OPTIMIZATION OF THE PYROLYSIS PROCESS

Bio Oil Production

NATIONAL UNIVERSITY OF SCIENCES & TECHNOLOGY
PAKISTAN

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Certificate

This is to certify that work in this thesis has been carried out by **Mr. Qassam Sarmad, Mr. Muhammad Sarmad Khalid** and **Mr. Uneeb Masood Khan** and has been completed under the supervision of **Dr. Arshad Chughtai** in the School of Chemical and Materials Engineering (SCME), National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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This project and this thesis is dedicated firstly to our parents, without whom none of this would have been possible. We also thank SCME and all the faculty in it for supporting us throughout the entirety of the project, specially our HoD, Dr. Arshad Hussain.

Acknowledgements

Dr. Arshad Chughtai has been the ideal project supervisor and mentor, who inspired us and pushed us to leave behind a legacy here at NUST. We would also like to thank Mr. Cory Jensen, who helped us in choosing a suitable project. The novel idea for the **Rotating Cone Reactor** comes from the researches, particularly in the field of BioFuels from Agricultural Waste, from University of Twente. Without their help the process efficiency would have been very low and the costs too high.

Abstract

The fossil fuel reserves are exhaustible apart from the fact that they pose serious environmental and health issues like the emission of Green House Gases. Moreover, other sources of energy must be explored and exploited to keep up with the ever increasing demand. Pakistan being an agricultural country produces a lot of agricultural waste which can provide an opportunity to meet the energy requirements. Fast pyrolysis is an effective method to convert biomass to bio-oil because using small residence times and fast heating rates favors the production of liquid (bio-oil) to produce transportation fuels. There are however some problems with bio-oil, it is corrosive hence it damages the equipment, has high levels Oxygen and moisture which significantly reduces the calorific value of the fuel.

The objectives of this research will be to optimize this process so that the maximum yield of bio-oil may be produced, minimizing the by-products. Moreover, the bio-oil will be upgraded using suitable catalytic routes so that it can be effectively used as a substitute or can be blended with the already existing crude oil derivatives, this way no modifications will be required in the engines, turbines, etc. The biofuel can also be blended with the already existing crude oil derivatives, this way no modifications will be required in the engines, turbines, etc. The process should be such that it fulfills the required parameters for the production because if the physical parameters do not align with the requirements, undesirable by-product like char is produced. Also the aging of the pyrolysis liquid should be avoided as it separates into various tarry, sludgy and waxy aqueous phase.

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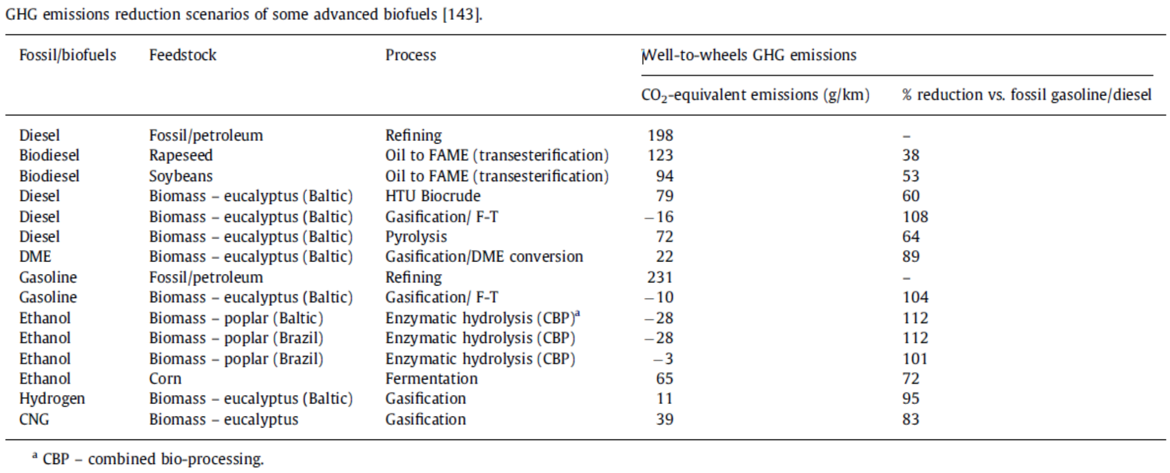
3 INTRODUCTION

3.1 BACKGROUND

Pyrolysis is a very cost effective process that can be used to produce bio fuel that can help us sustain our depleting resources of fossil fuels. Our heavy reliance on the fuel derived from crude oil means that it will be difficult to replace so easily, hence the need to sustain this resource.



GHG emissions reduction scenarios of some advanced biofuels [143].

Fossil/biofuels	Feedstock	Process	Well-to-wheels GHG emissions	
			CO_2-equivalent emissions (g/km)	% reduction vs. fossil gasoline/diesel
Diesel	Fossil/petroleum	Refining	198	–
Biodiesel	Rapeseed	Oil to FAME (transesterification)	123	38
Biodiesel	Soybeans	Oil to FAME (transesterification)	94	53
Diesel	Biomass – eucalyptus (Baltic)	HTU Biocrude	79	60
Diesel	Biomass – eucalyptus (Baltic)	Gasification/ F-T	−16	108
Diesel	Biomass – eucalyptus (Baltic)	Pyrolysis	72	64
DME	Biomass – eucalyptus (Baltic)	Gasification/DME conversion	22	89
Gasoline	Fossil/petroleum	Refining	231	–
Gasoline	Biomass – eucalyptus (Baltic)	Gasification/ F-T	−10	104
Ethanol	Biomass – poplar (Baltic)	Enzymatic hydrolysis (CBP)[a]	−28	112
Ethanol	Biomass – poplar (Brazil)	Enzymatic hydrolysis (CBP)	−28	112
Ethanol	Biomass – poplar (Brazil)	Enzymatic hydrolysis (CBP)	−3	101
Ethanol	Corn	Fermentation	65	72
Hydrogen	Biomass – eucalyptus (Baltic)	Gasification	11	95
CNG	Biomass – eucalyptus	Gasification	39	83

[a] CBP – combined bio-processing.

Table 3-1

The emissions from diesel and petrol/gasoline from crude oil is more harmful as compare to their bio oil versions. A concise comparison is shown in *Table 1-1 (Hossain & Davies, 2013).*

Pakistan produces 69 million tons of agricultural waste annually according to the US Agricultural Department. This can be used as a source of energy to power our homes, vehicles and industries. It is estimated that bio oil from the pyrolysis technology is capable of replacing 33% of crude oil globally.(Hernandez, 2011) Keeping this in mind, the process described in this thesis was designed to answer some, if not all, of our energy problems.

3.2 STATEMENT OF THE PROBLEM

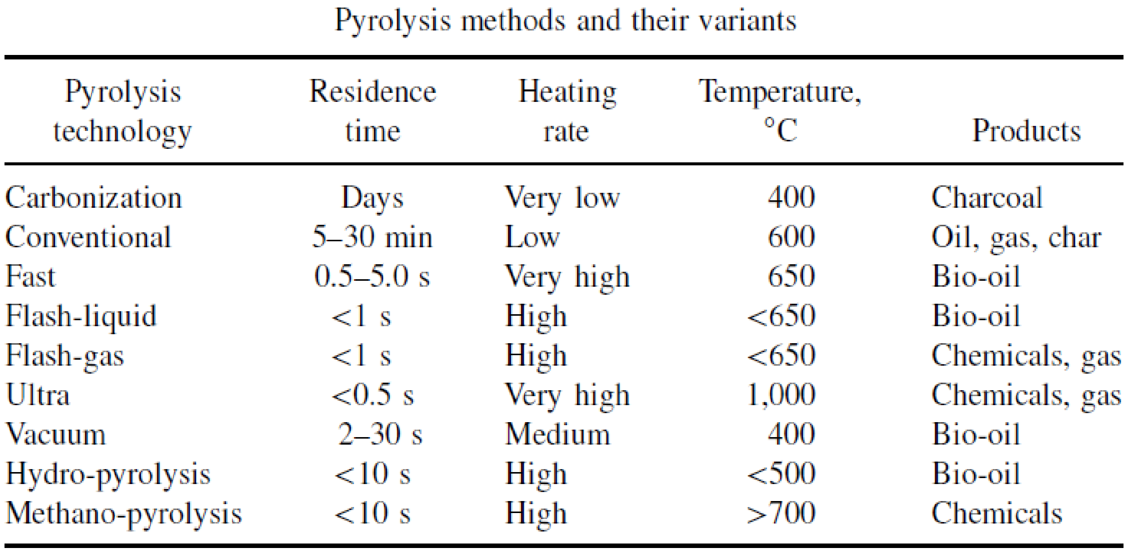
The increasing energy crisis in Pakistan is a major issue that needs to be addressed immediately. The natural resources for fuel are depleting at an alarming rate. The price of oil is greater now than it was ten years ago, with estimates that this will continue to rise. A solutions to all these problems must be found in time.

3.3 Purpose of the Study

The purpose of this project is to help resolve the issues related to energy, especially fuel, by using a cost effective technique to produce bio oil which can be blended with normal diesel.

Pyrolysis was selected over combustion or gasification as it was the process that suited all of our needs. Combustion produced products that were harmful to the environment and gave off less heat as compared to bio oil from pyrolysis. Gasification resulted in gaseous products that could not be blended with petrol and diesel. Also combustion and gasification required higher temperatures as compared to pyrolysis (800-1200 °C for combustion/gasification and 350-600 °C for pyrolysis).

Fast Pyrolysis was chosen over rest because of the lower residence time and lower temperature. The details of the pyrolysis techniques are shown in *Table 1-2*.



Pyrolysis methods and their variants

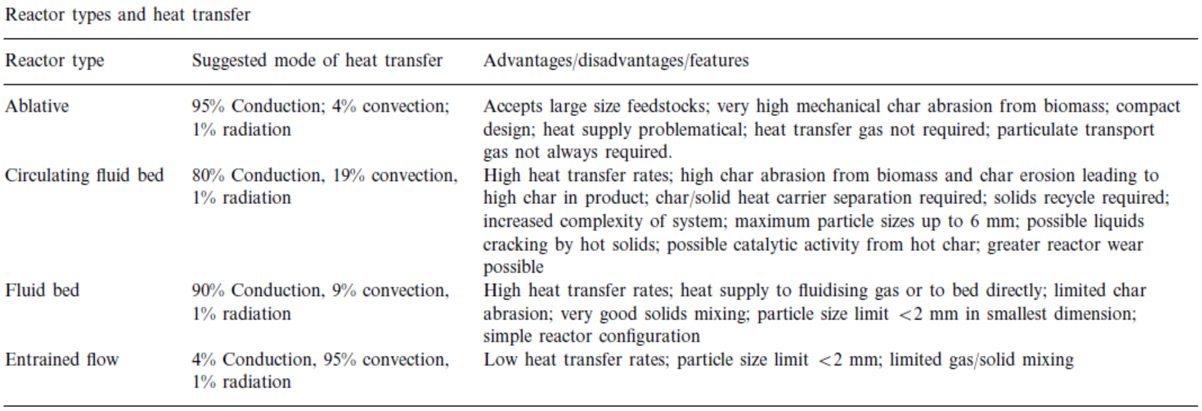
Pyrolysis technology	Residence time	Heating rate	Temperature, °C	Products
Carbonization	Days	Very low	400	Charcoal
Conventional	5–30 min	Low	600	Oil, gas, char
Fast	0.5–5.0 s	Very high	650	Bio-oil
Flash-liquid	<1 s	High	<650	Bio-oil
Flash-gas	<1 s	High	<650	Chemicals, gas
Ultra	<0.5 s	Very high	1,000	Chemicals, gas
Vacuum	2–30 s	Medium	400	Bio-oil
Hydro-pyrolysis	<10 s	High	<500	Bio-oil
Methano-pyrolysis	<10 s	High	>700	Chemicals

Table 3-2

3.4 Contributions of the Study

The process described in this thesis report is a proposal to solve some of Pakistan’s fuel related issues. Special considerations have been taken as to define all the processes in details so that the fabrication of a pilot plant can be done without many hindrances. The authors hope that this project will play a part, no matter how small, in resolving some of the issues identified in the report.

Rotating Cone Reactor was the main feature of the project. A small reactor was capable of processing large amounts of biomass. Design and calculations are discuss further in this thesis report. A brief comparison of the reactor types used in pyrolysis along with their respective heat transfer modes and possible advantages and disadvantages is shown in *Table 1-3 (Bridgwater, 1999).*

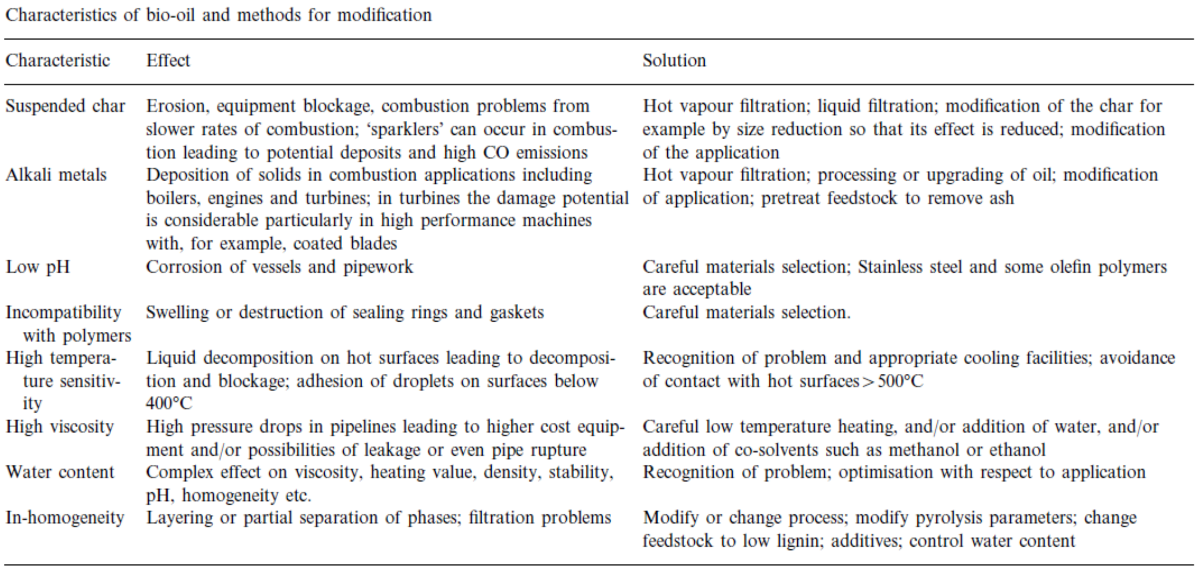


Reactor types and heat transfer

Reactor type	Suggested mode of heat transfer	Advantages/disadvantages/features
Ablative	95% Conduction; 4% convection; 1% radiation	Accepts large size feedstocks; very high mechanical char abrasion from biomass; compact design; heat supply problematical; heat transfer gas not required; particulate transport gas not always required.
Circulating fluid bed	80% Conduction, 19% convection, 1% radiation	High heat transfer rates; high char abrasion from biomass and char erosion leading to high char in product; char/solid heat carrier separation required; solids recycle required; increased complexity of system; maximum particle sizes up to 6 mm; possible liquids cracking by hot solids; possible catalytic activity from hot char; greater reactor wear possible
Fluid bed	90% Conduction, 9% convection, 1% radiation	High heat transfer rates; heat supply to fluidising gas or to bed directly; limited char abrasion; very good solids mixing; particle size limit <2 mm in smallest dimension; simple reactor configuration
Entrained flow	4% Conduction, 95% convection, 1% radiation	Low heat transfer rates; particle size limit <2 mm; limited gas/solid mixing

