petrol and diesel oil. Among all the fossil fuels, methane is surely the most ‘ecological’. The use of methane is expected to increase greatly in the near future. The natural gas reserves that are of ‘geological’ origin are estimated to be sufficient for 60-70 years and they are mostly concentrated in the areas surrounding the Persian Gulf. Much smaller amounts are currently obtained from waste products of zooculture, with the use of anaerobic digesters that enable the production of methane from animal sewage. Other small quantities can be obtained from self-produced methane in abandoned carbon mines; here, this naturally produced gas is tapped and at the same time is prevented from dispersing in the surroundings.

1.4. Usage of Methane in Pakistan

The methane used in Pakistan is largely factored into two types, *Natural Gas* and *Compressed Natural Gas*. The former is pumped as a domestic and industrial fuel from the gas fields in Sindh and Balochistan like Sui and Mari. While the latter is the compressed variant of natural gas and is utilized mainly in internal combustion engines.

Pakistan has a demand of around 4mcfd natural gas per day. Much of it goes to the burgeoning CNG sector which caters to the needs of a fleet of 2.85 million *Natural Gas Vehicles* (NGVs), the second largest in the world. The remaining gas is consumed as fuel by many households and industrial setups.

Due to unmitigated increase in natural gas application and demand, and the belief in *endless reserves* of natural gas by top leadership. Much of Pakistan’s energy profile has converted to gas; from power generation to vehicle fuel, it is everywhere. Consequently, there is a consistently gaping gulf between supply and demand of natural gas that has resulted in cyclic spells of load-shedding. Specifically during winters, when the situation is quite dire and gas is available for only 6 hours per day on average.

2 Chapter 2

Literature Overview

2.1. Introduction

Anaerobic digestion is a biological process in which organic matter is decomposed in the absence of oxygen, and by the action of a group of specific bacteria, into gaseous products "biogas" (CH_4, CO_2, H_2, H_2S, etc.) and digestate, which is a mixture of mineral products (N, P, K, Ca, etc.) and compounds difficult to degrade.

The controlled anaerobic digestion process is one of the most suitable for the reduction of greenhouse gas emissions; harness the energy potential of organic waste while providing great value fertilizer byproducts.

Anaerobic digestion can be applied, among others, to livestock waste, agricultural waste and waste from the processing industries. They can take the form of slurry, manure, agricultural residues or surplus crops, etc. The anaerobic digestion process is also suitable for treating wastewater with high organic load, such as the one produced in many food industries.

2.2. Process

Anaerobic digestion is characterized by the existence of several distinct consecutive stages in the process of degradation of the substrate (generic to describe, in general terms, the food of microorganisms), based upon 5 large populations of microorganisms, These populations are characterized by being composed of different growth rates and different sensitivities to each intermediate compound as an inhibitor (e.g H_2, acetic acid or ammonia produced acidogenesis can inhibit certain flora). This implies that each stage will present different reaction rates according to the composition of the substrate and the steady development of the overall process will require a balance to avoid the accumulation of intermediates inhibitors or accumulation of volatile fatty acids (VFA), which could produce a lowering in the pH. For pH stability, it is important to have bicarbonate-CO2 equilibrium. To enable symbiosis in some reactions the association between acetogenic and methanogenic bacteria is necessary, creating bacteria aggregates of these different populations. This implies that the set-up of the reactors are generally slow, requiring times that can be on the order of several hours to days. Thereby making it as a batch process by default.

As previously introduced, anaerobic digestion is a complex multistep process. It has 4 main steps (Fongsatitkul, Mavinic and Lo):

- **Hydrolysis**: Breaking down of complex sugars into sugars that are fermentable, or capable to being used by the bacteria. This is usually an enzymatic process that utilizes invertase at high temperature. It can be catalyzed exponentially by the amount of Sulfuric Acid present. The acid helps in the cleavage of glucosidic linkages between the molecules of a complex sugar.
- **Acidogenesis:** In the second stage, acidogenic bacteria transform the products of the first reaction into short chain volatile acids, ketones, alcohols, hydrogen and carbon dioxide. The principal acidogenesis stage products are propionic acid, butyric acid, acetic acid, formic acid, lactic acid, ethanol and methanol, among other. From these products, the hydrogen, carbon dioxide and acetic acid will skip the third stage, acetogenesis, and be utilized directly by the methanogenic bacteria in the final stage
- **Acetogenesis:** In the third step, the rest of the acidogenesis products, i.e. the propionic acid, butyric acid and alcohols are transformed by acetogenic bacteria into hydrogen, carbon dioxide and acetic acid. Hydrogen plays an important intermediary role in this process, as the reaction will only occur if the hydrogen partial pressure is low enough to thermodynamically allow the conversion of all the acids. Such lowering of the partial pressure is carried out by hydrogen scavenging bacteria, thus the hydrogen concentration of a digester is an indicator of its health.

- **Methanogenesis:** This is the final stage and the crux of the whole process. In it, the available substrates are utilized by methanogens to produce methane. This is strictly an anaerobic process and won’t occur in the presence of oxygen. Therefore the vessel has to be purged with Nitrogen before the initiation of a batch.

In general, the process speed is limited by the speed of the slowest stage, which depends on the composition of each residue. For soluble substrates, the limiting stage is methanogenesis. To speed the reaction up, the strategy to employ has been to employ condition specific flora of acetogenic and methanogenic bacteria in separate reactors. With this system, one can achieve optimum processing times on the order of days. Feed where the organic material is in particulate form, the limiting step is the hydrolysis. This is so because hydrolysis is an enzymatic process whose speed depends on the particle surface. However, in our case the feed is in the form of a slurry that contains soluble particles therefore the hydrolysis steps takes only a few hours.

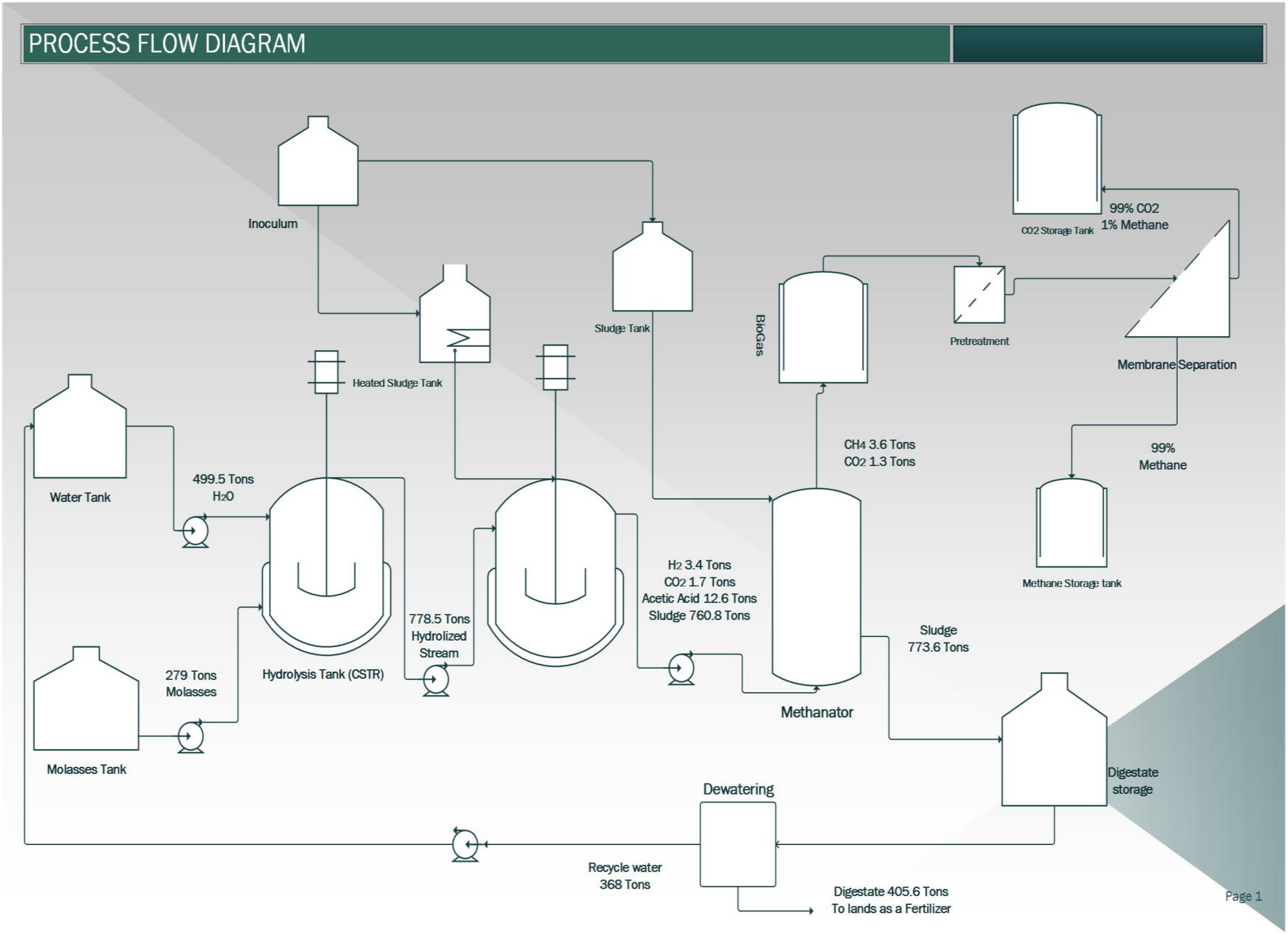
2.3. Parameters

The anaerobic process is very sensitive to environmental parameters. So, these need to be controlled in order for the process to be effective.

- pH, which must be maintained near neutral in a non-selective process. If the process is selective, as is the situation in our case, the pH needs to correspond with the active flora in the reactor.
- Alkalinity buffering capacity to ensure and prevent acidification. The alkalinity of the process is to be set to greater than 1.5 g / L CaCO3.
- Redox potential, with recommended values below -350 mV.
- Nutrients, with values that ensure the growth of microorganisms. Normally the manure can be used as a nutrient source for many types of microorganism. The sugar demand is fulfilled by molasses in our case.
- Toxic substances and bacterial inhibitors, like antibiotics, should have little to no concentration in each batch.

Apart from environmental parameters, several operational parameters also need to be maintained in optimal ranges for the process to be feasible:

- **Temperature**. This effects the type of bacterial flora active. Psychrophilic can be operated in the range of room temperatures, mesophilic at temperatures around 35 ° C and thermophilic at temperatures around 55 ° C. The variance in temperature is due to selective activation. For example, a mesophilic bacteria would form hard spores at high temperatures and remain dormant until favorable temperatures are attained.
- **Agitation.** Depending on the type of reactor system, energy must be provided to encourage the transfer of substrate to each aggregates\ of bacteria. Mixing also homogenizes the temperature and concentration profiles of the reactor and prevents undue accumulation.
- **Retention time**. It is the ratio between the volume and flow of treatment, ie, the average residence time of the influent in the reactor. It is subject to the action of microorganisms.
- **Organic loading rate**, OLR in abbreviation. This is the amount of organic matter introduced per unit volume and time. Low values result in very low effluent concentration and/or high retention time. Increased OLR means reduced gas production per unit of organic matter because of relatively high effluent organic matter concentrations. However, increased OLRs result in lower retention times. Therefore, it is advisable to find a technical/economic optimum value for each installation.


