DESIGN OF CIRCULATING FLUIDISED BED GASIFIER FOR SUGARCANE BAGASSE

By

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DESIGN OF CIRCULATING FLUIDISED BED GASIFIER FOR SUGARCANE BAGASSE

A thesis submitted in partial fulfillment of the requirements for the Degree of

Master of Technology

(Process Design Engineering)

By

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TO WHOM IT MAY CONCERN

This is to certify that Mr. Gnasasambandan S. s/o Sh. Swaminathan. C, Enrollment No. R67020900 - College of Engineering, University of Petroleum & Energy Studies, Dehradun has successfully undergone the training w.e.f. Day 03 of January'11 to Day 5 of May'11 in Process Department.

He did a project on Design of Circulating Fluidised Bed Gasifier for Sugarcane Bagasse.

During this tenure of training we found him very punctual, hardworking, sincere and dedicated towards the assignments allocated to him.

We wish him good luck and success for his bright future.

For Spray Engineering Devices Limited.

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Abstract

Bagasse is a residual biomass that is available in plenty from sugar industries. It is usually combusted to generate electricity. However combustion efficiency is poor as bagasse is powdery further combustion gives rise to particulate emission. Gasification of bagasse generates clean producer gas, which can be used in gas engines or gas turbines to generate electricity and thereby improve the electrical efficiency.

In this work a circulating fluidized bed gasifier(CFBG) for sugarcane bagasse has been designed. An equilibrium model was used to predict the product composition. The product gas estimated by the model has a heating value of 13.29 MJ/Nm^3. A gasification efficiency of 81% was calculated.

The design of the gasifier was based on hydrodynamics and a pressure balance around the system. The CFBG is an indirect heating gasifier where a medium transports heat between combustion and gasification section. Sand was used as the heat carrier and the circulation rate of sand was determined as to be 1620 kg/hr. The gasifier section is maintained at bubbling fluidized condition using steam and the riser section is maintained in fast fluidisation regime using air. The diameter of gasifier was determined as 0.38 m, bed height as 1m and freeboard height as 2 m for gasifying 100 kg/hr of bagasse using 30 kg/hr of steam. The riser diameter is 0.15 m and its height is 2 m.

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Dedicated to my parents.

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Nomenclature

English

a	decay constant of the riser, m^{-1}
A	area m^2
Ar	Archimedes number
B/T	ratio of bottom air to total air into riser
d_P	particle diameter, m
D	diameter, m
G_S	solids circulation rate, related to top of the riser, kg $m^{-2}s^{-1}$
g	gravitational acceleration, m s^{-2}
h	height, m
m	mass, kg
p	pressure, Pa
Q	volume flow, m^3s^{-1}
Re	Reynolds number
S/T	ratio of secondary air to total air into riser
u	velocity, m s^{-1}
u_T	terminal velocity, m s^{-1}
U	fluidisation velocity ms^{-1}
f_s	friction factor, dimensionless
K_∞	elutriation rate constant, kg $m^{-2}s^{-1}$
r	radius, m

Greek letters

ϵ	voidage, dimensionless
ρ	density, kg m^{-3}
δ_b	bubble fraction, dimensionless
Δp	pressure drop, Pa
μ	viscosity kg m$^{-1}s^{-1}$
ϕ	sphericity
Δu	velocity difference, m s^{-1}

Subscripts

a	annulus
b	bubble phase
Con	connection between gasification and combustion zone
Horizontal	horizontal section in seal loop
cyc	cyclone
avg	average
SLsta	seal loop standpipe
SLdow	seal loop downcomer
SLhor	seal loop horizontal
gas	gasifier
hyd	hydrostatic pressure drop
fr	fricitional pressure drop
s	solids
g	gas
mf	minimum fluidisation condition

Acronyms

FICFB	fast internal circulating fluidized bed
CFBG	circulating fluidized bed gasifier
BFB	bubbling fluidized bed
CFB	circulating fluidized bed
GE	gas engine
PFD	process flow diagram

Chapter 1

Introduction

1.1 Introduction

The use of biomass energy is today an urgent need in order to tide over the phenomena of 'peak oil'. Biomass is a carbon neutral source of energy. The stored chemical energy of the biomass can be converted into a usable form of energy such as electricity or as a fuel. This can be achieved using a number of different routes, each with specific advantages and disadvantages. In the recent times much emphasis is laid on research for generation of sustainable and green energy form to replace conventional fossil fuels. Bio-energy is seen as one of the key options to mitigate GHG emissions and substitute fossil fuels. The use of biofuels for transportation and as an alternative to fossil fuels helps in providing environment friendly solutions for the current energy crisis that is faced worldwide. Gasification of biomass helps to reduce the dependency on fossil fuels for generation of thermal and electrical power. Biomass gasification has both economic and ecological benefit and proves as an promising renewable source of energy.