Table 3-3

There were some reservations as to some of the properties of the product of the pyrolysis but oil from the rotating cone reactor proved to be of a higher quality than the normal pyrolysis oil. Shown in *Table 1-4 (Bridgwater, 1999).*



Characteristics of bio-oil and methods for modification

Characteristic	Effect	Solution
Suspended char	Erosion, equipment blockage, combustion problems from slower rates of combustion; 'sparklers' can occur in combustion leading to potential deposits and high CO emissions	Hot vapour filtration; liquid filtration; modification of the char for example by size reduction so that its effect is reduced; modification of the application
Alkali metals	Deposition of solids in combustion applications including boilers, engines and turbines; in turbines the damage potential is considerable particularly in high performance machines with, for example, coated blades	Hot vapour filtration; processing or upgrading of oil; modification of application; pretreat feedstock to remove ash
Low pH	Corrosion of vessels and pipework	Careful materials selection; Stainless steel and some olefin polymers are acceptable
Incompatibility with polymers	Swelling or destruction of sealing rings and gaskets	Careful materials selection.
High temperature sensitivity	Liquid decomposition on hot surfaces leading to decomposition and blockage; adhesion of droplets on surfaces below 400°C	Recognition of problem and appropriate cooling facilities; avoidance of contact with hot surfaces > 500°C
High viscosity	High pressure drops in pipelines leading to higher cost equipment and/or possibilities of leakage or even pipe rupture	Careful low temperature heating, and/or addition of water, and/or addition of co-solvents such as methanol or ethanol
Water content	Complex effect on viscosity, heating value, density, stability, pH, homogeneity etc.	Recognition of problem; optimisation with respect to application
In-homogeneity	Layering or partial separation of phases; filtration problems	Modify or change process; modify pyrolysis parameters; change feedstock to low lignin; additives; control water content

Table 3-4

3.5 ORGANIZATION OF THE REMAINDER OF THE STUDY

The authors have done their best in organizing every single detail in the process designed. It is hoped that the contents of this report should be sufficient for anyone looking to implement this process on a practical level. The specifications of the design are precise and some extra margins for contingencies have been considered throughout the designing phase of the project.

Further up-gradation of the product is possible by the removal of oxygen content.(Gabriele Centi & Santen) The lower the oxygen levels in the oil, the better will be the calorific value and it will be easier to blend it with the fuels. There are several ways to do this, some of which are discussed below:

- Decarboxylation – This is the removal of Oxygen as CO_2, as against water, and it improves the energy content of the Oil. This also reduces the acidity and corrosivity of the Bio Oil.
- Hydro-deoxygenation – A single ton of biomass requires 62 kg of H_2 which makes this process highly expensive. Also this requires that the products are refined before they can be added to diesel and gasoline.
- Cracking over Zeolite – Deep cracking and gas formation reduces liquid yields and deep de-oxygenation yields an aromatic product. Coke formation in this process leads to severe catalyst deactivation.
- Tail Gas Optimization – Through private communication with Mr. Najaf Awan (Ph.D. Scholar at Punjab University) the authors found out about some astounding results using this technique. Incondensable gases from the pyrolysis process can be recycled in the reactor to improve the yield and quality of the Bio Oil.

4 LITERATURE REVIEW

4.1 SURFACTANT WASHING

There are two main problems with washing, the first being that the biomass has to be subsequently dried so that the moisture content is below 10% due to fast pyrolysis requirements. Secondly, if the biomass is acid washed the acid has to be separated and recovered or disposed, which increases the operational costs of pre-treatment.

Addition of surfactant like Triton X-100 helps to increase the efficiency of inorganic removal by increasing cell permeability and should be considered when examining the economic merit, due to less water being retained by biomass; therefore reducing drying requirements.

Triton X-100 is biodegradable and relatively large amounts of the surfactant and its biodegraded by-products are currently released into the environment. Thus it can be used in a washing solution without having to be separated and recovered or disposed of after use, which saves on operational costs.(Banks, Nowakowski, & Bridgwater, 2014)

4.2 PYROLYSIS PRODUCT AS TRANSPORTATION FUEL

Pyrolysis can convert biomass from a variety of sources –including agricultural and forestry residues – into liquid, solid and gaseous forms. All three output fractions have potential as fuels (either directly or after up-gradation) in various types of prime mover for transport, power generation, combined heat and power(CHP) or combined cooling heat and power (CCHP). There are two main pyrolysis techniques: slow and fast. Slow pyrolysis has been used for many centuries in traditional charcoal making but can also be carried out using modern reactors.

The products (yields, properties and composition) of pyrolysis are dependent on the operating variables, biomass feedstock type, pyrolysis process used and reactor design. Operating parameters such as reaction temperature, residence time, inert gas flow rate and particle size affects the quantity and quality of pyrolysis liquid.

Pyrolysis liquid is a complex emulsion of water and oxygenated hydrocarbons. In crude condition and at room temperature it is a dark-brown free flowing liquid. Following hot-vapor filtration it may acquire a more translucent red-brown appearance due to the removal of char.

Some of the common problems associated with the combustion of the crude PL in standard CI engines are (i) difficulty in starting the engine due to the poor ignition quality of pyrolysis liquid, and subsequent unstable operation; (ii) higher fuel consumption than fossil diesel operation (in some cases it was not possible to attain full power due to the limitations of the fuel pumps and injectors capacity); (iii) higher CO emission (though NOx is typically lower than corresponding fossil diesel operation); (iv) corrosion.

4.2.1 Modifications required for engines for using pyrolysis liquid

Pyrolysis liquid up-gradation techniques discussed in the literature are concentration, hydrodeoxygenation, catalytic cracking, emulsification, esterification (blending/addition of chemicals such as alcohols), and steam reforming.

- Pyrolysis gas can substitute natural gas or producer gas in SI engines heating value, flame speed velocity and ignition energy are important parameters regarding the feasibility of using such low quality gaseous fuels. Low flame velocity may lead to knocking; nonetheless, the literature has reported a similar power output of SI engines when fuelled with PG instead of NG or methane.

- Storage stability, and corrosion and spray characteristics of pyrolysis liquid

Pyrolysis liquid is normally unstable; re-polymerization occurs with time and hence viscosity increases. Lighter compounds present in the pyrolysis liquid cause this ageing process. During ageing pyrolysis liquid tends to separate into various tarry, sludgy, waxy, and thin aqueous phases. The volatility of pyrolysis liquid decreases with time.(Hossain & Davies, 2013)

4.3 VACUUM PYROLYSIS FOR TRANSPORTATION FUEL

Pyrolysis oil is a valuable liquid fuel, as a mixture of hydrocarbons, fatty acids and methyl esters. A potential diesel fuel can be easily produced from the pyrolysis oil via esterification. This work provides the important data for scaling the vacuum pyrolysis process. Cracking process is easily carried by vacuum pyrolysis.

The pyrolysis oils obtained over alkali metal hydroxides (NaOH, KOH) showed the best evaporation properties. The next were the pyrolysis oils over alkali metal carbonates (Na_2CO_3, K_2CO_3). The non-catalytic pyrolysis oil had the lowest volatility, and it was similar to the pyrolysis oils over alkali metal chlorides (NaCl, KCl) and sulphates (Na_2SO_4, K_2SO_4). In the DTG curves, the peak temperatures appeared in the order of hydroxides < carbonates < non-catalyst ≈ chlorides ≈ sulphates.(Tang, Zheng, Wang, & Wang, 2014)

4.4 BIOFUELS FROM PYROLYSIS OIL

It was assumed that the woody biomass is the raw material for the process which is converted into liquid using the fast pyrolysis process. The excess oxygen from the oil is removed by means of catalytic hydro-cracking. Oxygen has to be removed as this lowers the calorific value of the bio oil. The final product is a direct replacement for gasoline and diesel. Wood chips are a waste and can be used in this process to produce biofuels. Forest residues and trimming can also be used as a likely source of raw materials. Although the variation in quality may result in process difficulties.

This process is economically feasible as the cost of the raw material i.e., woody biomass, is low and it is also abundantly available. Logistics cost will not be much of a factor as the facilities will be setup in the vicinity of the raw material, for example near pulp mills, petroleum refineries, etc. The production is very cost effective, making whole of the process quite feasible. The major fraction of the cost will be associated with the removal of oxygen.

It may be more economical to convert the heaviest upgraded pyrolysis oil in existing refineries or to blend it in the fuel oil. Equipment required are the same as those the refineries and chemical industries use at a large scale. For hydro-treating it is more attractive to covert Hydrogen from natural gas rather than from the process by-products. Also the catalyst hydro-treating uses will need to be frequently or continuously regenerated. Even with Oxygen removed it will still be a crude oil substitute and will not work well with the current fuel. Although it can be blended with fuel. The cost of Oxygen removal increases as the residual Oxygen levels decrease, but pyrolysis oil with less than 10% Oxygen content are miscible with petroleum products.(Arbogast, Bellman, Paynter, & Wykowski, 2012)

4.5 COMPARISON OF SLOW AND FAST PYROLYSIS

Parameter	Slow Pyrolysis	Fast Pyrolysis
Heat Transfer Rate	0.1 – 1 °C/s	10 – >1000 °C/s
Residence Time	Minutes to Hours	2 – 5 sec
Temperature	400 – 600 °C	400 – 650 °C

Table 4-1

(Bridgwater, 1999)

The pyrolysis oil can be burnt directly for static applications like in boilers and furnaces. They are also immiscible with fossil fuels but 35-40% miscible with water. They have a tendency to age due to light, O_2 and heat above 80 °C and their low pH causes some corrosion issues. Ideally low O/C and high H/C ratios are desirable.

4.6 UPGRADING TECHNIQUES FOR BIO OIL FROM FAST PYROLYSIS

This is basically lignocellulosic pyrolysis which uses zeolites for hydro-deoxygenation and hydrogenation of bio oil along with a wide variety of metal catalysts.

Biomass have naturally present ions of silica, Na, K, Mg and Ca that demonstrate catalytic properties under pyrolysis conditions. These minerals are normally stripped from the bio mass and then introduced individually according to their need.

4.6.1 Hydro-Deoxygenation

In HDO Carbon-Oxygen bonds are cleaved with Hydrogen in the presence of a catalyst producing CO_2 and H_2O, by removing Oxygen from the product. This can be achieved through sulfide/oxide and transition metal catalyst through a combination of chemisorption, proton donation and desorption. Catalyst reactivity relies on Lewis acid and Bronsted acid number and strength.

The advantages are that product is energy dense, non-corrosive and resembles naphtha. Also it is very easily separated from water.

Disadvantages include consumption of large quantities of high pressure Hydrogen gas, at around 600-700 L of H_2/kg Bio Oil. This is expensive as the cost of Hydrogen gas is very high and it is also difficult to handle.

HDO of wood based bio oil is done over a Pd/C catalyst inside a fixed bed reactor. The pressure is kept at 140 bar and 340 °C temperature.

4.6.2 Hydrogenation

In this process Hydrogen is co-fed with biomass at 75-300 bar pressure. This high pressure helps in increasing the solubility of H_2 in Bio Oil. Ru/C catalyst is used at 350 °C and 200 bar pressure inside an autoclave. The yield is highest for a 4 hour reaction time. The coke formation increases with time and temperature increases with deoxygenation.

Cobalt molybdate catalyst can also be used at 360-390 °C and 5 min to 1 hour of reaction time. In this case deoxygenation increases inside an autoclave.

4.6.3 Cracking with Zeolites

Zeolites are complex, three dimensional porous structures. Cracking with zeolites produce aromatic products at atmospheric pressure without the need for H_2 gas. The final product though has a lower heating value due to low H/C and high O/C ratios as compared to HDO. The reactions parameters are 350-600 °C.

(Dickerson & Soria, 2013)

5 PROCESS DESCRIPTION

The pyrolysis process in this thesis is designed to give the maximum yield for bio oil over gases and solids. The by-products that are produced i.e., incondensable gases and the bio char is used to power the necessary operations in the plant. Pyrolysis should ideally occur in the absence of Oxygen but the reactor used in this process has no such requirement, giving a good yield of pyrolysis oil even in the presence of Oxygen. Please note that the presence of Oxygen does NOT cause any problems and combustions does not occur. The feed set for the plant designed is at 1.25 tons/hr of bio mass giving 0.75 tons/hr of bio oil. The rest of the process description is explained in detail.

5.1 PRE-TREATMENT

5.1.1 Initial Screening

This screening is done using filters to remove any impurities that can be separated easily from the feed. This may include rocks, stones and other such big particles.

5.1.2 Washing

This process uses an industrial grade washer to wash the feed of any impurities that may have escaped the initial screening. This normally includes unwanted sand and dirt particles. After washing it is expected that the weight of the feed will increase by 10% due to the increased moisture content.

5.1.3 Drying

Drying is necessary to remove unnecessary moisture from the feed particularly after the feed passes through the washing process. This process utilizes a screen conveyor dryer that consists of a conveyor belt which carries the biomass in the drying chamber. Inside the drying chamber, steam is blown in to remove the additional moisture in the biomass through direct contact.

5.1.4 Size Reduction

The size reduction of the fibrous biomass is quite problematic as it cannot be done effectively using impact forces. Hence in this process, a jaw crusher is used that uses shearing forces to bring the biomass to the prescribed size i.e., 3 mm.

5.1.5 Final Screening

Final screening is done to ensure that the biomass being sent to the reactor is no greater than 3 mm in size. Using a filter mesh it is estimated that the efficiency of this process will be around 60%.

5.2 STORAGE

After the pre-treatment the biomass is stored in storage silos. This is done so that the reactor does not face a deficiency of biomass when the main process is in motion. The residence time for the reactor is only 5 seconds meaning that it can process a lot of feed in a short amount of time, hence storage silos are necessary to ensure that the process remains continuous.