PROCESS FLOW DIAGRAM
Inoculum
Heated Sludge Tank
Sludge Tank
BioGas
Pretreatment
CO2 Storage Tank
99% CO2
1% Methane
Membrane Separation
99%
Methane
Methane Storage tank
Water Tank
499.5 Tons
H2O
Hydrolysis Tank (CSTR)
778.5 Tons
Hydrolized
Stream
279 Tons
Molasses
Molasses Tank
CH4 3.6 Tons
CO2 1.3 Tons
H2 3.4 Tons
CO2 1.7 Tons
Acetic Acid 12.6 Tons
Sludge 760.8 Tons
Methanator
Sludge
773.6 Tons
Digestate
storage
Dewatering
Recycle water
368 Tons
Digestate 405.6 Tons
To lands as a Fertilizer
Page 1

2.4. Process Flow

As depicted in the above diagram, the process can be divided into four main parts; inoculum, digestion recycle and purification.

The process kicks off in the digestion phase. There are two **tanks**; one for water and the other for molasses. The molasses is pumped through the use of a specialty **rotary lobe pump** due to its viscosity. Both the water and molasses are pumped into the first reactor, **hydrolyzer**, as slurry. Here the complex sugars are broken down into simpler fermentable sugars like, glucose and fructose. The hydrolization, occurs under acidic conditions at a temperature of 55^0C. (Fongsatitkul, Mavinic and Lo) The enzyme used is industrial grade *invertase*.

Before the following reactions, **inoculum segregation** takes place. The same base sludge is heated and cooled alternatively to provide with selective spore activation. Via this method, the bacterial flora we want to be flourishing in a certain reactor, does so.

Following the introduction of inoculum, in the second reactor, a combination of acetogenesis and acidogenesis takes place under *mesophillic* conditions. It is a mixed flow reactor and is signified as a **fermenter**. Here all the substrates required for methane production are formed. In addition, hydrogen is also formed that is partially consumed by hydrogen-scavenging bacteria while the remaining is sent to the next reactor as substrate.

In the **methanator,** the novelty *Upflow Anaerobic Sludge Blanket Reactor*, the heart of the process resides. Here substrates like CO_2, H_2, acetic acid and propanoic acid are converted to methane under *thermophillic* conditions. The UASB, as it is concisely known, acts more like a three-phase separator than a pure reactor. In it the effluent of the last reactor is introduced at the bottom of the tank. From here, it rises up through a cake of aggregated bacteria at the bottom, remains in the middle as liquid for a while before the evolving gas is captured by the shed up top. As the process of the methanator occurs at strictly anaerobic conditions, it needs to be purged with nitrogen prior to the start of every batch so as to ensure lack of oxygen.

The methanator branches into two further processes; recycle and purification. In recycle, a portion of the solids in the reactor are washed out into the **digestate storage tank**. Nominal washings occur every day but at the lapse of the Solid Retention Time (67 days), the whole of the prior sludge in the UASB is offloaded into the storage tank. Following this, the **decanter centrifuge** comes into play as it separates the heavier residue from the water content. The water, is recycled back to the water storage tank at the start. This is to ensure low capital cost in terms of water usage. This is especially considerable because of the water situation in many Southern parts of Pakistan where the supply falls well short of demand. The digestate, on the other hand, is separated out as solid residue that can be applied as a cheap, yet effective, fertilizer. (Fongsatitkul, Mavinic and Lo) The digestate, due to its BOD, has a fairly high content of nitrogenous elements that can prove beneficial for plants.

As stated before, the **methanator** yields to two processes. The second one of these is the purification one. Here, the biogas obtained from the methanator (which contains several gases apart from methane) is purified into high value methane. This is done via a series of steps. In the first step, represented by the *Pretreatment* black box, the H_2S and moisture content is removed. This is done so by a combination of **condensate traps** and **H_2S Scavengers**. Following the pretreatment, the gas is passed through an *ethyl acetate membrane* to give high purity CO_2 and CH_4. The **membrane** is designed such that up to 99% purity of the effluent gases is attained. At these purity levels the carbon dioxide can be sold to industries such as bottling plants while methane can be used as a high calorific value fuel for both burning and internal combustion engines. This methane can also be burnt to generate electricity. There is a precedent for this as well. Shakarganj Sugar Mill is generating 6 MW of on-site electricity using this method.

As a utility and HAZOP requirement, the gases are compressed and stored in tanks below the surface of the Earth. This is especially requisite in the case of methane, which is combustible.

2.5.Process Simulation

We had also modeled the aforementioned process in Aspen HYSYS. The specifications were the same as in the actual process. The simulation mechanics were different.

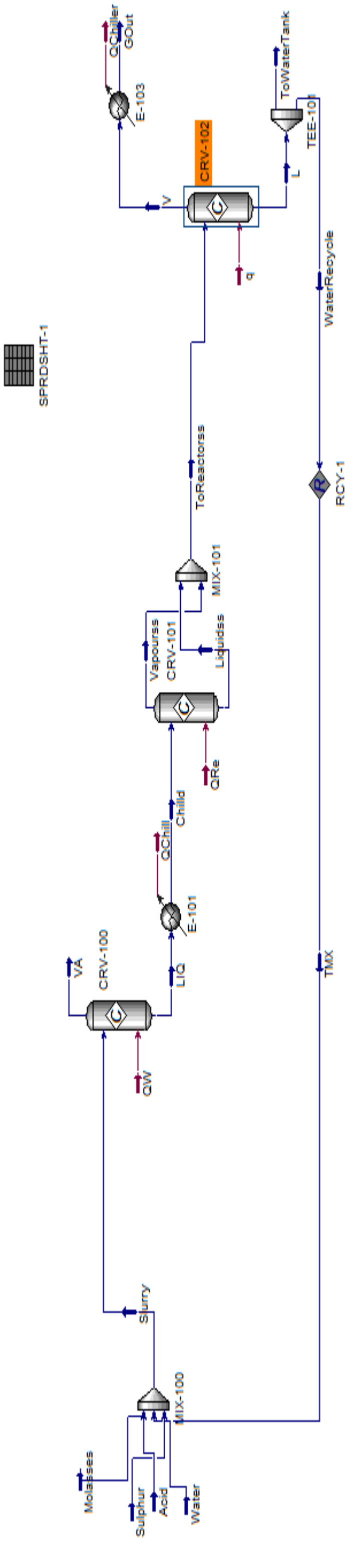
The fluid packaged used is *Non-random Two Liquid Model* (NRTL). There were some inorganic salt molecules that weren't there in the HYSYS component library. We constructed hypothetical molecules for these using properties available on online encyclopedias.

Another important point is the reaction set. To gain a rudimentary understanding of the process we had used *Conversion* models using conversion factor that are universally available in several research articles. As a result the reactors employed, were also conversion based as symbolized by a *C* on their vessels.

Additional unit-ops employed in the simulation are *Mixers* and *Tees*. The mixer is the flipped triangle at the start that takes in a variety of streams at varying conditions and *flashes* them into a combined, single outlet stream in a cumulative effect. *Tees*, conversely, split streams according to a predefined ratio. They are symbolized as triangles, similar to *Mixers*, but are flipped the other way in order to signify the flipping effects that take place.

The combinative effects of the streams can be seen in the exit stream compositions as simulated. We managed to converge the whole process with the requisite flashing parameters. However, operational limitations in the software have mitigated the efficacy of the simulation process. The batch process is inherently unstable during a simulation so we derived the process as a continuous one in order to calculate per unit hour conversions and products in order to upscale and use them to compare with the values obtained from manual calculations using our balances.

The entire converged process is depicted in the HYSYS flowsheet on the following page.

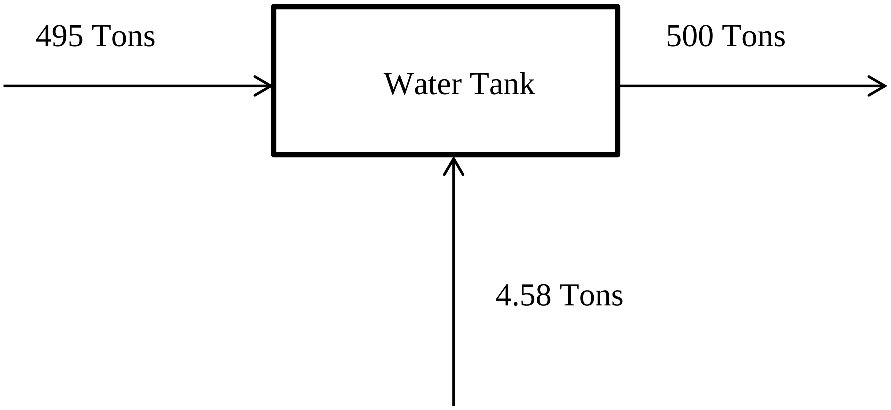

Molasses
Sulphur
Acid
Water
MIX-100
Slurry
QW
CRV-100
VA
LIQ
E-101
QChill
Chilld
QRe
CRV-101
Vapourss
Liquidss
MIX-101
ToReactorss
SPRDSHT-1
q
CRV-102
V
E-103
QChiller
GOut
L
TEE-10
ToWaterTank
WaterRecycle
RCY-1
TMX

3. Chapter 3

Material Balance

Basis: 1 day operation

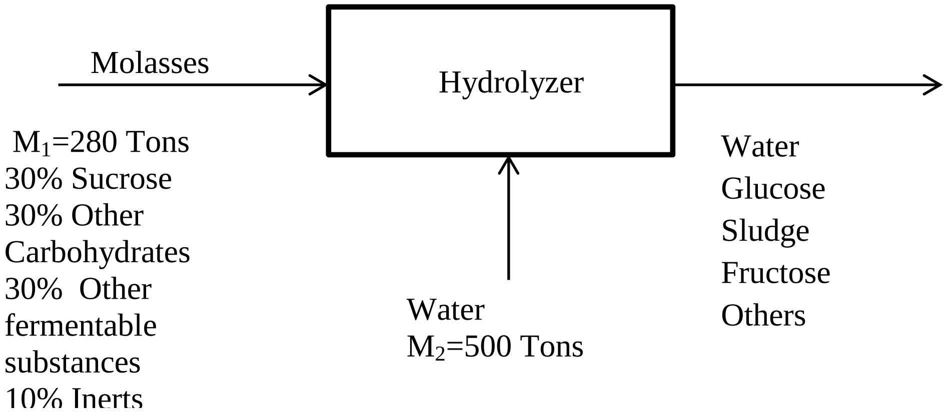
3.1 Water Tank



Outlet Stream to hydrolyzer = 495 Tons + 4.58 Tons (Recycled Stream)

Outlet Stream to hydrolyzer =500 Tons of water

3.2 Hydrolyzer



3.2.1. Component Balance

Feed M_1 Contains

30% Sucrose = 0.3*280 = 84 Tons

30% Other Carbohydrates = 0.3*280 = 84 Tons

30% Other fermentable substances = 0.3*280 = 84 Tons

10% inerts = 0.1*280 = 28 Tons

Formation of Glucose:

30% from Sucrose = 0.3*84 = 25.2 Tons

1 % from Other fermentable substances = 0.01*84 = 0.84 Tons

Total Glucose formed = 26.04 tons

Formation of Fructose:

30% from Other Carbohydrates = 0.3*84=25.2 Tons

Others:

29% from Other fermentable substances = 0.29*84= 24.36 tons

Sludge:

Hydrolyzed Stream (M_3) – Glucose – Fructoses – Others = Sludge

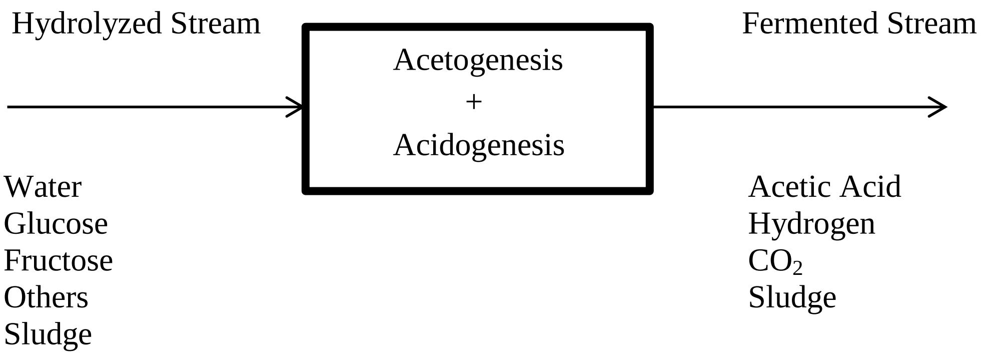
780 – 26.04 – 25.2 – 24.36 = 704.4 Tons

3.2.2. Overall Balance

$M_1 + M_2 = M_3$

280 + 500 = 780

3.3 Fermenter



3.3.1. Acetogenesis

Reaction or Conversion of Glucose (From 31%):

13 % Acid + Alcohol = 06.04*0.13= 3.3852 Tons

12 % Acetic acid = 26.04*0.12= 3.12478 Tons

6% Hydrogen and Carbon dioxide= 26.04* 0.06= 1.525 Tons

Conversion of Fructoses (From 30%):

16% Acid + Alcohol = 25.2 Tons*0.16= 4.032 Tons

2% Hydrogen and Carbon dioxide= 25.2*0.02= 0.504 Tons

Conversion of Others (From 29%):

20% Acetic Acid = 4.875 Tons

9% Hydrogen and Carbon dioxide= 2.19 Tons

Total Acid + Alcohol= 3.3852 + 4.032 = 7.4172 Tons

Total Acetic Acid = 11.02 Tons

Total Hydrogen and Carbon dioxide = 4.256 tons

3.3.2 Acidogenesis

Acid + Alcholo produce (From 29%):

20% Acetic Acid = 0.2*7.4172 = 1.48344 Tons

9 % Hydrogen and Carbon dioxide = 0.09*7.4172= 0.9918

Total Acetic Acid produced= 1.48344 + 11.02 = 12.5 Tons

Total Hydrogen and Carbon dioxide = 0.9918 + 4.256 = 5.2487 tons

1 part CO_2 and 2 part H_2

1.73:3.46

Sludge = 762.3 tons

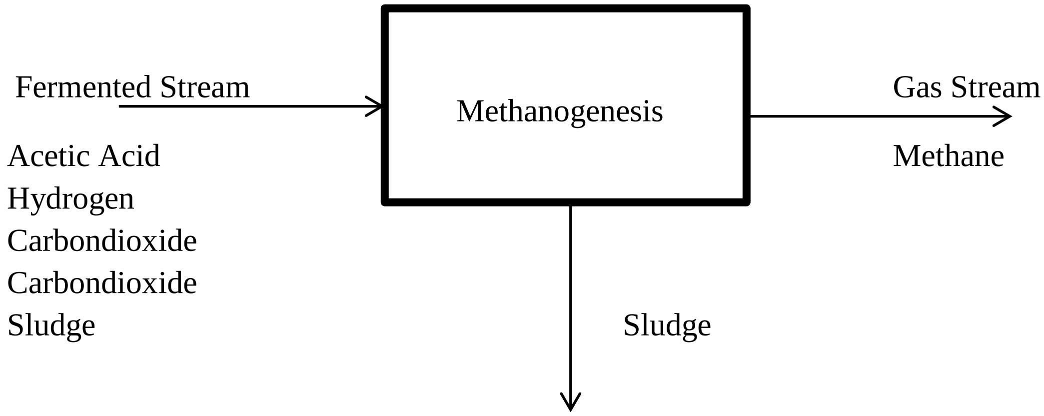
3.3.3 Overall Balance

Hydrolyzed Stream = Fermented stream

780=762.3+5.2+12.5

780=780

3.4 Methanator:



3.4.1. Reaction 1

$$CH_3COOH \xrightarrow{\qquad} CH_4 + CO_2$$

From Above Equation

Mass of Acetic Acid = 12.5 Tons

Moles of Acetic acid = 208.33 Kmol

Moles of CH_4 =104.165 kmol

Mass of CH_4 = 1.66 Tons

Moles of CO_2 = 104.165 Kmol

Mass of CO_2 = 4.58 Tons

Total CO_2=4.58 + 1.73 =6.31 Tons

3.4.2. Reaction 2

$$CO_2 + 4H_2 \xrightarrow{\qquad} CH_4 + 2H_2O$$

80% of CO_2 is converted into methane

80 % of 143.4 kmol is converted into methane= 114.7 kmol

Mass of CH_4 = 1.8352 Ton

114.7 kmol of CO_2 forms 2*114.7kmol of H_2O

Mass of H_2O= 4.129 Ton

Total Methane = 1.8352 + 1.66 Tons =3.5 Tons

Total Water remaining =4.129 Ton

CO_2= 1.262 Tons

Sludge = 771 tons

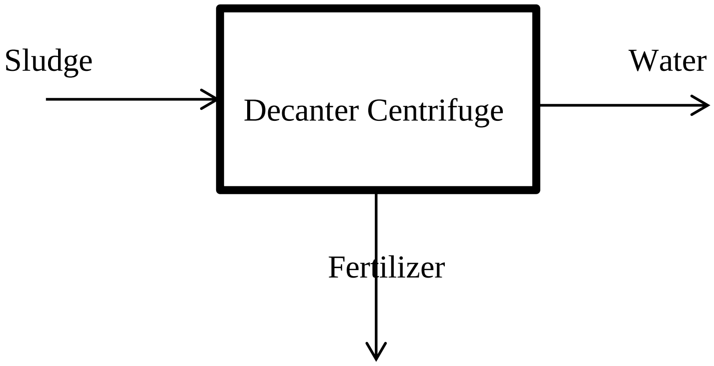
3.4.3. Overall Balance

Fermented Stream = Methanated Stream

780 = 3.5 + 4.129 +1.262 + 771

780 = 780

3.5 Centrifuge:



Sludge = Water + Fertilizer

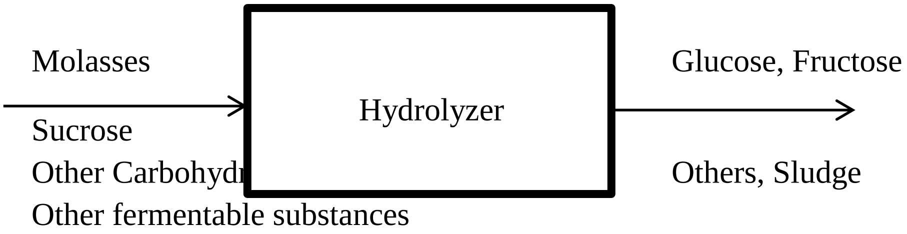
771 tons = 370 tons + 401 tons

4. Chapter 4

Energy Balance

Basis: 1 day Operation

4.1 Hydrolyzer:



4.1.1 Heat In

Heat in Sucrose = mCpΔT

= 84000*3.43*10

=2,881,200 KJ

Heat in Other Carbohydrates= mCpΔT

= 84000*3.07*10

=2,578,800 KJ

Heat in Other fermentable substances = mCpΔT

= 84000*2.52*10

=2,116,800 KJ

Heat in water = $mCp\Delta T$

= 5,00,000*4.18*10

=2,09,00,000 KJ

4.1.2. Heat Generated

Heat Generated by Sucrose = $mX_A\Delta H$

= 84,000*0.3*17,000

= $4.28*10^8$ KJ

Heat Generated by Other Carbohydrates = $mX_A\Delta H$

= 84,000*0.3*17,000

= $4.28*10^8$ KJ

Heat Generated by Other fermentable substances = $mX_A\Delta H$

= 84,000*0.3*17,000

= $4.28*10^8$ KJ

Total Heat Generated = $18.03*10^9$ KJ

4.1.3. Heat Out

Heat out by Glucose = $mCp\Delta T$

= 26040*0.115*10

=29946 KJ

Heat out by Fructose = $mCp\Delta T$

= 25200*0.178*10

=44856 KJ

Heat out by Others= $mCp\Delta T$

= 24360*2.621*10

=638232 KJ

Heat out by Sludge = $mCp\Delta T$

= 704400*2.92*10

=$2.057*10^7$ KJ

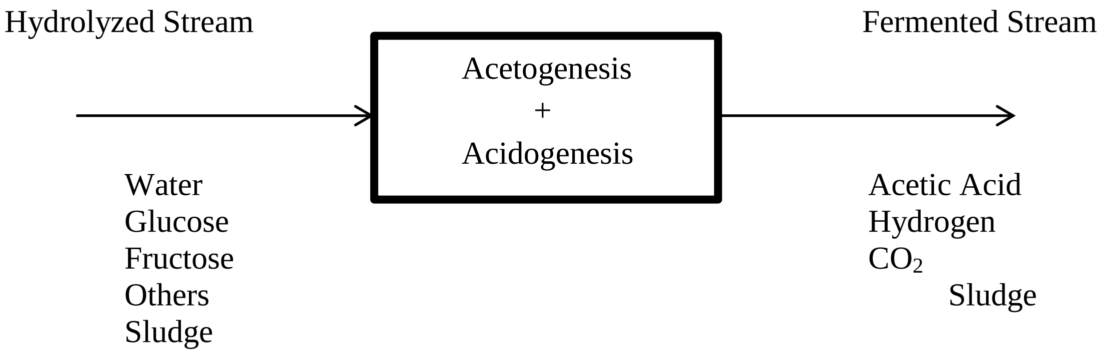
Total Heat Out= $2.12*10^7$ KJ

4.1.4. Overall Balance

Heat In = Heat Out

$2.847*10^7$ KJ = $2.12*10^7$ KJ

4.2. Fermenter



4.2.1. Heat In

Total heat in = $2.12*10^7$ KJ

4.2.2. Heat Generated

$$C_6H_{12}O_6 + 2H_2O \xrightarrow{\hspace{3cm}} CH_3COOH + 2CO + 2H_2O$$

Heat of formation of Glucose = -7.04 KJ/Kg

Heat of formation of Water = -15.87 KJ/Kg

Heat of formation of Acetic Acid = -8.08 KJ/kg

Heat of formation of Carbondioxide= -8943.6 KJ/Kg

ΔH_r = [2*(-8.08KJ/Kg)] + [2*(-8943.6 KJ/Kg] + [4*0] + [2*(-15.87 KJ/Kg)] – [-7.04 KJ/Kg]