5.3 MAIN PROCESS

5.3.1 Reactor

As mentioned before the process described in this thesis report uses a very novel reactor technology, first proposed and patented by University of Twente. This is the rotating cone reactor, which does not need an inert environment for pyrolysis to occur. It uses sand, which enters from the bottom of the cone, as the heating medium. The hot sand which enters at approximately 450-500 °C interacts with the biomass at high frequency which results in more efficient heat transfer. The sand leaves the reactor at a lower temperature than at which it entered and is sent into the char combustor to be heated again. The vapors formed from the pyrolysis include several compounds which are then sent to the condenser to separate oil from gases. The process also produces char which is sent along with the sand into the char combustor, where it is used to reheat the sand.

5.3.2 Char Combustor

Char combustor has the used up sand from the reactor and by-product, char coming in as the feed. The char is burnt in the combustor and is used to heat up the cool sand. This produces some flue gases and ash, all of which is separated in the cyclone separator.

5.3.3 Cyclone Separator

The flue gases alone with the ash come into the cyclone at a high velocity and they are separated according to their relative densities. Ash being the heaviest comes out of the bottom of the cyclone which the flue gases are passed into the after burner where they are used for combustion of incondensable gases.

5.3.4 Condenser

Condenser is extremely important as this is where the bio oil is separated from the gas. It is estimated the overall yield of the plant is 60% so the oil produced is approximately 0.75 tons/hr. The incondensable gases are sent to the after burner where they are combusted and the heat evolved is used to make steam.

5.4 TURBINES

The process described produces a lot of energy that can be used to completely power the plant. Steam is produced by the cooling water from the condenser and the char combustor, and also from the after burner. The steam from the after burner is at a higher temperature of about 1200°C and 10,000 kPa. This steam is used in a steam turbine where its outlet temperature is 230 °C and 1000 kPa.

The steam from the condenser and the char combustor are also at 230°C and 1000 kPa. These are combined with the steam coming in from the first turbine and sent into the second turbine, which reduces the temperature to 30°C and 100 kPa. Together these two turbines produce a power of 218.4 kW which is enough to power whole of the plant, including drying, crushing and the plant utilities.

5.5 PROCESS FLOW DIAGRAM

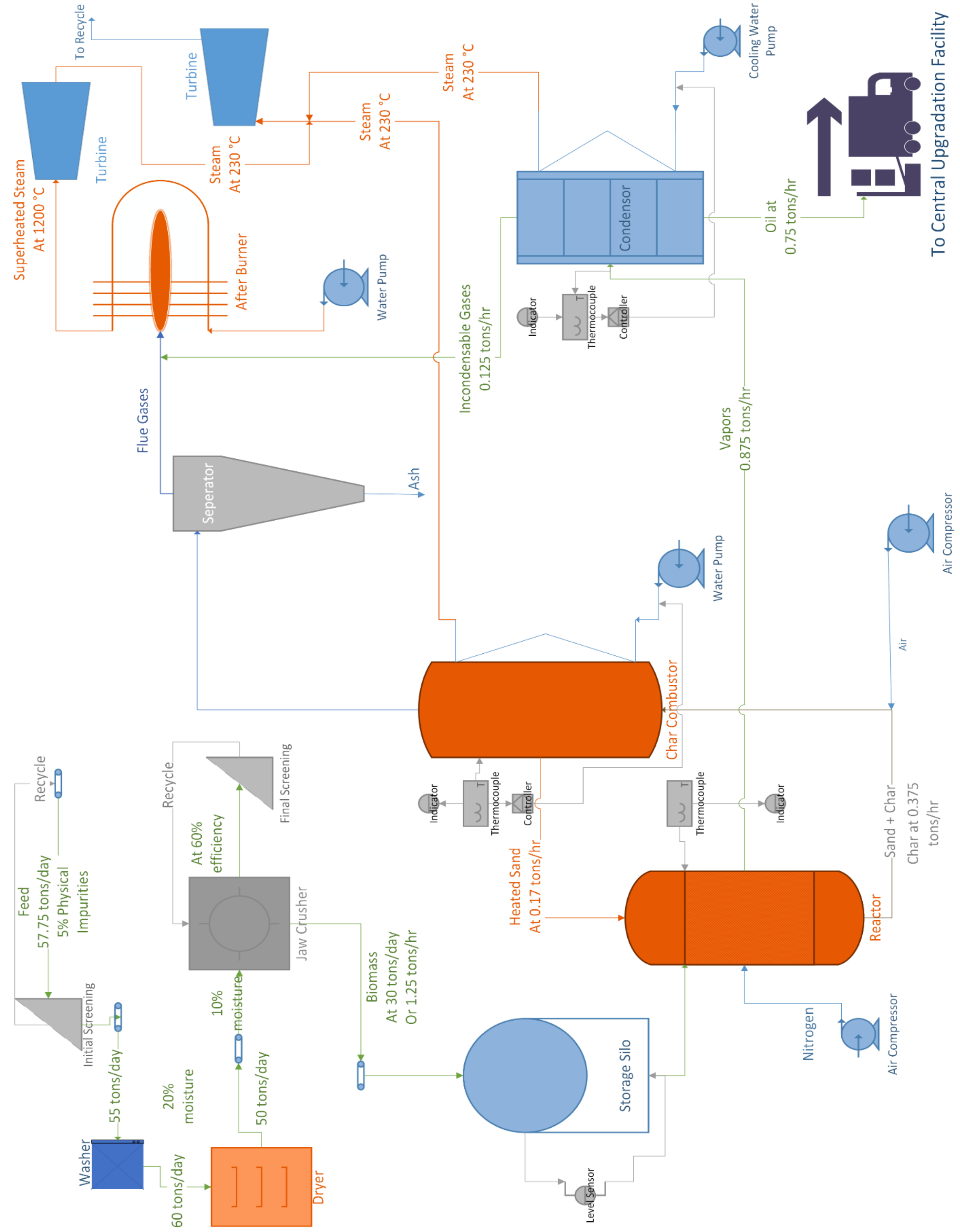


Figure 5-1

6 PROCESS CONTROL SYSTEM

The control system attached in it is a cascade control loop.

A cascade control arrangement, there are two (or more) controllers of which one controller's output drives the set point of another controller. For example: a level controller driving the set point of a flow controller to keep the level at its set point. The flow controller, in turn, drives a control valve to match the flow with the set point the level controller is requesting.

6.1 REACTOR

The control system attached in it is a cascade control loop.

A cascade control arrangement, there are two (or more) controllers of which one controller's output drives the set point of another controller. For example: a level controller driving the set point of a flow controller to keep the level at its set point. The flow controller, in turn, drives a control valve to match the flow with the set point the level controller is requesting.

Temperature Controller: Master Control Loop

Flow Controller: Slave control Loop

- Thermocouple in the reactor senses the temperature and gives the signal to temperature controller
- The temperature controller is attached in a cascade system with the flow controller.
- Hence flow controller takes the input from both temperature sensor and the flow sensor and according to the set point given, it regulates the supply of heated sand through the Flow control valve.

6.2 COMBUSTOR

Temperature Controller: Master Control Loop

Flow Controller: Slave control Loop

- Thermocouple in the combustor senses the temperature and gives the signal to temperature controller
- The temperature controller is attached in a cascade system with the flow controller.
- Hence flow controller takes the input from both temperature sensor and the flow sensor and according to the set point given, it regulates the supply of water through the Flow control valve.

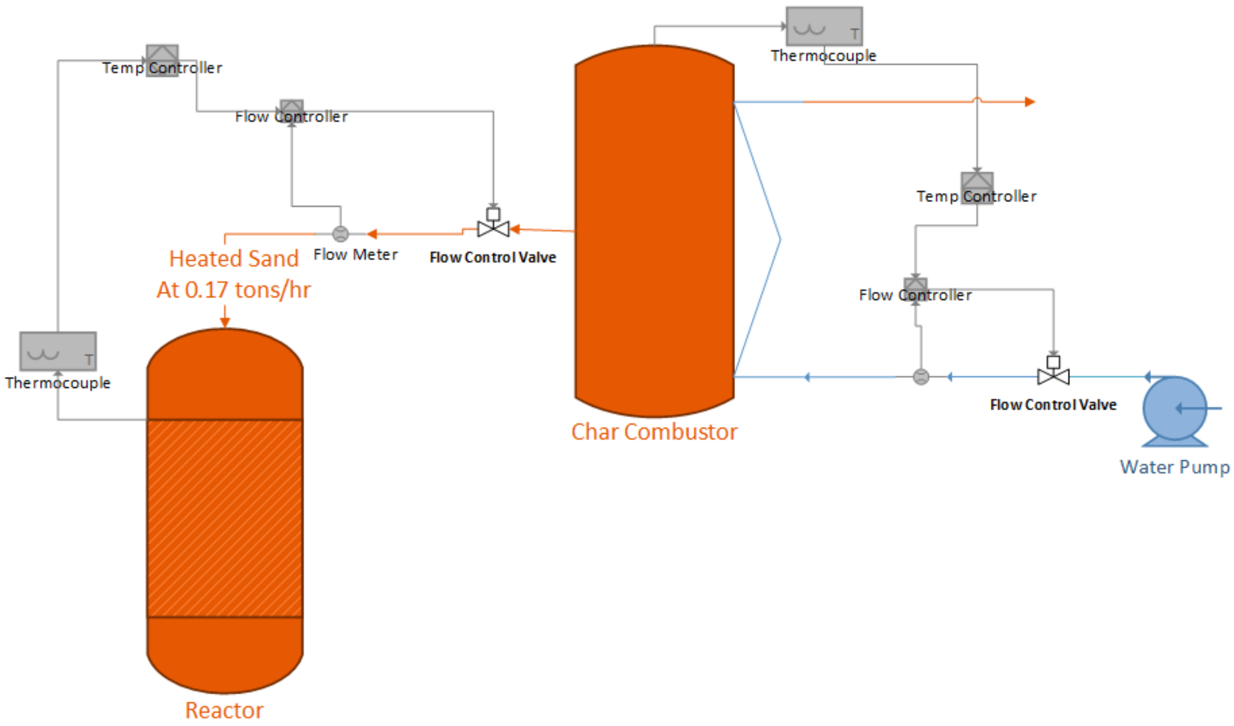

Temp Controller
Flow Controller
Thermocouple
Temp Controller
Flow Controller
Heated Sand
At 0.17 tons/hr
Flow Meter
Flow Control Valve
Thermocouple
Char Combustor
Flow Control Valve
Water Pump
Reactor

Figure 6-1

6.3 CONDENSER

Temperature Controller: Master Control Loop

Flow Controller: Slave control Loop

- Thermocouple in the condenser senses the temperature and gives the signal to temperature controller
- The temperature controller is attached in a cascade system with the flow controller.
- Hence flow controller takes the input from both temperature sensor and the flow sensor and according to the set point given, it regulates the supply of water through the Flow control valve.

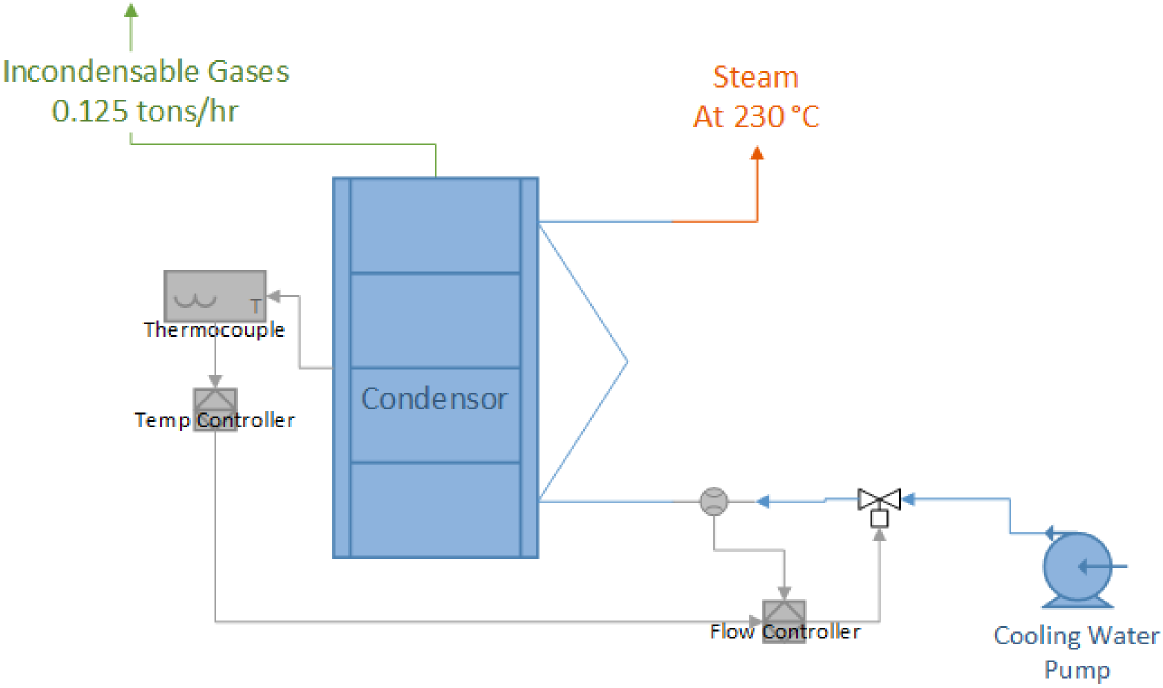


Figure 6-2

7 MASS BALANCE

Objective: To make sure the feed enters at 30 tons per day

7.1 INITIAL SCREENING

Mass entering: 57.75 tons per day at 5% impurity, 10% moisture

Mass leaving out: 57.75 – (0.05) (57.75)

=57.75- 2.75

= 55 tons per day

Assuming all impurities have been removed by the screener.

7.2 INDUSTRIAL WASHER

Mass in: 55 tons/day, 10% moisture

Mass out: 20% moisture

= (0.20) (55)/0.10

= 60 tons per day

7.3 SCREEN CONVEYOR DRYER

Mass in: 60 tons per day at 20% moisture

Mass out: Assuming all moisture removed

= 60 – (0.20) (60)

= 50 tons per day

Jaw Crusher

Mass in= Mass out

Mass out: 50 tons per day feed

7.4 FINAL SIEVE SHAKE

60% efficiency of the sieve shakers

Diameter size of particle< 3mm

Mass out: (0.60) (50)

= 30 tons per day feed

7.5 ROTATING CONE REACTOR

Mass in: 30 tons/ day or 1.25 tons/ hr feed

Efficiency of the rotating cone reactor= 70%

Mass of Bio vapors formed= (0.70) (1.25)

= 0.875 tons per hour

Char formed= 1.25- 0.875

= 0.375 tons per hour

Mass out= 0.875 tons/hr Bio vapors + 0.375 tons/hr char

7.6 CHAR COMBUSTOR

Mass in char = Mass out char

Mass out char = 0.375 tons per hour

High Heating Value (Char) = 21.93 MJ/kg (*value obtained from energy balance)

Heat Balance for Sand and Char

$Q_{char}= Q_{sand}$

Q_{char} = 21.93 MJ/kg

$Q = m_s c_p \Delta T$

$(21.93 \times 10^6) = m_s (860) (550-400)$

m_s= 170 kg/h OR 0.170 tons/h

Hence mass of sand needed= 0.17 tons per hour sand used

Mass of air needed

Mass of Bio Char= 375 kg/hr or 0.375 tons/hr

Bio Char

- O = 24.8%
- C = 67.2%
- H = 1.0%
- Ash = 7%

Carbon: 0.672 x 375 = 252 kg

Hydrogen: 0.01 x 375 = 3.75 kg

Oxygen= 0.266 x 375 = 93 kg

Ash= 0.07 x 375 = 26.25 kg

Moles of oxygen needed

$C + O_2 \longrightarrow CO_2$

Carbon	Oxygen
12g	32g
252kg	x

X = (252)(32)/ 12

= 572kg of oxygen needed

$2H_2 + O_2 \longrightarrow 2H_2O$

Hydrogen	Oxygen
2g	16g
3.75 kg	x

X= (3.75) (16)/(2)

= 30 kg of oxygen needed

Total Mass of Oxygen needed

=Mass for Hydrogen + Mass for Carbon + Mass of oxygen present in the fuel

= 30 + 672 + 93

= 795 kg

Total Moles of Oxygen= 795/ 16

= 43.28 kmols

Moles of Oxygen present in fuel = Mass of oxygen in fuel/ Molecular Mass

=93/16

= 5.813 kmols

Total Moles of Oxygen needed from air= 43.28 – 5.813

= 37.46 kmols of Oxygen

21% of oxygen present in air

Total Air= 37.46/ 0.21

= 178.381 kmols of air

Air used is 9% in excess to avoid the formation of Carbon Monoxide and soot formation

Total air= (1.09) (178.381)

= 194.43 kmols of air

7.7 FLUE GAS COMPOSITION

Since all of Carbon is burned so no carbon monoxide and no soot is formed

7.7.1 Carbon Balance

Since 1 mole of Carbon forms 1 mole of C02 hence 252 kg of Carbon will form (252/12) kmols C02

Moles of CO2 = 252/12 = 21 kmols CO2

7.7.2 Hydrogen Balance

Since 1 mole of will form 1 mole of water so 3.75 kg of hydrogen will form (3.75/1) kmols of water

Moles of H20 = 3.75/2= 1.875 kmols

7.7.3 Oxygen Balance

21% present in air + total oxygen in the fuel

= (0.21) (194.43) + 5.813

= 46.64 kmols

7.7.4 Nitrogen Balance

78% present in the air

=(0.78)(194.43)

= 151.65 kmols

Hence,

CO_2= 21 kmols

H_2O= 1.875 kmols

O_2 = 46.64 kmols

N_2 = 151.65 kmols

So the total Flue Gas Compositions are

CO_2= 9.50%

H_2O = 0.80 %

O_2 = 21.10 %

N_2 = 68.60 %

7.8 CYCLONE SEPARATOR

7% ash with the char gases goes into the cyclone separator with the composition

CO_2= 9.50%

H_2O = 0.80 %

O_2 = 21.10 %

N_2 = 68.60 %

Total Mass in= 0.375 tons/hr char

Total Mass out = 7% Ash + Flue Gases

Ash= (0.07) (0.375)

= 0.026 tons/hr

Total Flue Gases= 0.375 – 0.026

= 0.349 tons/ hr

7.9 CONDENSER

Bio oil produced are 60% of the original cotton stalks fed into the reactor

Mass in= 0.875 tons/hr biomass

Mass out = (0.60) (1.25)

= 0.75 tons/hr bio oil

Mass of incondensable Gases = 0.875 – 0.75

= 0.125 tons/ hr

7.10 SUMMARY OF MASS BALANCE

Equipment	**Mass in (tons/hr)**	**Mass out (tons/hr)**
Initial Screening	2.40	2.29
Industrial Washer	2.29 (10% moisture)	2.5 (20% moisture)
Dryer	2.5 (20% moisture)	2.08
Jaw Crusher	2.08	2.08
Final Screening	2.08	1.25
Reactor	1.25 (Biomass)	0.875 + 0.375 (Vapors + Char)
Condenser	0.875	0.75 + 0.125 (Oil + Gases)
Combustor	0.375	0.375
Cyclone Separator	0.375	0.349 + 0.026 (Gases + Ash)

Table 7-1

8 ENERGY BALANCE

8.1 BIO CHAR

O = 24.8%

C = 67.2%

H = 1.0%

Ash = 7%

8.2 BIO OIL

C = 59.7%

H = 7.8%

N = 1.8%

O = 36.6%

(Demirbas, 2004)

HHV= 0.3491*C + 1.1783*H - 0.1034*O - 0.015*N - 0.0211*Ash

HHV (Char) = 21.93 MJ/kg

HHV (Bio Oil) = 26.2 MJ/kg

(Chantal Bergeron, Danielle Julie Carrier, & Ramaswamy, 2012)

8.3 STANDARD POWER REQUIREMENTS

- Jaw Crusher: 7.5 kW
- Screen Conveyor Dryer: 3.5 kW

The energy required for every kg of water evaporated equals the energy for every 2kg steam produced.

(McCabe, Smith, & Harriott, 1993)

8.4 THE ROTATING CONE REACTOR

The BTG technology uses 0.2 tons/hr processing of the biomass produced and the power required to move the motor is 23 kW. Hence the rotating Cone reactor can easily be scaled up for the energy requirement.

Since we are processing 1.25 tons/hr of cotton stalk so the power requirement hence the power requirement can be scaled down using to the unitary method.

23 kW energy used to process 2 tons/hr feed.

So to process 1.25 tons/hr feed, the power required is 14.375 kW.

(Kyritsis, 2000)

8.5 POWER OF CHAR COMBUSTOR COOLING WATER PUMP

Since the outlet steam that we see during our design feature is 230℃ so we need to calculate the amount of water needed for the energy balance equation

Cooling water at 100 kPa, 30°C is increased to 1000 kPa

By steam Tables:

Specific Volume= 0.1043 m^3/kg

Work required by pump = V (P2 – P1)

= 0.1043(1000-100)

= 93.87 kJ/kg

The amount of water used in cooling pump of a char combustor

Energy required by water = Energy given by char

Mass flow rate of sand = 170 kg/hr

Cp of sand = 860 J/kg℃

Initial temperature of sand= 400℃

Final Temperature of sand = 550℃

Change in temperature of water dT = (230-30) ℃

Cp of water = 4185.5 J/kg℃

$Q_{sand} = Q_{water}$

mc dT = mc dT

170 x 860 x (550 – 400) = m x 4185.5 x (230 – 30)

M_{water}= 26.2 kg/hr

Power of pump = (26.2 kg/hr x 93.87 kJ/kg) / 3600

= 0.683 kW

8.6 POWER OF COOLING WATER CONDENSER PUMP

Since the outlet steam that we see during our design feature is 230℃ so we need to calculate the amount of water needed for the energy balance equation

Cooling water at 100 kPa, 30°C is increased to 1000 kPa

By steam Tables:

Specific Volume= 0.1043 m^3/kg

Work required by pump = V (P2 – P1)

= 0.1043(1000-100)

= 93.87 kJ/kg

The amount of water used in cooling pump of the condenser

Energy required by water = Energy given by bio vapors

Mass flow rate of bio vapors = 875 kg/hr

Cp of bio vapors = 1300 J/kg℃

Initial temperature of bio vapors= 550℃

Final Temperature of bio oil = 50℃

Change in temperature of water dT = (230-30) ℃

Cp of water = 4185.5 J/kg℃

$Q_{bio\text{-}vapors} = Q_{water}$

mc dT = mc dT

875 x 1300 x (550 – 50) = m x 4185.5 x (230 – 30)

M_{water}= 679 kg/hr

Power of a water pump = (679 kg/hr x 93.87 kJ/kg) / 3600

= 17.7 kW

8.7 POWER OF AFTER BURNER PUMP

Steam at 10 MPa, 1200℃ is leaved and the water is entering at 30℃ , 100 kPa

So first we need to calculate the energy of the incondensable gases that enters the after burner.

Gases	Composition (%)	HHV (Btu /lb)
Hydrogen	3.145	51628
Oxygen	1.51	-
Nitrogen	34.27	-
Methane	5.64	20525
Carbon Monoxide	22.77	4368
Carbon Dioxide	30.04	14156
Ethylene	0.76	20525
Ethane	0.65	20295
Propene	0.54	19683
Propane	0.65	19834

Table 8-1

Total HHV of incondensable gases =

Σ (HHV of a gas x composition)

= 4301.7 Btu/lb = 10005 kJ/kg

Mass flow rate of incondensable gases = 125 kg/hr

Energy of incondensable gases = 10005 x 125 = 1250,000 kJ/hr

Energy lost by gases = Energy gained by water

Hence the temperature of water needs to be increased to 1200 ℃

$Q_{water} = m\ c\ \Delta\ T$

1250000 = m x 4.185 x (1200 -30)

m = 255.2 kg / hr

Work required by pump = V (Δ P)

0.1043 (10000 – 100)

= 103.257 kJ/kg

Power required by pump = (103.257 kJ/kg x 255.2 kg/h)/ 3600

= 7.32 kW

8.8 ENERGY PRODUCED FROM THE TURBINE

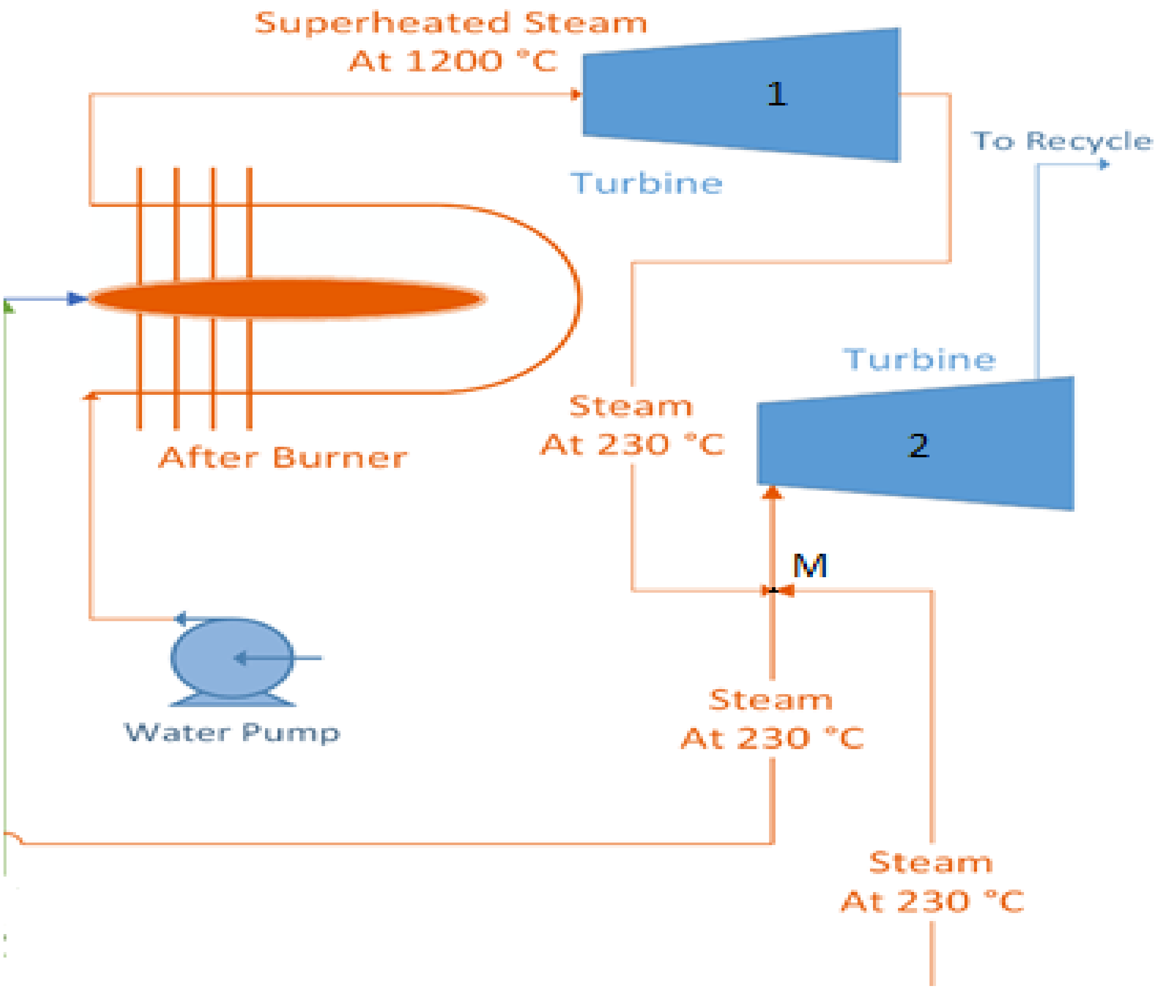


Figure 8-1

8.8.1 Steam Turbine 1 at the after burner

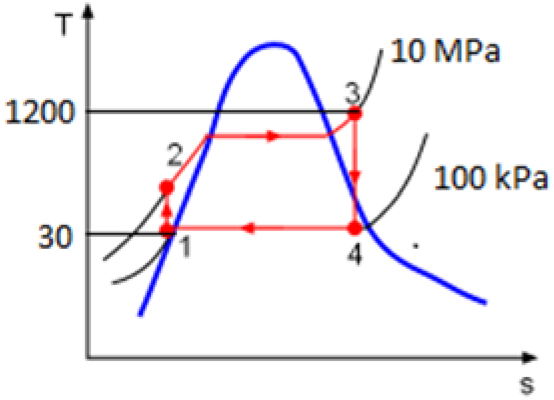


Figure 8-2

From the steam tables

30℃, 100 kPa

Enthalpy = 5131.7 kJ/kg

1200℃ , 10 MPa

Enthalpy = 2897.18 kJ/kg

Total work output produced at point 3 = 5131.7 – 2897.18

= 2234.52 kJ/kg

Total Mass flow rate of water used = 255.2 kg/hr

Total Power output = (255.2 kg/hr x 2234.52 kW) / 3600

= 158.9 kW

8.8.2 Energy Produced at Steam Turbine 2

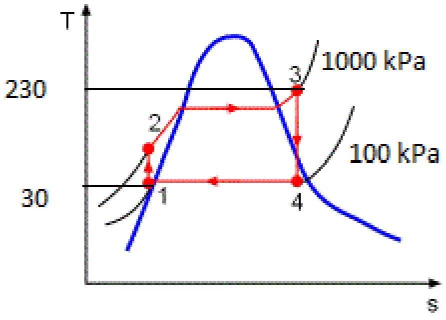


Figure 8-3

Work done by pump = 0.9387 kJ/kg

100 kPa, 30°C = 417.51 kJ/kg

h_2 = 417.51 + 0.9387

= 418.44 kJ/kg

1000 kPa, 230 °C

H_3= 2897.18 kJ/kg

S_3 =S4= 6.83414 kJ/kg

30°C, 100 kPa

H_f =417.5 kJ/kg

S_f = 1.302 kJ/kg

H_{fg} = 2257.5 kJ/kg

S_{fg} = 6.05 kJ/kg

X = quality of steam

X= (S_4 – S_f)/ S_{fg}

X= (6.83414 – 1.3028) / 6.0562

X = 0.913

$H_4 = h_f + Xh_{fg}$

$H_4 = 417.51 + (0.91334)(2257.5)$

$= 2479.36$ kJ/kg

$W_3 = h_3 - h_4$

$2897.18 - 2479.36$

$= 417.816$ kJ/kg

Since the steam meet at point "M" as shown by the above figure at same temperatures and pressure, hence their mass flow rates are added,