= -17864.58 KJ/Kg

Heat generated = $mX_A\Delta H_r$

= 26040*0.9*-17864.58

= $-4.18*10^8$ KJ

4.2.3. Heat Out

Heat Out with Acetic Acid= $mCp\Delta T$

=11020*2043*10

=$2.25*10^5$ KJ

Heat Out with Hydrogen = $mCp\Delta T$

=3460*14.345*10

=$4.95*10^5$ KJ

Heat Out with CO_2= $mCp\Delta T$

$=1730*0.859*10$

$=14860.7$ KJ/Kg

Heat Out with Sludge= $mCp\Delta T$

$=762000*2.9*10$

$=2.21*10^7$ KJ

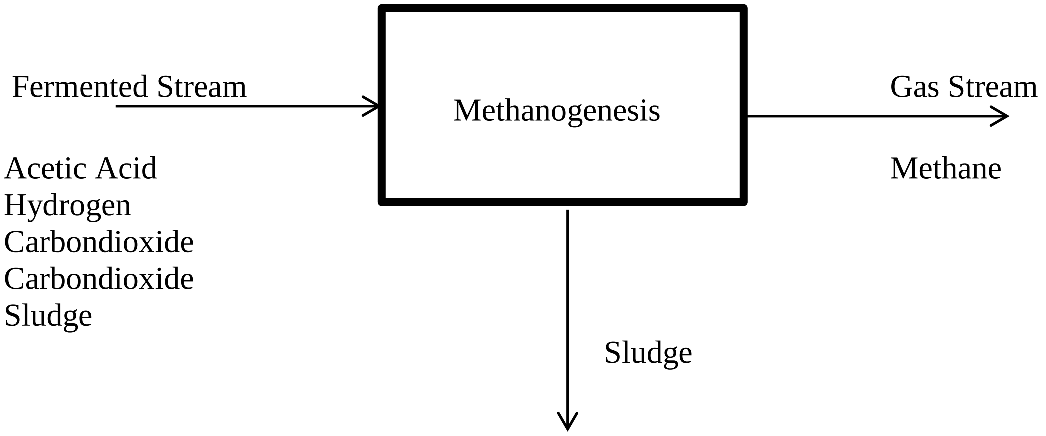
Total heat Out = $2.28*10^7$ KJ/Kg

4.2.4. Overall Balance

Heat in = heat Out

$2.12*10^7$ KJ $=2.28*10^7$ KJ

4.3. Methanator:



4.3.1. Heat In

Heat in = $2.28*10^7$ KJ/Kg

4.3.2. Heat Generated

$CH_3COOH \xrightarrow{\qquad\qquad} CH_4 + CO_2$

Heat of Reaction = (1*Heat of formation of Methane) + (1*Heat of formation of CO_2) – (1*Heat of formation of CH_3COOH)

ΔH_r = (-4.675 KJ/Kg) + (-8.943 KJ/Kg) – (-8.08 KJ/Kg)

= - 5.538 KJ/Kg

Heat Generated = $mX_A\Delta H$

=12500*1*-5.538

= - 69225 KJ/Kg

$CO_2 + 4H_2 \xrightarrow{\qquad\qquad} CH_4 + 2H_2O$

ΔH_r = [-4.675 KJ/Kg] + [2*(-15.78 KJ/Kg)] – [-8.943 KJ/Kg]

= -27.472 KJ/Kg

Heat Generated = $mX_A\Delta H$

=0.8*6310*-27.472

=-138678.6 KJ/Kg

Total Heat Generated = - 207,903.656 KJ/ Kg

4.3.3. Heat Out

Heat out with CH_4 = $mCp\Delta T$

=(3500)(2.62)(10)

=79100 KJ

Heat out with CO_2 = $mCp\Delta T$

$=1262*0.856*10$

$=10802$ KJ

Heat out with Sludge = $mCp\Delta T$

$=762300*2.9*10$

$=22106700$ KJ

Total Heat Out= $2.2*10^7$ KJ

4.3.4. Overall balance

Heat In = Heat Out

$2.28*10^7$ KJ = $2.19*10^7$ KJ

5. Chapter 5

Equipment Designing

5.1.1. Hydrolyzer

Design equation

$C_{12}H_{22}O_{11} + 2H_2O \xrightarrow{\qquad\qquad} 2C_6H_{12}O_6$

Rate Equation

$R_s = -k_bS$……………………………………………………………….. (1)

$dN_A/dt = r_AV$

$N_A = C_AV$

N_A in equation (1)

$VdC_A/dt = r_AV$

$dC_A/dt = r_A$ ……………………………………………………….…(2)

Putting r_A in eq (2)

$dC_A/dt = -k_bC_A$

$N_{Ao} = C_AV$

Moles of water = 5,00,000Kg/18= 27,777kgmol

Moles of molasses = 27,777kgmol

Mass of water = 500 Tons

Mass of Molasses = 280 Tons

Density of Water= 1000 kg/m^3

Specific Gravity of Molasses is 1.4

So,

Density of molasses is 1400 kg/m^3

Volume occupied by water = 500,000/1000= 500 m^3

Volume occupied by molasses = 280,000/1400 = 200 m^3

Total volume occupied by fluid = V_f = 700 m^3

C_{Ao}=N_{AO}/ V = 27777/700 = 39.68 kgmol/m^3

C_A=C_{AO}(1-X_A)

C_A=39.8(1-0.65)=13.885 kgmol/m^3

Now,

$$\int_{Cao}^{Ca} \frac{dCa}{Ca} = -kb \int_0^t dt$$

ln C_A/C_{Ao}= -k_bt

ln (13.8/39.8)= - k_bt

t=3.3 hrs

H/D ratio for the Hydrolizer tank

D=H/2

Volume of cylindrical vessel

V=π/4*H^2/2*H

For Volume of Vessel

20% free space is Allowed

So,

Total Volume of Hydrolyzer= 700*1.2= 840m^3

For Height H

$V=\pi/4*H^2/2*H$

H^3=4*840*4/3.14 = 16.19 m

Diameter is given as

D = H/2 = 16.19/2 = 8.1 m

5.1.2. Mechanical Desgin

Pressure due to Fluid

Density of fluid = (1400 + 1000)/2 = 1200 Kg/m^3

Height of fluid = $4V/\pi D^2$

=4*700/3.14*$(16.19)^2$

=13.6 m

Pressure due to fluid =$\int gh$

=1200*9.8*13.6

=159,936Pa = 159.936 KPa

Maximum Allowable Pressure = Atmospheric Pressure + Pressure due to fluid head

= 101.3 KPa + 159.936 KPa = 261.236 KPa

Design pressure

10% of maximum allowable pressure = 261.236*1.1=287.35KPa

5.1.3. Wall Thickness for Cylindrical Shell:

$e = P_i D / 2 J f - P_i$ (Coulson and Sinnott)

P_i= Internal pressure of the vessel

D= Diameter of Vessel

J= welding efficiency

f= maximum allowable stress = 165 N/mm^2

$e = P_i D / 2 J_f - P_i$

$e = (2.8735*10^5)(8.1) / [(2)(1)(165*10^6)-(2.875*10^5)]$

$e = 0.00705$ m

$e = 7.05$ mm

Corrosion allowance = 0.3mm/year

For 15 years (plant life)

Corrosion allowance = 0.3*15 = 4.5 mm

$e = 4.5 + 7.05 =$ *11.55 mm*

5.1.4. Thickness for Spherical head:

Tank Head; (Coulson and Sinnott)

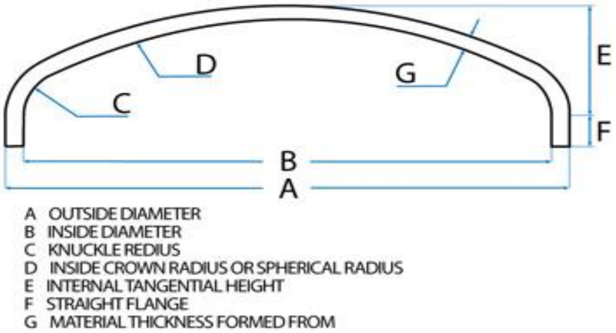


Figure 1-Torisperical Head

Torispherical head will be used because P<15 bar

$$e = \frac{P_i * R_c * C_s}{2*f*j*-P_i(C_s-0.2)}$$

R_c= Crown Radius= D_i=8.1m

R_i=Knuckle radius

R_c/R_i=0.06

Ri=0.06*8.1= 0.486m

C_S is stress concentration factor for torispherical head

$$C_s = \frac{1\left(3+\sqrt{R_c/R_i}\right)}{4}$$

C_s=1.77

e = $(2.873*10^5)(8.1)(1.77) / [(2) (1) (165*10^6) + (2.835*10^5)(1.77-0.2)]$

e = 0.012 m = 12.4 mm

Corrosion allowance = 0.3 mm/year

For 15 years

Corrosion allowance = 0.3*15 = 4.5 mm

e= 4.5 + 12.4 = 16.9 mm

5.1.5. Impeller Design

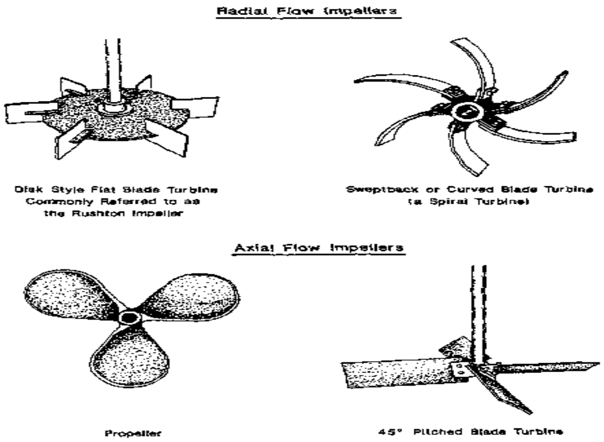


Figure 2-Impeller Designs

(Coulson and Sinnott)