So the mass flow rate = 200 + 26.2 + 255.2 = 481.4 kg/hr

Power output = (481.4 kg/hr) (417 kJ/kg) / 3600

= 55.7 kW output

8.9 SUMMARY OF THE ENERGY BALANCE

Power required for plant	Power output in the process
Jaw Crusher = 7.5 kW **Screen Conveyer dryer = 3.5 kW** **Reactor = 14.375 kW** **Char combustor cooling pump = 0.693 kW** **Condenser Water pump = 17.1 kW** **Pump for after burner = 7.32 kW**	**Steam Turbine 1 = 159 kW** **Steam Turbine 2 = 55.7 kW**

Table 8-2

9 EQUIPMENT SPECIFICATION

9.1 SCREEN CONVEYOR DRYER

Component	Specification
Screen Mesh Size	30 mesh
Conveyor Width	1 m
Conveyor Length	10 m
Heating Time	20 mins
Energy Requirement	3 kW/h

Table 9-1

9.2 JAW CRUSHER

Component	Specification
Feed Opening	200x300
Max Feed Size	20 mm
Discharge Feed Size	1.5 mm
Reduction Ratio	12:1
Processing Capacity	3 tons/h
Rotational Speed	260 RPM
Motor Power	7.5 kW
Dimensions	910x750x990 mm

Table 9-2

*Calculations shown in *Energy Balance*

10 EQUIPMENT DESIGN

10.1 CHAR COMBUSTOR

10.1.1 Sand Specifications

Porosity = 0.5

Flow rate = 170 kg/h

Density = 1281 kg/m^3

Volumetric Flow Rate = $\frac{170\ kg/h}{1281\ kg/m^3}$ = 0.1327 m^3/h

10.1.2 Char Specifications

Porosity = 0.78

Flow rate = 375 kg/h

Density = 450 kg/m^3

Volumetric Flow Rate = $\frac{375\ kg/h}{450\ kg/m^3}$ = 0.833 m^3/h

10.1.3 Total Volumetric Flow

0.1327 + 0.833 = 0.966 m^3/h

10.1.4 Effective Porosity

$\frac{(0.1327)(0.5)+(0.78)(0.833)}{(0.966)}$ = 0.74

OR

Particle Size = 1.64 mm

10.1.5 Design Calculations

Residence Time = 120 sec

Volume Occupied = 1 – 0.74 = 0.26

$$t = \frac{V}{V°}$$

$$V_1 = \frac{0.966}{3600} \times 120s = 0.0322\, m^3$$

$$V_2 = \frac{0.0322}{0.26} = 0.1238\, m^3$$

Volume Allowance = 10%

$$V_3 = 0.0322 \times 0.10 = 0.00322 m^3$$

Total Volume = V_1+ V_2+V_3 = 0.159 m^3

L/D ratio = 3/1

L = 3D

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

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

D = 0.407 m

L = 1.22 m

(Sinnott, 2005)

10.2 CYCLONE SEPARATOR

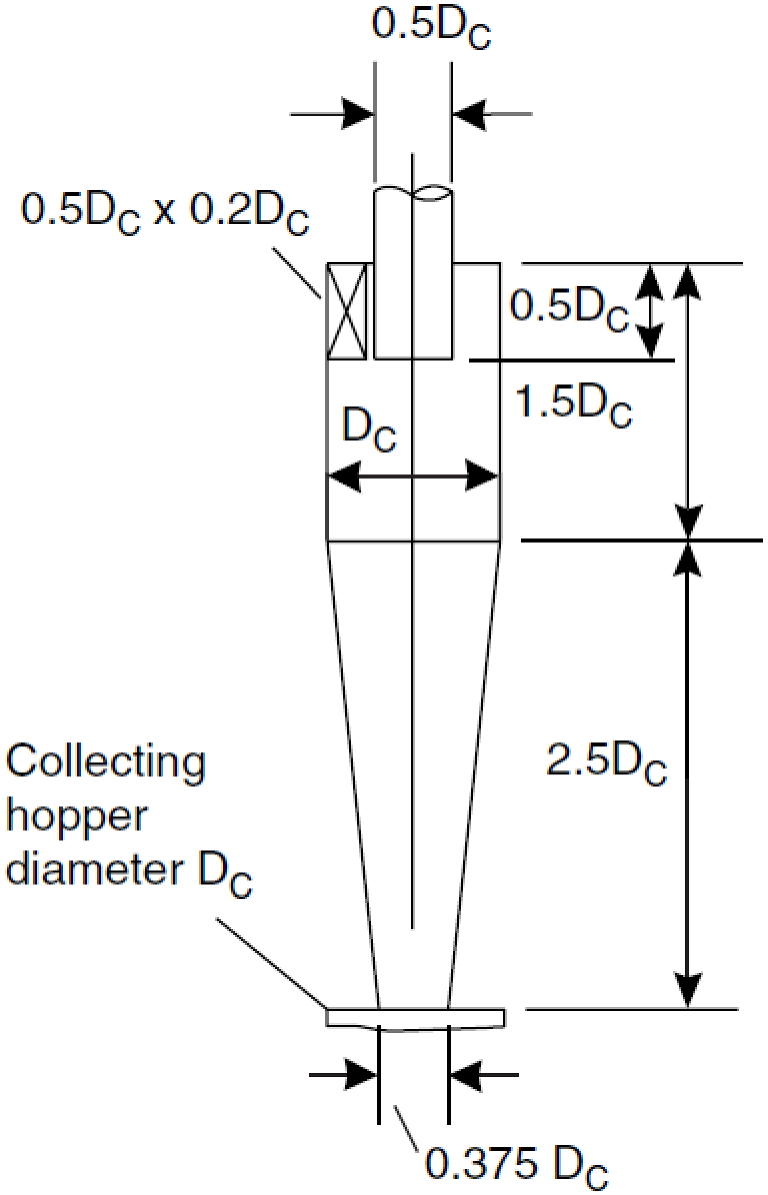


Figure 10-1

Taking standard diameter of cyclone, $D_c = 200$ mm

Duct area = $0.5D_c \times 0.2D_c = (0.5)(0.15)*(0.2)(0.15) = 2.25 \times 10^{-3}$ m^2

ρ_{gas} at 500 °C = $\frac{28}{22.4} \times \frac{273}{773} = 0.44$ kg/m^3

Viscosity of N_2 at 500 °C = 0.036 cp

A_1 = Area of inlet duct = 100*40 = 4000 mm^2

A_s = Cyclone surface area = π(200)[(300)+(500)] = 5.026 x 10^5 mm^2

Friction factor for cycl ones, f_c is taken as 0.005

$$\psi = \frac{f_c * A_s}{A_1} = \frac{0.005*5.026*10^5}{4000}$$

ψ = 0.63

$$\frac{r_t}{r_e} = \frac{200-((40)/2)}{100} = 1.8$$

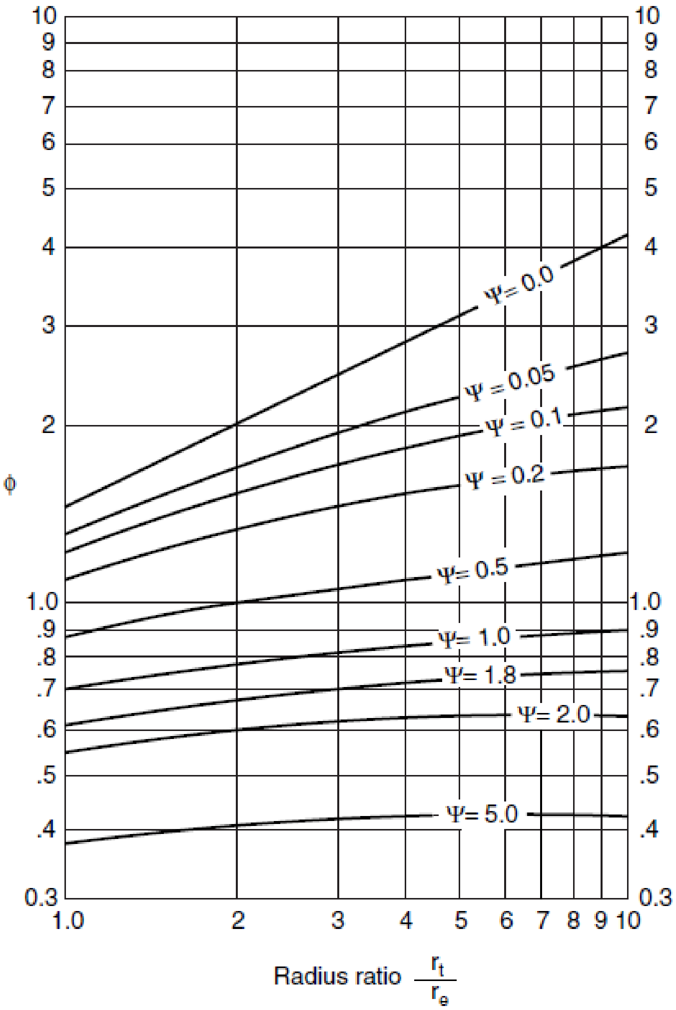


Figure 10-2

From the figure ϕ = 0.9

$u_1 = \frac{1000}{3600} \text{ X } \frac{10^6}{4000} = 69.4$ m/s

Area of exit pipe = $\frac{\pi * 10^6}{4} = 7854 \text{ mm}^2$

$u_2 = \frac{1000}{3600} \text{ X } \frac{10^6}{7854} = 35.4$ m/s

From equation $\Delta P = \frac{\rho_f}{203} [u_1^2 \, [1 + 2\phi^2 \, (\frac{2r_t}{r_e} - 1)] + 2u_2^2]$

Where ΔP = cyclone pressure drop, millibars,

ρ_f = gas density, kg/m3,

u_1 = inlet duct velocity, m/s,

u_2 = exit duct velocity, m/s,

r_t = radius of circle to which the centre line of the inlet is tangential, m,

r_e = radius of exit pipe, m,

ϕ = factor from Figure

ψ = parameter in Figure, given by:

$$\psi = f_c \frac{A_s}{A_1}$$

f_c = friction factor, taken as 0.005 for gases,

A_s = surface area of cyclone exposed to the spinning fluid, m2.
For design purposes this can be taken as equal to the surface area of a cylinder with the same diameter as the cyclone and length equal to the total height of the cyclone (barrel plus cone).

A_1 = area of inlet duct, m2.

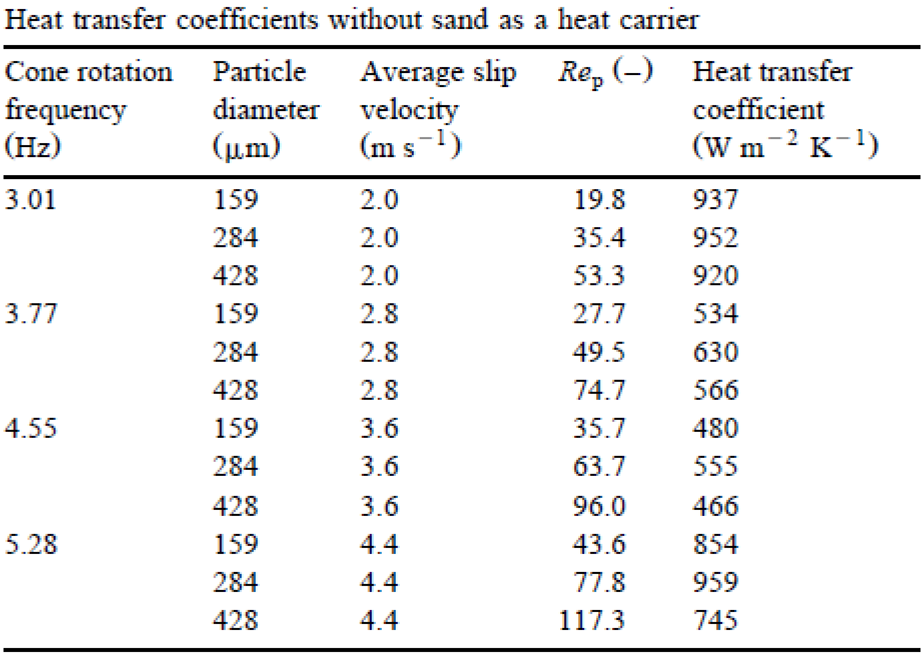
$$\Delta P = \frac{0.44}{203} [69.4^2 \, [1 + 2(0.9)^2 \, (2 * 1.8 - 1)] + 235.4^2]$$

$$\Delta P = 60 \, millibars$$

(Sinnott, 2005)

10.3 REACTOR DESIGN

The figure shows different heat transfer coefficients at different frequencies and we chose the one which is used as follows:



Heat transfer coefficients without sand as a heat carrier

Cone rotation frequency (Hz)	Particle diameter (μm)	Average slip velocity (m s^{-1})	Re_p (–)	Heat transfer coefficient (W m^{-2} K^{-1})
3.01	159	2.0	19.8	937
	284	2.0	35.4	952
	428	2.0	53.3	920
3.77	159	2.8	27.7	534
	284	2.8	49.5	630
	428	2.8	74.7	566
4.55	159	3.6	35.7	480
	284	3.6	63.7	555
	428	3.6	96.0	466
5.28	159	4.4	43.6	854
	284	4.4	77.8	959
	428	4.4	117.3	745