Number of Baffles = 4

Number of Blades = 6

Shape factors

S_1=Da/D_t=1/6

S_2=E/D_t=1/3

S_3= L/D_a=1/4

S_4=W/D_a=1/5

S_6=J/D=1/2

S_6= H/D_t=2

Impeller Diamter

D_a=D_t*1/6

=8.1*1/6=1.35m

Impeller height above vessel

E=D_t*1/3 = 8.1/3= 2.7m

Impeller length

L= D_a/4=1.35/4=0.3375m

Width of impeller

W= Da/5=1.35/5=0.27m

Width of bladdes

$J= D_t/12=8.1/12=0.67m$

Height of Liquid In the Tank H= 13.6m

For Non-Newtonian fluids

$$N_{R=\frac{n^{2-n'}-D_a^2}{k^{n'}-R}}\rho$$

k=0.069

n’=0.664

No. of impeller revolutions = 0.83 r/s

By solving equation

$N_R= 5.2*10^4$

Power $=P= N_p n^3 Da^5 \int$

$=(1.63)(0.82)^3(1.35)^5(1200)$

= 5015 W

5.1.6. Jacket Design

Spiral Baffle jacket gives low pressure drop and it is less expensive. (Albright)

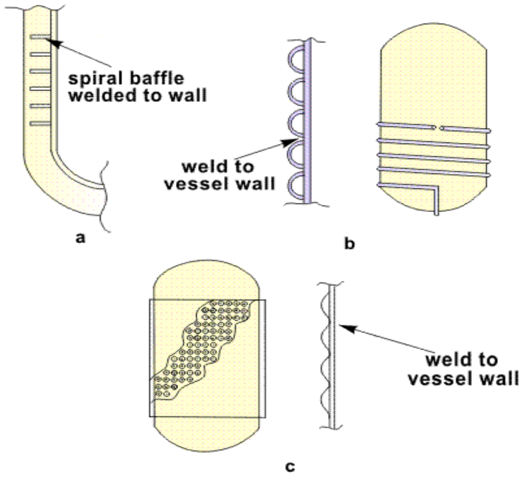


Figure 3-Jacket Designs

Height of jacket is 66% of the height of reactor.

=16.19*0.66=10.68m

Spacing Between jacket and reactor wall = 0.3m

Pitch between spiral= 0.6 m

Number of spirals= Height of jacket/pitch

=10.68/0.6=17.8

Length of channel = 2πr = 2πd/2=n*πd

=17.8*3.14*8.1=452.7m

Cross Sectional Area of channel = Pitch* Spacing

$=0.6*0.3=0.18m^2$

Hydraulic mean diameter=de=4*Cross-sectional Area / wetted perimeter

=4*0.18 / 2(0.3+0.6)

=0.4m

Water Inlet Temperature=25^oC

Water outlet temperature=60^oC

$Q=mCp\Delta T$

$Q=1.803*10^9$KJ/day$=2.08*10^7$J/s

Cp=4.18KJ/kg ^oC

FlowRate of Water = m = Q / $Cp\Delta T$

$=2.08*10^7/ 4.18*10^3*35$

=331.73 Kg/s = $0.331m^3/s$

At mean temperature:-

Density of Water = 1000 kg/m^3

Viscosity of water=$7.69*10^{-4}$ Pas

Thermal Conductivity of water = k =0.620W/m^oC

$Pr = Cp\mu/K$

$=4.18*7.96*10^{-4}/0.620=5.18*10^{-3}$

Velocity through channel =u= mass flow rate / density * area

=1.84m/s

$Re=de*u*\int/\mu$

=957087

For Water (Shell Side)

$Nu=CRe^{0.8}Pr^{0.33}$

$=(0.023)(9.5*10^{6})^{0.8}(5.18*10^{-3})^{0.33}$

=245

$Nu=h_{j}*de/k$

$h_{j=}Nu*k/de= 245*0.62/0.4$

$h_{j=}379.75\ W/m^{2\ o}\ C$

By using graph from book plant Desgin engineering by colson

$J_{r}=1.5*10^{-3}$

$\Delta P=8*Jr*L*\int*u^{2}/2de$

$=8*1.5*10^{-3}*452.7*1000*(1.84)^{2}/2*0.4$

$=2.2*10^{4}Pa$

=3.19 Psi

For Reactor Side:-

Agitated Speed=0.83r/s

Cp value of mixture=3.049 Kj/kg ^{o}C

K value of mixture=0.208 W/m$^{2\ o}$C

μ value of mixture = 0.02pas

$Pr=Cp*\mu/k = 0.293$

$Re = \int * u * de / \mu = 1.49*10^5$

For flat Blade

$Nu = 0.74\ Re^{0.67} Pr^{0.33}$

$= 0.74(1.49*10^5)^{0.67}(0.293)^{0.33}$

$= 1443.2$

$h_i * d/k = Nu$

$h_i = 222\ W/m^{2\ o}C$

$$\frac{1}{U} = \frac{1}{h_j} + \frac{e}{k} + \frac{1}{h_r}$$

$$\frac{1}{U} = \frac{1}{379.75} + \frac{0.115}{16} + \frac{1}{227}$$

$U = 128.9\ W/m^{2\ O}C$

$A = Q/U*\Delta T = 2.08*10^7/\ 128.9*15$

$A = 1.075*10^4\ m^2$

5.2. Fermenter

Design equation

$$C_6H_{12}O_6 + 2H_2O \xrightarrow{\hspace{2cm}} 2CH_3COOH + 2CO_2 + 4H_2$$

Volume Calculations:

mass of glucose = 26040 Kg

Density of glucose is 1500 kg/m^3

Volume occupied by Glucose = 26040/1500= 17.36 m^3

Mass of Fructoses = 25200kg

Density of Fructoses = 1356 kg/m^3

Volume of Fructose = 18.58 m^3

Mass of Other fermentable substances = 24360kg

Density of Other fermentable substances = 877.2 kg/m^3

Volume of Other fermentable substances = 27.66 m^3

Mass of water= 103500kg

Density of water = 1000 kg/m^3

Volume of water = 103.8 m^3

Uncovered material = 600600 kg

Density = 1100 kg/m^3

Volume = 546 m^3

Total Volume = 713.4 m^3

Key Component Glucose is represented by subscript A

N_{AO} =26040kg/180=144.7 Kmol

C_{AO}= 144.67/713.4 =0.202 kmol/m^3

C_A= $(1-X_A)C_{Ao}$

X_A=0.9

C_A= (1-0.9)*0.202 =0.0202 Kgmol/m^3

For Acetogenic bacteria

$\mu_{max}=0.33h^{-1}=7.92d^{-1}$

ks=1.7 kg/m^3

By putting all values in the rate equation

$$\int_{Cao}^{Ca} \frac{dCa}{Ca} = 0.174 \frac{7.29Ca}{1.7 + Ca} \int_{0}^{t} dt$$

t=16 hrs

H/D ratio for the Fermenter tank

D=H/2

Volume of cylindrical vessel

V=π/4*H^2/2*H

For Volume of Vessel

20% free space is Allowed

So,

Total Volume of Hydrolyzer= 713*1.2= 856m^3

For Height H

V=π/4*H^2/2*H

H^3=4*856*4/3.14 = 16.29 m

Diameter is given as

D = H/2 = 16.29/2 = 8.14 m

5.2.1. Mechanical Desgin

Pressure due to Fluid

Density of fluid = (1400 + 1000)/2 = 1200 Kg/m^3

Height of fluid = $4V/\pi D^2$

$=4*713/3.14*(16.29)^2$

=14.4 m

Pressure due to fluid =∫gh

=1200*9.8*14.4

=169515.8 = 169.51 KPa

Maximum Allowable Pressure = Atmospheric Pressure + Pressure due to fluid head

= 101.3 KPa + 169.51 KPa = 270.81 KPa

Design pressure

10% of maximum allowable pressure = 270.81*1.1=297.891KPa

5.2.2. Wall Thickness for Cylindrical Shell:

$e = P_i D / 2 J f - P_i$

P_i= Internal pressure of the vessel

D= Diameter of Vessel

J= welding efficiency

f= maximum allowable stress = 165 N/mm^2

$e = P_i D / 2 J_f - P_i$

$e = (2.9735*10^5)(8.14) / [(2)(1)(165*10^6)-(2.975*10^5)]$

e = 0.00735 m

e = 7.35 mm

Corrosion allowance = 0.3mm/year

For 15 years (plant life)

Corrosion allowance = 0.3*15 = 4.5 mm

e = 4.5 + 7.35 = *11.85 mm*

5.2.3. Thickness for Spherical head:

Tank Head;

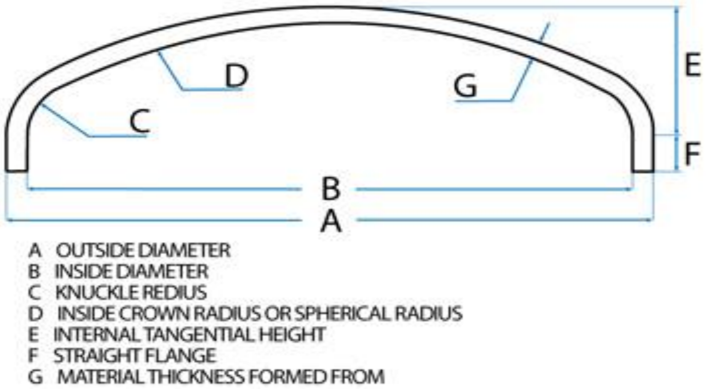


Figure 4-Torispherical Head Fermenter

Torispherical head will be used because P<15 bar

$$e = \frac{P_i*R_c*C_s}{2*f*j*-P_i(C_s-0.2)}$$

R_c= Crown Radius= D_i=8.14m

R_i=Knuckle radius

R_c/R_i=0.06

Ri=0.06*8.14= 0.4884m

C_S is stress concentration factor for torispherical head

$$C_s = \frac{1\left(3 + \sqrt{R_c/R_i}\right)}{4}$$

C_s=1.77

$e = (2.973*10^5)(8.14)(1.77) / [(2)\ (1)\ (165*10^6) + (2.97*10^5)(1.77-0.2)]$

e = 0.0129 m = 12.9 mm

Corrosion allowance = 0.3 mm/year

For 15 years

Corrosion allowance = 0.3*15 = 4.5 mm

e= 4.5 + 12.9 = 17.4 mm

5.2.4. Impeller Design

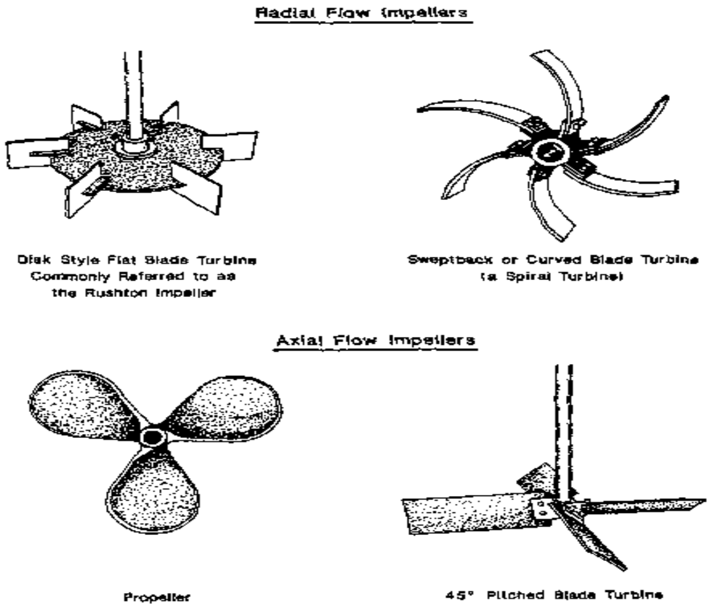


Figure 5- Flate Blade Impeller

(Coulson and Sinnott)

Number of Baffles = 4

Number of Blades = 6

Shape factors

S_1=Da/D_t=1/6

S_2=E/D_t=1/3

S_3= L/D_a=1/4

S_4=W/D_a=1/5

S_6=J/D=1/2

S_6= H/D_t=2

Impeller Diamter

$D_a=D_t*1/6$

$=8.14*1/6=1.356$m

Impeller height above vessel

$E=D_t*1/3$ = 8.14/3= 2.713m

Impeller length

L= $D_a/4$=1.356/4=0.339m

Width of impeller

W= Da/5=1.356/5=0.2712m

Width of bladdes

J= $D_t/12$=8.14/12=0.678m

For Non-Newtonian fluids

$$N_{R=\frac{n^{2-n'}-D_a^2}{k^{n'}-R}}\rho$$

k=0.069

n’=0.664

No. of impeller revolutions = 0.83 r/s

By solving equation

$N_R= 5.2*10^4$

$N_p=K_t$ when $N_{re}>1000$

For turbulent Flow

$N_p=K_t=1.63$

Power =P= $N_p n^3 Da^5 \int$

$=(1.63)(0.82)^3(1.356)^5(1200)$

= 5127.4 W

5.2.5. Jacket Design

Spiral Baffle jacket gives low pressure drop and it is less expensive. (Coulson and Sinnott)

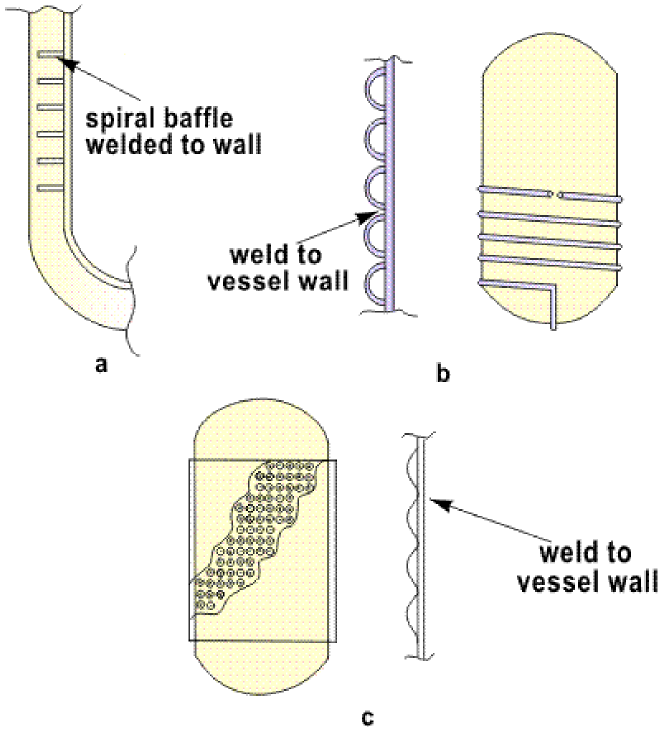


Figure 6- Baffle Jacket

Height of jacket is 66% of the height of reactor.

$=16.29*0.66=10.75m$

Spacing Between jacket and reactor wall = 0.3m

Pitch between spiral= 0.6 m

Number of spirals= Height of jacket/pitch

$=10.75/0.6=17.91$

Length of channel = $2\pi r = 2\pi d/2=n*\pi d$

$=17.91*3.14*8.14=457m$

Cross Sectional Area of channel = Pitch* Spacing

$=0.6*0.3=0.18m^2$

Hydraulic mean diameter=de=4*Cross-sectional Area / wetted perimeter

=4*0.18 / 2(0.3+0.6)

=0.4m

Water Inlet Temperature=25^{o}C

Water outlet temperature=40^{o}C

Q=mCpΔT

Q=4.18*10^{8}KJ/day

Cp=4.18KJ/kg ^{o}C

FlowRate of Water = m = Q / CpΔT

=4.18*10^{8}/ 4.18*10^{3}*15

=77.16 Kg/s = 0.07m^{3}/s

At mean temperature:-

Density of Water = 1000 kg/m^{3}

Viscosity of water=7.69*10^{-4} Pas

Thermal Conductivity of water = k =0.620W/m^{o}C

Pr = Cpµ/K

=4.18*7.96*10^{-4}/0.620=5.18*10^{-3}

Velocity through channel =u= mass flow rate / density * area

=0.42 m/s

Re=de*u*∫/µ

=218,465

For Water (Shell Side)

Nu=CRe$^{0.8}$Pr$^{0.33}$

=(0.023)(218,465)$^{0.8}$(5.18*10^{-3})$^{0.33}$

=75.5

Nu=h_j*de/k

$h_{j=}$Nu*k/de= 75.5*0.62/0.4

$h_{j=}$117 W/m^2 ^{o} C

For Reactor Side:-

Agitated Speed=0.83r/s

Cp value of mixture=3.049 Kj/kg ^{o}C

K value of mixture=0.208 W/m^2 ^{o}C

µ value of mixture = 0.02pas

Pr=Cp*µ/k = 0.49

Re=∫*u*de/µ = 4*10^4

For flat Blade

$$Nu=0.74\ Re^{0.67}Pr^{0.33}$$

$$=0.74(4*10^4)^{0.67}(0.49)^{0.33}$$

$$=708.5$$

h_i*d/k =Nu

$h_{i=}$108.67 W/m^2 ^{o} C

Thermal Conductivity of Stainless Steel=k=16 W/m^2 ^{O}C

$$\frac{1}{U}=\frac{1}{h_j}+\frac{e}{k}+\frac{1}{h_r}$$

$$\frac{1}{U}=\frac{1}{117}+\frac{0.011}{16}+\frac{1}{108}$$

U=92.15 W/m^2 ^{O}C

A=Q/U*ΔT =

A=3500 m^2

5.3. Methanogenic Reactor Design

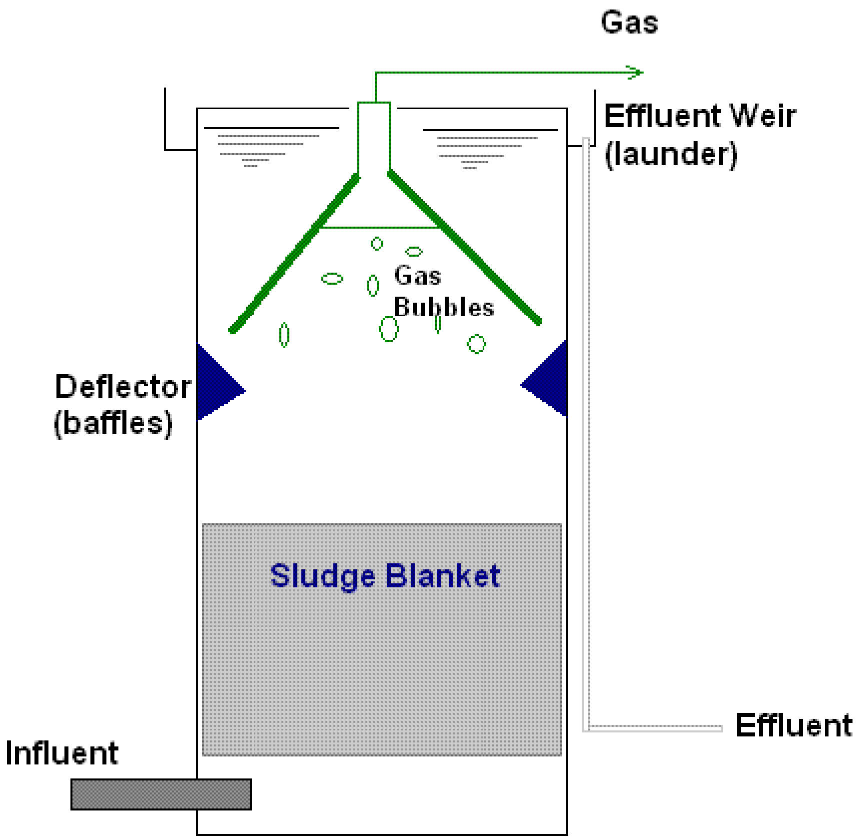


Figure 7- UASB Reactor

UASB Reactor design (Fongsatitkul, Mavinic and Lo).

5.3.1. For reactor volume

$$V_n = \frac{QS_o}{L_{org}}$$

Vn = (766 m3/d)(13.82 kg sCOD/m3) / (10 kg sCOD/m3/d)

$Vn = 1058 \text{ m}^3$

For effective reactor volume

$$V_L = \frac{V_n}{E}$$

$V_L = (1058 / 0.85)$

$V_L = 1244 \text{ m}^3$

5.3.2. For reactor diameter

$$A = \frac{Q}{v}$$

$A = (766 \text{ m}^3/\text{d}) / (1.0 \text{ m/h})(24 \text{ h/d})$

A = 31.87 m2

$$A = \frac{\pi D^2}{4}$$

D = 6.37 m

5.3.3. For reactor height

$$H_L = \frac{V_L}{A}$$

$H_L = 1244 / 31.87$

$H_L = 34.67 \text{ m}$

$H_T = H_L + H_G$

$H_T = 34.67 + 2.0$

$H_T = 36.67 \text{ m}$

5.3.4. For HRT

$$\tau = \frac{V_L}{QC_f}$$

= (1244 m^3)(24 h/d) / (766 m^3/d)(1.1)

= 33 hours

5.3.5. For reactor solid retention time

The value of the SRT (the number of days the sludge should reside in the reactor) *can be estimated by assuming that all the wasted biological solids are in the effluent flow. A conservative design approach is to assume that the given effluent VSS concentration consists of biomass.*

The effluent soluble COD concentration at 98% COD removal is:

S = (1-0.98)(13820 g/ m^3)

S = 276.4 g/ m^3

The effluent nbVSS concentration given that 50% of the influent VSS is degraded is:

nbVSS = 0.50(150 g/ m^3)

nbVSS = 75 g/ m^3

The pCOD degraded is:

pCOD degraded = 0.50(15400-13820) g/m^3

pCOD degraded = 790 g/m^3

The total degraded influent COD, So:

So = (13820 + 790)

So = 14610 g/m3

We know two equations

$$QX_e = P_{x,VSS} = Solids\ wasted/day$$

$$P_{x,VSS} = \frac{Q(Y)(So - S)}{1 + (k_d)SRT} + \frac{f_d(k_d)Q(Y)(So - S)SRT}{1 + (k_d)SRT}$$

Equating these two and substituting the data we have:

QXe = (766 m3/d)(790 g/m3) = (766 m3/d)(0.08 g VSS/g COD)[(14610-276.4)g/m3] / [1+0.03 g VSS/g VSS.d)SRT] + (0.15g VSS/ g VSS)(0.03 g VSS/g VSS.d)(766 m3/d)(0.08 g VSS/g COD))[(14610-276.4)g/m3]SRT / [1+0.03 g VSS/g VSS.d)SRT] + (766 m3/d) (75 g/m3)

605,140 g/d = 878,363 / [(1.03)SRT] + 3952 / [(1.03)SRT] + 57450 g/d

SRT = 67 days

The effluent soluble COD at SRT of 67 days at 55 ^{0}C

We know that,

$$k = \frac{\mu_{max}}{Y}$$

k = (0.25 g VSS/ g VSS.d) / (0.08 g VSS/g COD)

k = 3.125 g COD/g VSS.d

Now,

$$S = \frac{Ks[1 + (k_d)SRT]}{SRT\ (Y_k - k_{d)} - 1}$$

S = 7585 mg/L (1 + (0.03 g/g.d)] / [(67 d[(0.08 g/g)(3.125 g/g.d)-(0.03 g/g.d)]-1]

S = 497 mg/L

Now checking the adequacy of the given SRT value by comparing the fraction of influent sCOD in effluent = (497 mg/L / 13820 mg/L) = 0.036 = 3.6%S

This value is considerably less than the critical value of 10%, so our calculated SRT holds.