Table 10-1

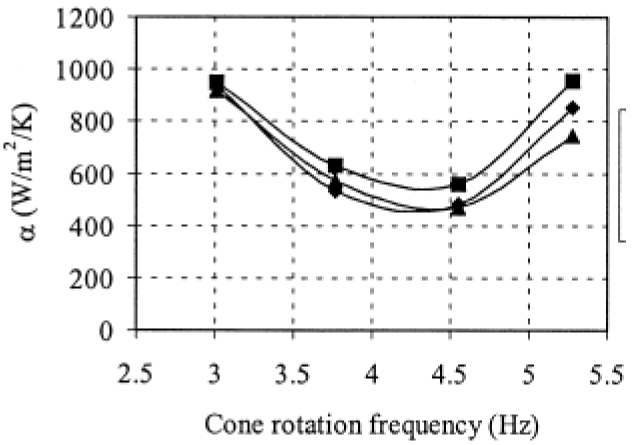


Figure 10-3

Hence we used he Particle velocity of 2 m/s with the maximum Heat Transfer Coefficient since it has an optimum cone frequency and Reynolds number of 35.4

Cone Frequency = 3.01 Hz

Particle Velocity = 2 m/s

Diameter size= Greater than 0.284 mm

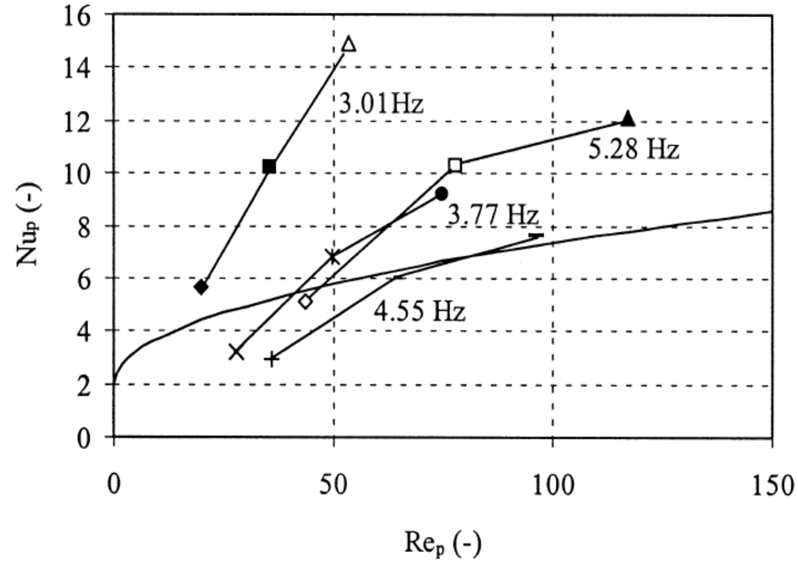


Figure 10-4

Nusselt number calculated from above figure 8-4

Minimum Velocity which occurs at the bottom from where sand is added

V= $\omega\ r$

ω= 2πf

V =2πfr

2 = 2π (3.01) r

r= 10.57 cm

Where r is the radius from where sand enters

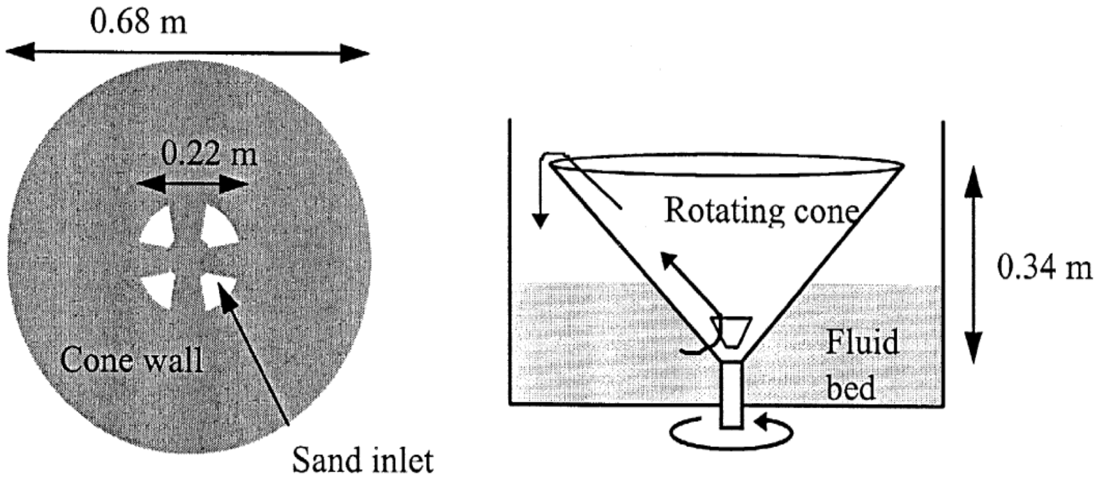


Figure 10-5

Generally the ratio of sand inlet to the cone outer radius = 1:3.08

Hence the outer radius = 32.55 cm

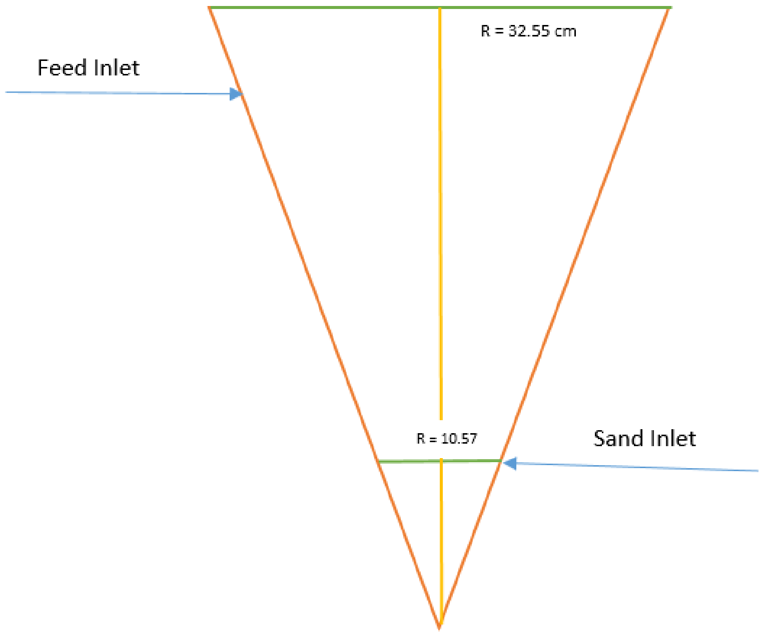


Figure 10-6

Maximum Trajectory Angle = 45°

Hence the height is calculated by

Tan 45° = (32.55)/ x

X= 32.55 cm

Where x is the vertical height

The slant length is calculated by

$$32.55^2 + 32.55^2 = c^2$$

C = 46 cm

Where c is the slant height

Volume = $\frac{1}{3}\,\pi\, r^2 h$

Where r = outside radius

h = vertical height

Volume = $\frac{1}{3}\,\pi\,(32.55^3)(32.55)$

Volume = 36100 cm3 = 36.1 L

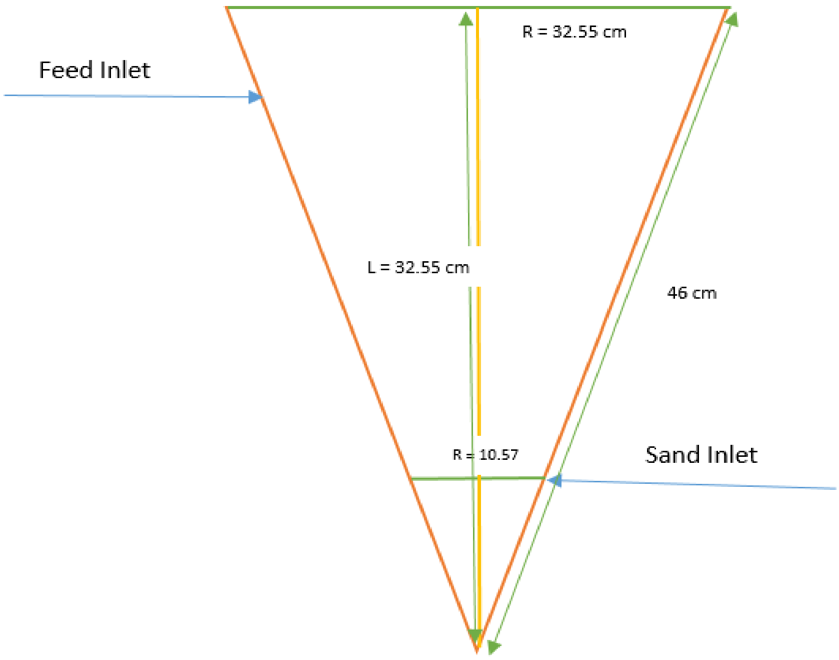


Figure 10-7

Cone Radius (Top)	**32.55 cm**
Cone Radius (Bottom)	10.57 cm
Cone Trajectory	45°
Cone Height	32.55 cm
Cone Slant Length	46 cm
Volume	36.1 liters
Cone Frequency	3.01 Hz

Table 10-2

(Janse, Jong, Prins, & Swaaij, 1999)

10.4 CONDENSER DESIGN

Mass flow rate of bio vapors = 875 kg/hr

Density of bio vapors= 400 kg/m^3

Volumetric Flow rate = 2.185 m^3/hr

Hence the maximum volume that it can accommodate, the minimum time a condenser takes is 15s, hence the volume is [(2.185 x 15)/3600] = 0.145 m^3

Volume =145 L

Length to Diameter ratio of a condenser is 3:1

V= $\frac{\pi\, d^2 l}{4}$

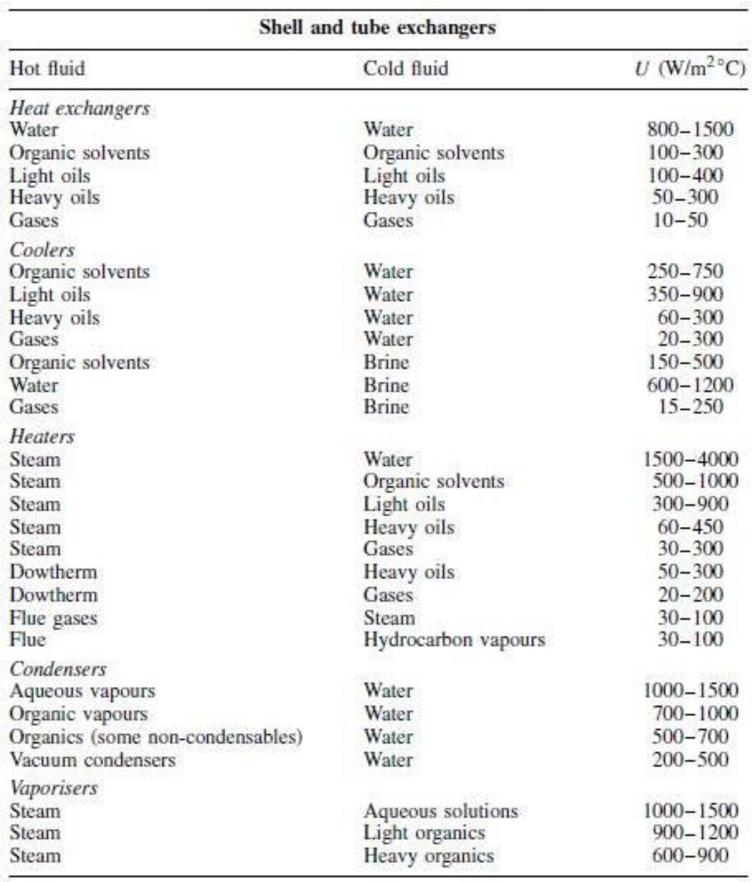
Hence Diameter is 0.395m and length is 1.18m

The enthalpy of vapors = (26.2 + 10) MJ/kg

= 36.2 MJ /kg

Heat transferred from vapor = $\frac{[875 \times (36.2 - 26.2)]}{3600}$

= 2430 kW



Shell and tube exchangers

Hot fluid	Cold fluid	U (W/m^2°C)
Heat exchangers		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
Coolers		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
Heaters		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
Condensers		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
Vaporisers		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

Table 10-3

Assumed overall heat transfer coefficient is shown above = 900 W/m K

Inlet temperature of bio vapors: 550 °C

Outlet Temperatures of bio oil: 50°C

Inlet Temperature of water: 30°C

Outlet Temperature of water: 230°C

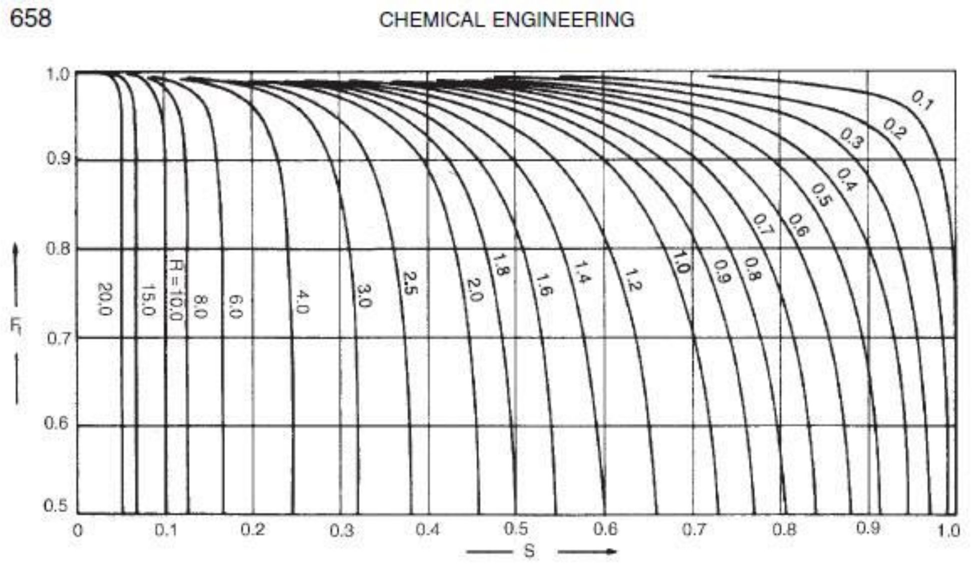


Figure 10-8

R: (550 – 50) / (230 – 30) = 2.5

S: (230 – 30) / (550 -30) = 0.385

From graph: f_r = 0.6

LMTD: $\frac{(550-50)-(230-30)}{\ln\frac{500}{200}}$

LMTD: 327.4

Correction Temperature: 0.6 x 327.4 = 196.4 ℃

Trial Area = [Q/ U (ΔT)]

(2430 x 10^3)/(900 x 196.44)

A = 13.74 m^2

Tubes of inner dia 16.8 mm, outer dia 20 mm

Surface area of one tube: $\pi\ 0.02^2$(1.18)

= 0.74 m^2

Number of Tubes: (13.74/0.74)

= 19 tubes

Square pitch: 1.25 x O.D

1.25 x 20

= 25 mm

10.4.1 Shell Side Coefficient

Use pull-through floating head, no need for close clearance.