5.3.6. The average sludge blanket concentration,

X_{RSS} is computed using:

$$SRT = \frac{V(X_{RSS})}{(Q - Q_w)X_e + Q_w X_R}$$

Because it was assumed that all wasted solids were in the effluent flow, therefore $Q_w = 0$ and,

$$X_{RSS} = \frac{(Q)X_e.SRT}{V}$$

X_{RSS} = (766 m^3/d)(630 g/m^3)(67 d)(1 kg/ 10^3g) / 1044 m^3

$X_{RSS =}$ 30.97 kg/m^3

5.3.7. The amount of Methane produced using these COD amounts

The total COD degraded:

COD_d = (14610 – 276.4) g/ m^3

COD_d = 14333 g/ m^3

The COD removed by Sulfate as the electron acceptor:

0.67 g COD removed / g SO_4 reduced

COD_{SR} = 0.90(1383 g SO_4/ m^3)(0.67)

COD_{SR} = 834 g/m^3

The COD removed by competing bacterial flora:

For the thriving conditions of methanobacterium beijingese the competition factor is 0.14 of the relevant COD

COD_{COMP} = 0.20(COD_d - COD_{SR})

COD_{COMP} = 0.14(14333-834) g/m^3

COD_{COMP} = 1895 g/m^3

Now the COD removed by methanogenic bacteria:

$$COD_{MB} = (14333 - 834 - 1895)\ g/m^3 \times 766\ m^3/d$$

$$COD_{MB} = 11{,}604{,}664\ g/d$$

Therefore, the rate of methane production is:

Methane Production at 55 $^0C = (0.40\ L/g)\ (273.15+35\ /\ 273.15+55) = 0.53245\ L/g$

Amount of CH_4 produced/d = 0.53245 L/g (11,604,664 g/d)

= 6,178,903 L/d

= 6178.9 m^3

5.4. Membrane

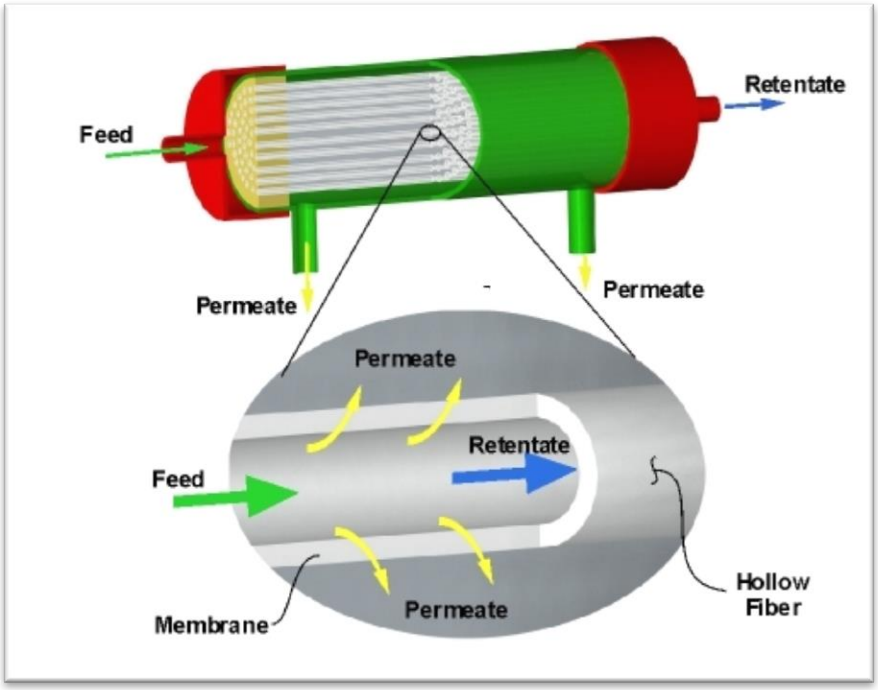


Figure 8- Membrane Separation

Driving equations for membrane designing (Geankoplis and Geankoplis)

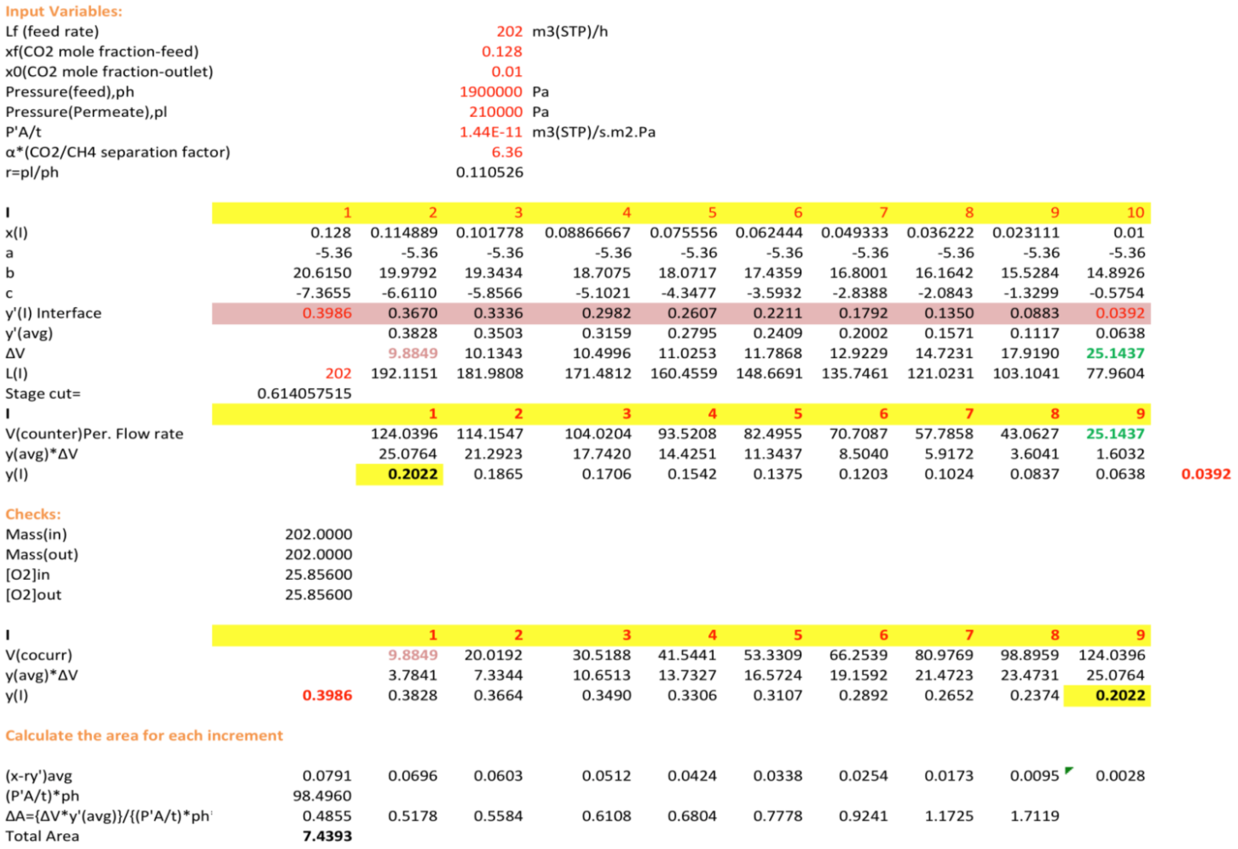
$\mathbf{J_A=(D_AS_A(P_{A1}-P_{A2}))/z}$

Formula used to calculate area:

$\mathbf{\Delta A=\{\Delta V*y'(avg)\}/\{(P'A/t)*ph*(x-ry')avg\}}$

Characteristic	**Details**
Module	Hollow fiber
Flow Regime	Counter current flow
Lf (feed rate)	$202m^3$(STP)/hr
Xf (CO_2 mole fraction-feed)	0.128
Xo (CO_2 mole fraction-outlet)	0.01
Pressure (feed) ph	1900kpa
Pressure (permeate) pl	210kpa
P’A/t	1,44 E-11m^3(STP)/s.m^2.pa
a* (selectivity CO_2 / CH_4)	6.36
R= pl/ph	0.11
Stage cut	0.61
Area	$7.44m^2$

Detailed calculations of membrane are as follows



Input Variables:

Lf (feed rate)	202	m3(STP)/h
xf(CO2 mole fraction-feed)	0.128	
x0(CO2 mole fraction-outlet)	0.01	
Pressure(feed),ph	1900000	Pa
Pressure(Permeate),pl	210000	Pa
P'A/t	1.44E-11	m3(STP)/s.m2.Pa
α*(CO2/CH4 separation factor)	6.36	
r=pl/ph	0.110526	