Select baffle spacing = shell diameter, 45 per cent cut.

Clearance = 90 mm.

Shell internal diameter: 118.5 + 90 =208.5 cm= 2085 mm

Use Kern's method to make an approximate estimate.

Cross-flow area A = $[[(25 - 20)/25] \times 2085 \times 2085 \times 10^{-6}]$

$= 0.869\ m^2$

Condensing Coefficient: 1500 W/m^2℃

Shell side mean temperature: 130 ℃

Tube side mean temperature: 300 ℃

Wall temperature:

$(130 - Tw) \times 1500 = (300 - 130) \times 900$

Tw = 28 ℃

Mean Temperature: (130 + 28) /2 = 79℃

At 79℃

Density: 1150 kg/m^3

Viscosity: 0.0022 Ns/m^2

Thermal Conductivity: 0.15 W/m℃

Average molecular weight: Combination of butane, propane, propene, ethane, ethene

M_r= 39.8999= 40

Vapor Density: (M_r /22.4) x [273/(273 + Shell side mean temperature)] x ($\Delta T_{cooling\ water}$ / passes)

(40/22.4) x [273/ (273 + 130)] x (200/4)

= 60.5 kg/m^3

Γ = (Mass of bio vapors/ (length of tube x number of tubes)

= (875/3600) x [1/(1.18 x 19)]

= 0.011 kg/s

h_c= 0.95 Kc $[[p_l\ (p_l - p_v)\ g] / (\Gamma\ v)]^{0.33}$

h_c = 0.95x (0.15)x [[1150 x (1150 – 60.5) x 9.81] / (0.011 x 0.0022)] $^{0.33}$

h_c = 1173 W/m℃

10.4.2 Tube Side Coefficient

Cross Sectional Area

$A = \frac{\pi r^2}{4}$ x (Number of tubes / Passes)

$A = \frac{\pi 0.0168^2}{4}$ x (19 /4)

A = 1.05 mm^2

Tube Velocity = (0.1886/1000) x (1/ 0.00105)

V = 0.18 m/s

Nu = 0.023 x $(Re)^{0.8}$x $Pr^{0.3}$

hd/k = 0.023 x $(\frac{cu}{k})^{0.3}$ x $(pvd/u)^{0.8}$

h x (16.8 x 10^{-3}) /(0.15) = 0.023 x (2.01 x 0.0022 / 0.15 $)^{0.3}$ x (60.5 x 0.18 x 0.0168 / 0.022$)^{0.8}$

h_o= 5200 W/m℃

$$\frac{1}{U} = \left(\frac{1}{hi}\right) + (\frac{Do}{Di})(\frac{1}{ho})$$

$$\frac{1}{U} = \left(\frac{1}{1173}\right) + (\frac{20}{16.8})(\frac{1}{5200})$$

U= 924.5 W/m℃

Since 924.5 is closer to the assumed value (900) so it's the correct heat transfer coefficient assumed

Effective pressure drop

Viscosity of bio oil = 0.022 kg/ms

Re= (ρud/μ)

(0.18 x 60.5 x 0.0168)/ 0.022

Re= 8.316

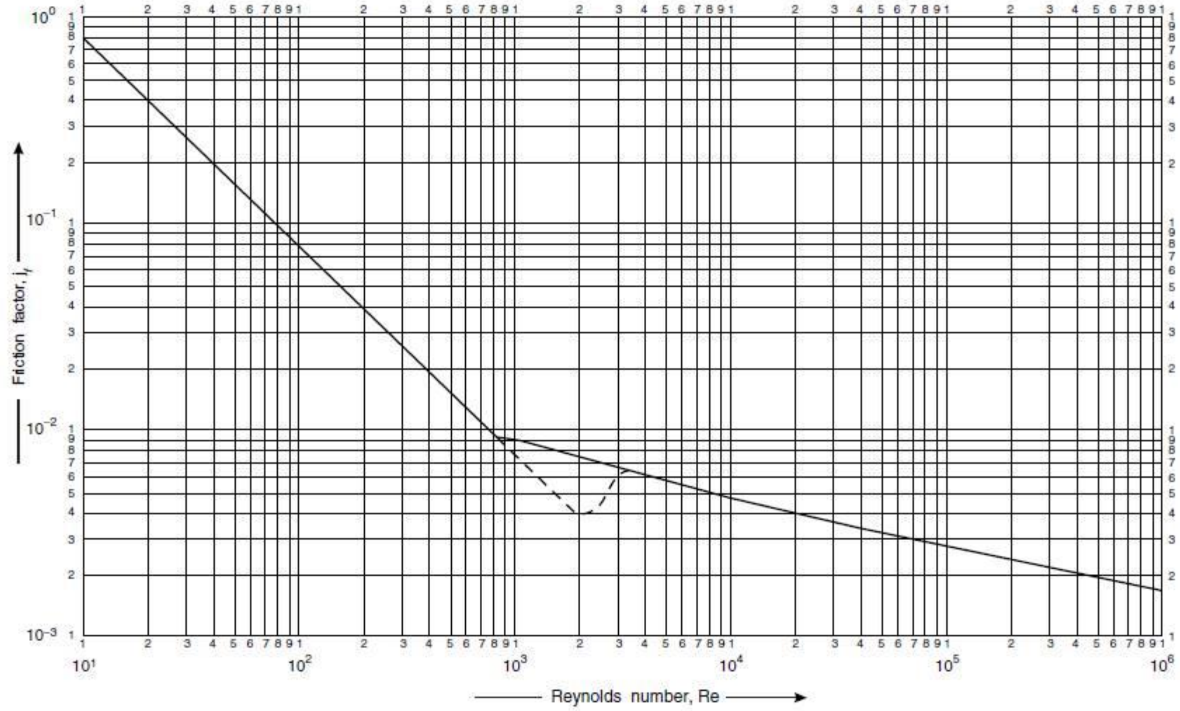


Figure 10-9

From Figure 12.24, j_f= 0.091

Neglect Viscosity correction

$\Delta P=4\ [8 \times j_f \times (\frac{L}{(I.D(tube)}) +2.5] \times (\rho\times v^2)/2$

$\Delta P=4\ [8 \times 0.091 \times (1.19/0.0168) + 2.5] \times (60.5 \text{ x } 0.18^2)/2$

ΔP = 211.9 Pa = 0.031 psia

Hence the pressure drop is negligible in the tube side which makes it an effective design.

11 COSTING AND ECONOMICS

All costing has been done with the help from Coulson Richardson vol 6 unless specified otherwise.

11.1 EQUIPMENT COSTING

11.1.1 Combustor

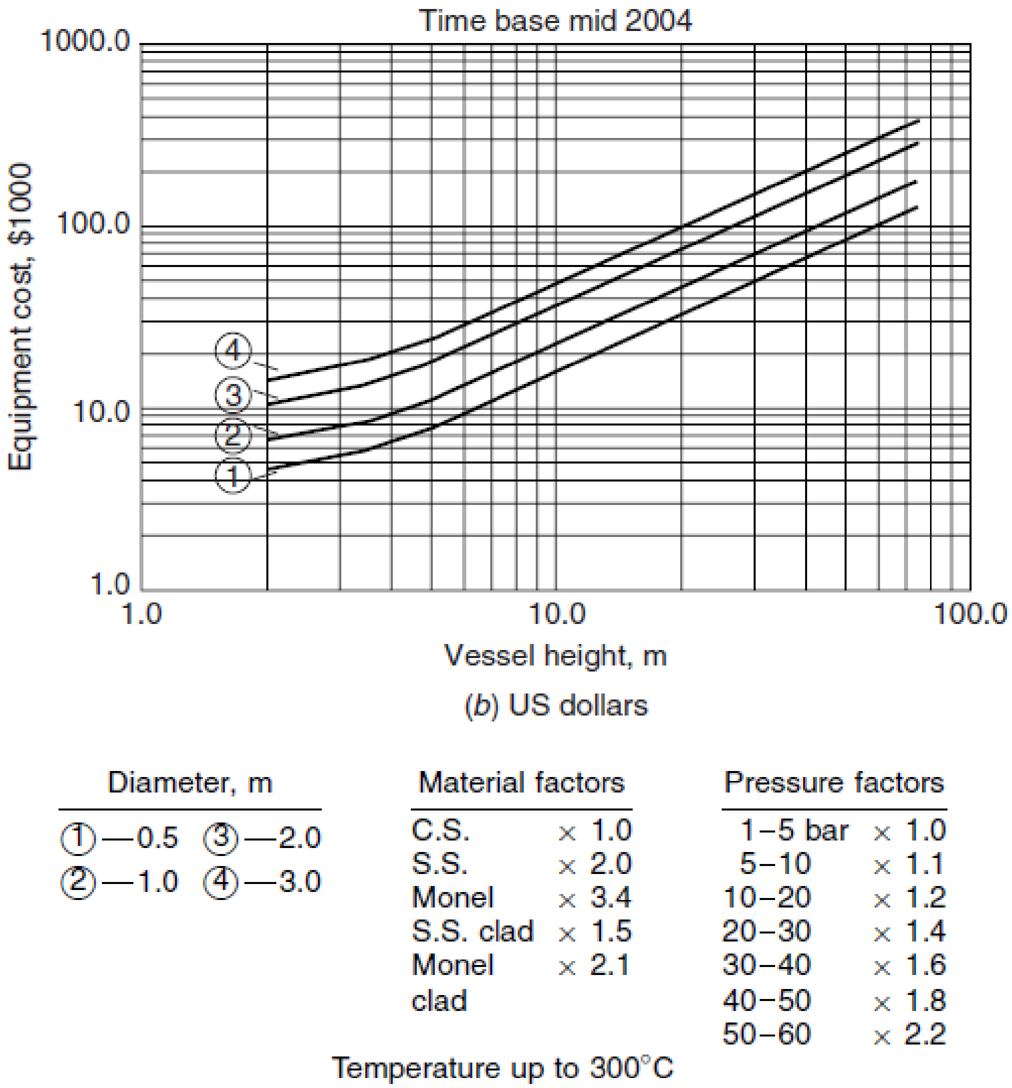


Figure 11-1

Purchase Cost = Bare cost from figure x Material factor x Pressure factor

Bare cost = $4000

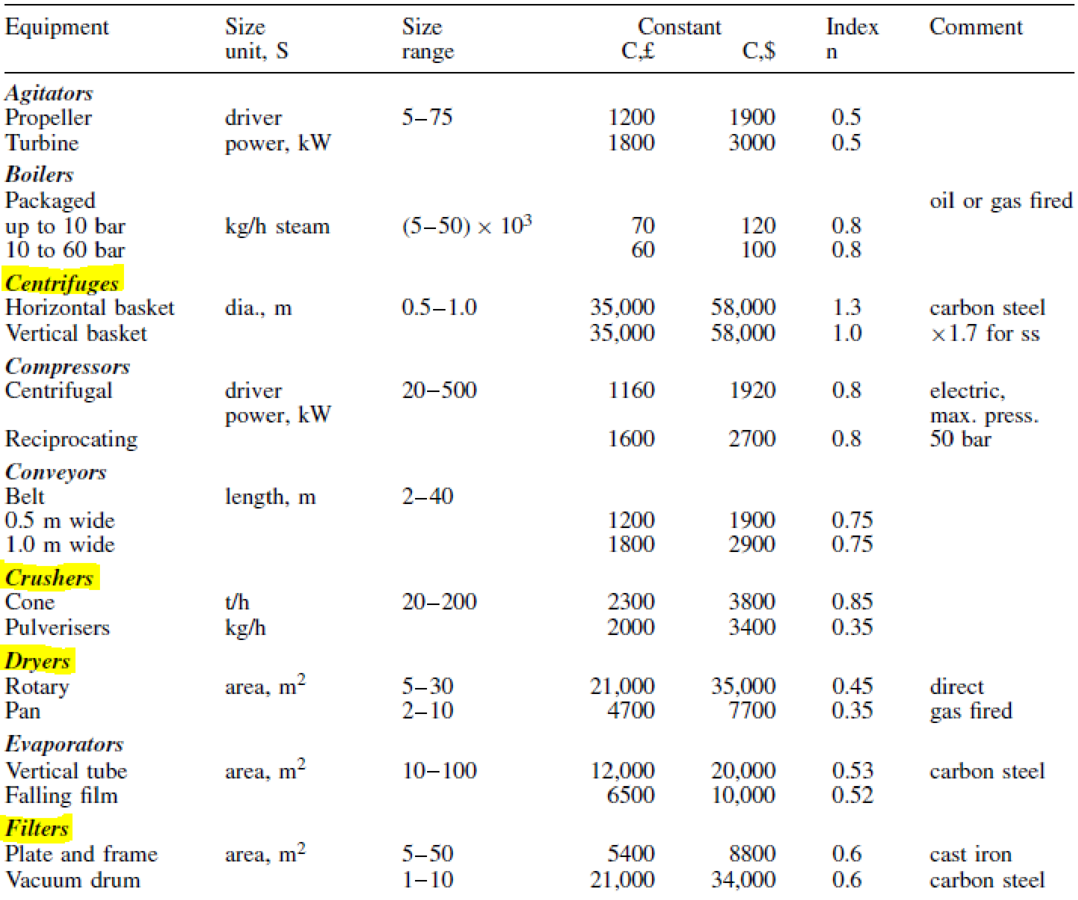
Material factor = 2.0 for S.S

Pressure factor = 1.0

Purchase Cost = 4000*2*1

Purchase Cost = $8000

11.1.2 Miscellaneous Equipment



Equipment	Size unit, S	Size range	Constant C,£	C,$	Index n	Comment
Agitators						
Propeller	driver	5–75	1200	1900	0.5	
Turbine	power, kW		1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5–50) \times 10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	×1.7 for ss
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press.
Reciprocating			1600	2700	0.8	50 bar
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel

Table 11-1

11.1.2.1 Cyclone Separator

$C_e = CS^n$

C_e = Purchase Cost of equipment

S = Characteristic size parameter = 0.2m

C = cost constant = \$58000

n = index for that type of equipment = 1.0

C_e = 58000*0.2$^{(1.0)}$

C_e = \$11,600

11.1.2.2 Dryer

$C_e = CS^n$

C = \$7700

S = 10m

n = 0.45

C_e = 7700*10$^{(0.45)}$

C_e = \$21,700

11.1.2.3 Filters

Filter size = 5m^2/each

Final Mesh = 3mm

$C_e = CS^n$

C = \$8800

S = 2 x (5m)

n = 0.60

C_e = (8800*5$^{(0.60)}$)*2

C_e = \$46,200

11.1.2.4 Crusher

Capacity = 200 kg/hr

$C_e = CS^n$

C = \$3400

S = 200 kg/hr

n = 0.35

$C_e = 3400*200^{(0.35)}$

C_e = \$36,800

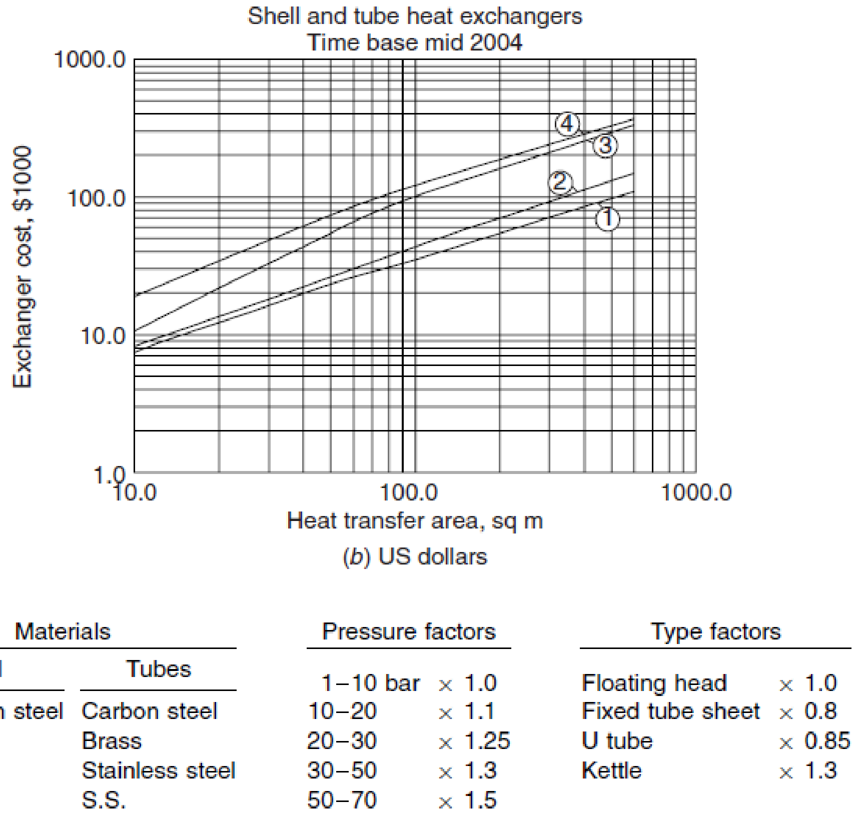
11.1.3 Reactor

Purchase Cost = \$17000 (Estimated from data provided by University of Twente)

11.1.4 Washer

Industrial Washer with required Capacity = \$30,000 (Ali express)

11.1.5 Condenser



Materials	
Shell	Tubes
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 11-2

Purchase Cost = Bare cost from figure x Type factor x Pressure factor

30 tubes, Triple pass. Area = $11 m^2$

Bare cost = $20,000

Type factor = 0.85

Pressure factor = 1.0

Purchase Cost = 20000*0.85*1

Purchase Cost = $17,000

Total purchase cost of equipment (PCE)

Combustor = 8000

Cyclone Separator = 11,600

Dryer = 21,700

Reactor = 17,000

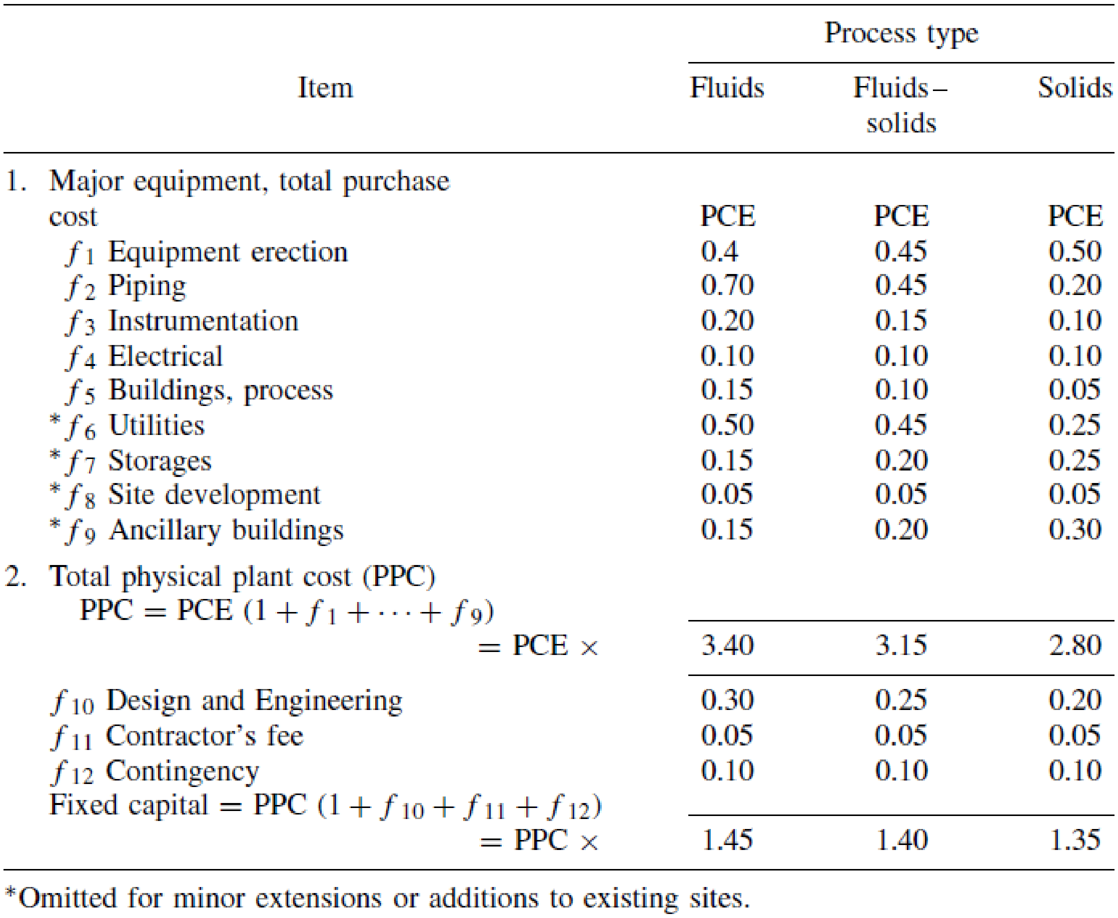
Filters = 46,200

Crusher = 36,800

Washer = 30,000

Condenser = 17,000

Purchase Cost of Equipment = $188,300



Item	Process type		
	Fluids	Fluids–solids	Solids
1. Major equipment, total purchase cost	PCE	PCE	PCE
f_1 Equipment erection	0.4	0.45	0.50
f_2 Piping	0.70	0.45	0.20
f_3 Instrumentation	0.20	0.15	0.10
f_4 Electrical	0.10	0.10	0.10
f_5 Buildings, process	0.15	0.10	0.05
*f_6 Utilities	0.50	0.45	0.25
*f_7 Storages	0.15	0.20	0.25
*f_8 Site development	0.05	0.05	0.05
*f_9 Ancillary buildings	0.15	0.20	0.30
2. Total physical plant cost (PPC) $\text{PPC} = \text{PCE}(1 + f_1 + \cdots + f_9)$			
$= \text{PCE} \times$	3.40	3.15	2.80
f_{10} Design and Engineering	0.30	0.25	0.20
f_{11} Contractor's fee	0.05	0.05	0.05
f_{12} Contingency	0.10	0.10	0.10
Fixed capital $= \text{PPC}(1 + f_{10} + f_{11} + f_{12})$			
$= \text{PPC} \times$	1.45	1.40	1.35

*Omitted for minor extensions or additions to existing sites.

Table 11-2

11.2 PHYSICAL PLANT COST

For estimating Physical Plant Cost, following factors are to be considered

Equipment erection = 0.45

Piping = 0.45

Instrumentation = 0.15

Electrical = 0.1

Utilities = 0.45

Storage = 0.2

Site development = 0.05

Total physical plant cost (PPC) = 188,300(1+0.45+0.45+0.15+0.1+0.45+0.2+0.05) = $536,655

PPC = $536,655

11.3 FIXED COST

Design and Engineering = 0.25

Contingencies = 0.1

Fixed Capital = 536,655(1+0.25+0.1) = $724,485

FC = $724,485

Operating time = 16hrs/day x 365 days = 5840 hours

11.4 VARIABLE COST

Raw Material = $22/ton * 1.25 tons/hr * 5840 hours = $240,900

Miscellaneous Cost = 0.1*Maintenance cost = $3622

Utilities = 2990 tons water x 1¢/ton = $30

Total Variable Cost = 240,900 + 3622 + 30 = $244,552

Variable Cost = $244,588

11.5 FIXED COSTS

Maintenance = 5% of FC = 536,655*0.05 = $36,244

Operating labor = $140,160

Plant overhead = 50% of operating labor = 140,160*0.50 = $70,080

Laboratory = 30% of operating labor = 140,160*0.30 = $42,048

Capital charges = 6% of FC = 536,655*0.06 = $43,469

Insurance = 1% of FC = 536,655*0.01 = $7245

Total fixed cost = 36,224+140,160+70,080+42,048+43,469+7,245 = $339,226

Fixed Cost = $339,226

Direct Production Cost = Fixed Cost + Variable Cost = 339,226 + 244,552

Direct Production Cost = $583,778

11.6 PAYBACK PERIOD AND RATE OF RETURN

Bio oil sold at = $0.18/kg

Bio oil production = 4,380,000 kg/year

Sales = 0.18*4,380,000 = $788,900

RoR = $\frac{\boldsymbol{Sales - Direct\ Production\ Cost}}{\boldsymbol{Fixed\ Capital}}$ **=** $\frac{\mathbf{788{,}900 - 583{,}778}}{\mathbf{536{,}655}}$ **= 0.27 = 27%**

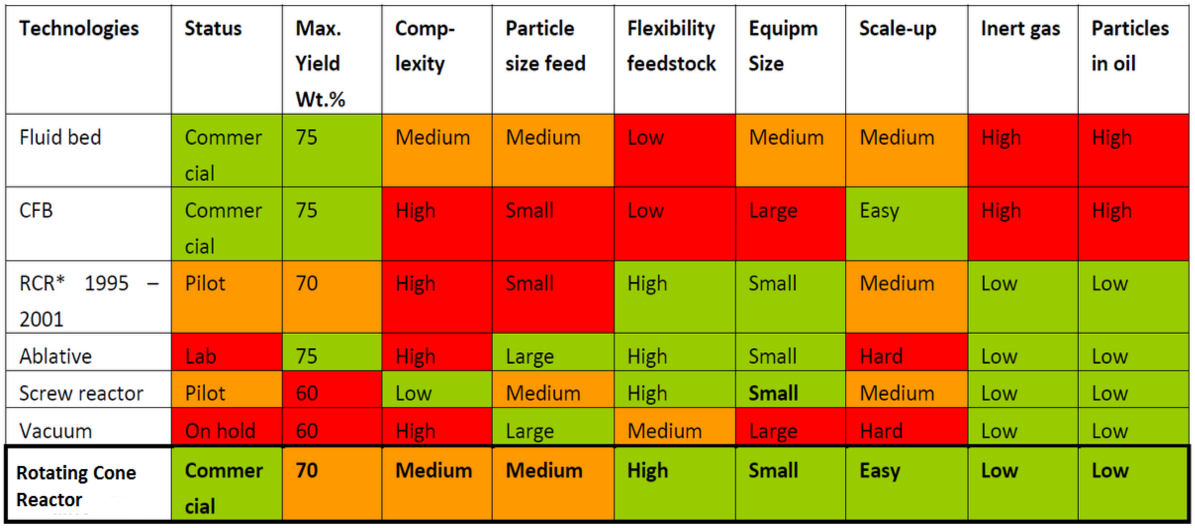
Payback period = $\frac{\mathbf{27}}{\mathbf{100}} = \mathbf{3.7}\ \boldsymbol{years}$

12 CONCLUSION

- According to US Agricultural Department Pakistan produces a total of 69 million tons of agricultural waste that can be utilized to answer some of our energy problems
- This process is capable of displacing 33% of petroleum globally according to Rafael Hernandez in the journal Current Opinion in Chemical Engineering 2011.(Hernandez, 2011)
- Normally the bio oil produced from pyrolysis has a High Heating Value of 16-19 MJ/kg. Our bio oil theoretically has a HHV of 26 MJ/kg compare to that of crude oil diesel that is 44 MJ/kg.
- Our pyrolysis process can convert up to 70 wt. % of the biomass feedstock into bio-oil and the remaining part into char and gas.
- Our proposed pyrolysis technology is characterized by an intense mixing without the need for an inert carrier gas. The improved RCR (Rotating Cone Reactor) design results in a remarkably small reactor, reduced system complexity and minimum downstream equipment size compared to competing pyrolysis technologies.(Kyritsis, 2000)

  The main advantages of using this method are:

  - High biomass throughput per reactor volume resulting in compact reactor design.
  - Absence of inert carrier gas resulting in minimum downstream equipment size.
  - Maximum caloric value of pyrolysis gas.
  - Very simple process: no gas recycle required.
  - Straightforward to scale-up.
  - Able to produce electricity or to produce a combination of electricity and steam for other applications.
  - High flexibility for feedstock: waste material, large particle size, etc.
  - Low amounts of solids in the oil (down to 0.01 %wt.).



Technologies	**Status**	**Max. Yield Wt.%**	**Comp-lexity**	**Particle size feed**	**Flexibility feedstock**	**Equipm Size**	**Scale-up**	**Inert gas**	**Particles in oil**
Fluid bed	Commercial	75	Medium	Medium	Low	Medium	Medium	High	High
CFB	Commercial	75	High	Small	Low	Large	Easy	High	High
RCR* 1995 – 2001	Pilot	70	High	Small	High	Small	Medium	Low	Low
Ablative	Lab	75	High	Large	High	Small	Hard	Low	Low
Screw reactor	Pilot	60	Low	Medium	High	**Small**	Medium	Low	Low
Vacuum	On hold	60	High	Large	Medium	Large	Hard	Low	Low
Rotating Cone Reactor	**Commercial**	**70**	**Medium**	**Medium**	**High**	**Small**	**Easy**	**Low**	**Low**

Table 12-1

The compact design of our modified rotating cone reactor makes scaling-up to capacities larger than 5 t/h straightforward. Because of the simplicity of the rotating cone process, investment costs can be considerably lower in comparison to other pyrolysis technologies. CFB and fluid bed systems are more capital intensive, also because of the larger down-stream equipment, including ATEX and other safety issues.

Our design includes recovery of excess heat in the form of steam which can be used for industrial or local heating applications and electricity production. Theoretically more electricity can be produced than required for the whole plant, therefore making it self-sustaining.

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