I	1	2	3	4	5	6	7	8	9	10	
x(I)	0.128	0.114889	0.101778	0.08866667	0.075556	0.062444	0.049333	0.036222	0.023111	0.01	
a	-5.36	-5.36	-5.36	-5.36	-5.36	-5.36	-5.36	-5.36	-5.36	-5.36	
b	20.6150	19.9792	19.3434	18.7075	18.0717	17.4359	16.8001	16.1642	15.5284	14.8926	
c	-7.3655	-6.6110	-5.8566	-5.1021	-4.3477	-3.5932	-2.8388	-2.0843	-1.3299	-0.5754	
y'(I) Interface	0.3986	0.3670	0.3336	0.2982	0.2607	0.2211	0.1792	0.1350	0.0883	0.0392	
y'(avg)		0.3828	0.3503	0.3159	0.2795	0.2409	0.2002	0.1571	0.1117	0.0638	
ΔV		9.8849	10.1343	10.4996	11.0253	11.7868	12.9229	14.7231	17.9190	25.1437	
L(I)	202	192.1151	181.9808	171.4812	160.4559	148.6691	135.7461	121.0231	103.1041	77.9604	
Stage cut=	0.614057515										
I		1	2	3	4	5	6	7	8	9	
V(counter)Per. Flow rate		124.0396	114.1547	104.0204	93.5208	82.4955	70.7087	57.7858	43.0627	25.1437	
y(avg)*ΔV		25.0764	21.2923	17.7420	14.4251	11.3437	8.5040	5.9172	3.6041	1.6032	
y(I)		**0.2022**	0.1865	0.1706	0.1542	0.1375	0.1203	0.1024	0.0837	0.0638	**0.0392**

Checks:

Mass(in)	202.0000
Mass(out)	202.0000
[O2]in	25.85600
[O2]out	25.85600

I		1	2	3	4	5	6	7	8	9
V(cocurr)		9.8849	20.0192	30.5188	41.5441	53.3309	66.2539	80.9769	98.8959	124.0396
y(avg)*ΔV		3.7841	7.3344	10.6513	13.7327	16.5724	19.1592	21.4723	23.4731	25.0764
y(I)	**0.3986**	0.3828	0.3664	0.3490	0.3306	0.3107	0.2892	0.2652	0.2374	**0.2022**

Calculate the area for each increment

(x-ry')avg	0.0791	0.0696	0.0603	0.0512	0.0424	0.0338	0.0254	0.0173	0.0095	0.0028
(P'A/t)*ph	98.4960									
ΔA={ΔV*y'(avg)}/{(P'A/t)*ph	0.4855	0.5178	0.5584	0.6108	0.6804	0.7778	0.9241	1.1725	1.7119	
Total Area	**7.4393**									

5.5. Centrifuge

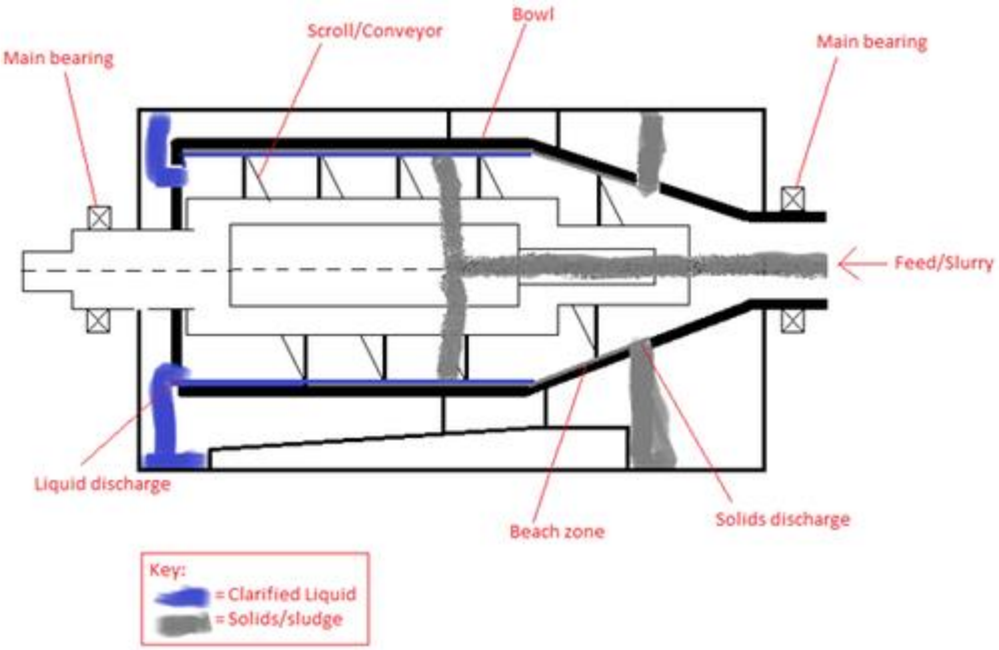


Figure 9- Decanter Centrifuge

For a decanter centrifuge for separeation of molasses and water the suitable diameter is 1.6m according to literature so radius will.

r= D/2
r= 1.6/2
r= 0.8m
From Richard and Coulson L/D can be between 1.5 to 5.
we are using L/D = 3

r_i = light liquid (water)= 0.085m

r_w = heavy liquid (molasses)= 0.137m
p_1= light liquid (water)= 1.0 g/cm^3

p_2= heavy liquid (molasses)= 1.4 g/cm^3

Interface diameter (r_s^2)= $((p_2)(r_w^2) - (p_1)(r_i^2))/ (p_2 - p_1)$

Interface diameter (r_s^2)= ((1.4)(0.0188) – (1)(0.00722))/ (1.4-1)

Interface diameter (r_s^2)= (0.02632 – 0.00722) / 0.4

Interface diameter (r_s^2)= 0.04751m

r_s= 0.218m

Now

L/D= 3

L= D*3

L= 0.8 * 3 = 2.4m

$V = pi\ (r^2 - r_i^2)\ (L)$

$V = 3.14\ (0.8^2 - 0.085^2)\ (2.4)$

V = 3.14 (0.64 – 0.00722) (2.4)

$V = 4.76\ m^3$

Now

$t_v = Q/ (pi\ (r^2 - r_i^2)\ (L)) \quad = Q/ V$

Q= V/day = (m/p)/day

$Q= ((773.6*1000)\ kg /(1200)\ m^3/kg\ )/day$

$Q= 644\ m^3/day$

$Q= 26.83\ m^3/hr$

$t_v= Q/V \quad = (26.83)/(4.76)$

$t_v= 5.64$ hrs

6. Chapter 6

Cost Analysis

6.1. Initial investment

Firstly we will calculate the purchased cost of major equipment's. (Coulson and Sinnott)

WATER TANK

Volume of reactor = 605m^3

Material factor= 1.0

Pressure factor= 1.1

Total cost of equipment= 2300*$(605)^{0.55}$

= $78,000/

MOLASSES TANK

Volume of reactor = 220m^3

Material factor= 1.0

Pressure factor= 1.2

Total cost of equipment= 2300*$(220)^{0.55}$

= $45,000/

MOLASSES BUFFER TANK

Volume of reactor = 715 m^3

Material factor= 1.0

Pressure factor= 1.2

Total cost of equipment= 2300*$(715)^{0.55}$

= $86,000/

SLUDGE TANK

No of equipment= 2

Volume of reactor = 14m^3

Material factor= 1.0

Pressure factor= 1.1

Total cost of equipment= (2300*$(14)^{0.55}$)*2

= $20,000/

SEED TANK

Volume of reactor = 5m^3

Material factor= 1.0

Pressure factor= 1.1

Total cost of equipment= 2300*$(5)^{0.55}$

= $56,000/

HYDROLYZER REACTOR

Volume of reactor = 840m^3

Material factor= 1.5

Pressure factor= 1.1

Total cost of equipment= (15000*$(5)^{0.55}$)*1.5

= $326,000/

FERMENTER

Volume of reactor = 856m^3

Material factor= 1.5

Pressure factor= 1.1

Total cost of equipment= $(15000*(856)^{0.55})*1.5$

= $348,000/

METHANATOR

Volume of reactor = 970m^3

Material factor= 1.5

Pressure factor= 1.1

Total cost of equipment= $(12000*(5)^{0.55})*1.5$

= $280,000/

CENTRIFUGE

Total cost of equipment=$93,000/

MEMBRANE

Total cost of equipment=$74,000/

Total purchase cost of major equipment items (PCE)	
Water tank	$78,000/
Molasses tank	$45,000/
Molasses buffer tank	$86,000/
Sludge tanks	$20,000/
Seed Tank	$56,000/
Hydrolyzer	$326,000/
Fermenter	$348,000/
Methanator	$280,000/
Total	$1,355,600/

PCE $1,355,600

f1 Equipment erection	0.2
f2 Piping	0.45
f3 Instrumentation	0.15
f4 Electrical	0.1
f5 Buildings	not required
f6 Utilities	0.15
f7 Storages	0.25
f8 Site development	not required
f9 Ancillary buildings	not required

SO

Total physical plant cost (PPC) =1,355,600132,300*(1+0.2+0.45+0.15+0.1+0.15+0.25)

= **$3,250,000/**

f10 Design and Engineering	**0.28**
f11 Contractor's Fee	**0.02**
f12 Contingencies	**0.1**

Fixed capital = 3,250,000 * (1+0.28+0.02+0.1)

= $4,550,000/

Working capital, allow 5% of fixed capital to cover the cost of the initial charge

Working capital = 0.05 * 4,550,000

= $227,500/

Total investment required for project = 4,550,000 + 227,500

= **$4,777,500/**

6.2. Variable cost

Raw material cost	**$300,000/yr @$130/ton**
Miscellaneous material	**$50,000/yr**
Shipping and packing	**negligible**
Total Variable Cost	**$350,000/yr**

6.3. Fixed cost

Maintenance (take as 5% of fixed capital)	= 4,550,000* 0.05
	= $227,500/yr
Operating labor	= $250,000/yr
Supervision (no additional supervision would be needed)	
Plant overheads (take as 50% of operating labor)	= 250,000* 0.5
	= $125,000/yr
Laboratory (take as 30% of operating labor)	= 250,000* 0.3
	= $75,000/yr
Capital charges, 6% of fixed capital	= 4,550,000* 0.06
	= $273,000/yr
Insurance, 1% of fixed capital	= 4,550,000* 0.01
	= $45,500/yr

Total Fixed Cost = $1,021,000/yr

Direct Production Cost = Variable Cost + Fixed Cost

= 350,000 + 1,021,000

= **$1,371,000/yr**

6.4. Revenue generated

Sales of CH_4	**$1,386,000/yr @77 $/ton**
Recycles & Waste as fertilizer	**$608,000/yr**
Sales of CO_2	**$250,000**
Total Revenue Generated	**$ 2,244,000/yr**

6.5. Profit

Profit = Revenue Generated – Direct Production Cost

= 2,244,000 – 1,371,000

= **$873,000/yr**

6.6. Rate of return

Rate of return = Profit per year / Total initial investment

= (873,000/4,777,500)* 100

= **18.23**

6.7. Payback period

Pay Back Period = 100 / Rate of return

= 100 / 18.23

= 5 years and 4 months (approx.)

7. Chapter 7

Conclusion

Through this green process, we have managed to produce methane at 3.5 tons/d. Carbon Dioxide is also produced as a side product at 1.3 tons/d. However, focusing on the greener aspect of energy generation, we haven't released the carbon dioxide into the atmosphere but rather have stored it for commercial use.

The 99% pure methane produced using this process is cost-effective and competitive at generating electricity as well as a fuel for both domestic and automobile usage. The digestate of this process, on the other hand, can be used as field discharge for its fertilizing effect. Thus this can be an effective process to compensate our dwindling gas reserves while also being economically viable.

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