Thermochemical biomass conversion includes a number of different possible routes for the production of useful fuels and chemicals from the initial biomass. The basis of thermochemical conversion is the pyrolysis process, which include all chemical changes occurring when heat is applied to a material in the absence of oxygen. The products of biomass pyrolysis include water, char, oils or tars, and permanent gases including methane, hydrogen, carbon monoxide, and carbon dioxide. Gasification is a complex thermal process that depends on the pyrolysis mechanism to generate gaseous products, which in the presence of reactive gases such as oxygen and steam convert the majority of the biomass into a fuel gas. The fuel gas product obtained by the gasification mechanism can in turn be further purified to syngas, which is mainly composed of carbon monoxide and hydrogen. Syngas may be used to produce chemicals and liquid fuels over catalysts. However, a majority of fuel gas is used directly to fire processes such as kilns, as fuel in steam boilers, and as a gaseous fuel in internal combustion engines (ICEs) and gas turbines thereby replacing conventional fossil fuels.

India has just over 500 sugar mills, with nine states (Uttar Pradesh, Bihar, Punjab and Haryana in the northern region; Maharashtra and Gujarat in the western region and Andhra Pradesh, Tamil Nadu and Karnataka in the southern region) holding 95 percent of them. Most mills are either privately owned or co-operatives. Bagasse which is a waste produced from the sugar industry has been used for cogeneration. However only 1/3rd of bagasse is used for cogeneration and the remaining is wasted or incinerated.

The producer gas is affected by various processes and hence one can expect variations in the gas composition produced from various biomass sources. The gas composition is also a function of gasifier design and thus, the same fuel may give different calorific value as when used in two different gasifiers. The major advantage of fluidized bed gasifier over downdraft is its flexibility with regard to its feed rate and composition. Fluidized Bed systems can also have high volumetric capacity and the temperature can be easily controlled.

Country	Area Harvested	Production Ranking	Yield	Production
Australia	423,000	8	85.13	36,012,000
Brazil	5,303,560	1	73.83	386,232,000
China	1,328,000	3	70.71	93,900,000
Colombia	435,000	7	84.14	36,600,000
Cuba	1,041,200	9	33.33	34,700,000
India	4,300,000	2	67.44	290,000,000
Mexico	639,061	6	70.61	45,126,500
Pakistan	1,086,000	5	47.93	52,055,800
Philippines	385,000	11	67.10	25,835,000
Thailand	970,000	4	76.36	74,071,952
USA	403,390	10	77.29	31,178,130
Other	4,091,132			244,581,738
Total	20,405,343			1,350,293,120
Average			68.53	

Table 1.1: Top 11 Sugar Growing Countries

1.1.1 Global Sugarcane Processing Industry

Sugarcane is currently grown under a wide range of conditions, in tropical and subtropical regions between 35°N in Spain to 35°S in South Africa. In absence of irrigation, a good distribution of rainfall is required as water requirements for the crop are 1.2-1.6 m/year. Depending on crop variety sugarcane harvesting generally occurs every 9-14 months.

A WADE report (2004) reported that the three largest sugarcane growers in terms of production are Brazil, India and China, which accounts for more than half of total sugar production. Table 1.1 (on Page 3) compares the production and yield figures for the top 11 sugar-growing countries.

Bagasse is the fibrous residue of cane stalk obtained after crushing and the extraction of juice. Pickering, S. (2000) reported that each tonne of sugarcane can yield 250kg of bagasse. The composition of bagasse varies with variety and maturity of sugarcane as well as with harvesting methods used and efficiency of the sugar mill in processing the sugarcane.

1.1.2 Potential of Bagasse

The bagasse potential availability in the world is 337357 thousand tonnes at 50 percent humidity and 168679 thousand tonnes on a dry basis based on the report published by International Sugar Organization (2005).The Sugar Technologists Association of India (2002) states that Indian sugar mills are currently self-sufficient in terms of energy, already using bagasse to meet their steam and power requirements. However only 20-30 percent of all bagasse is used for cogeneration purposes. This suggests that the remaining 2/3 of bagasse is currently being wasted, as it is being incinerated for disposal purposes rather than energy recovery. Hence there is enormous potential for energy recovery from sugarcane bagasse.

1.2 Gasification

Gasification of biomass gives carbon monoxide, carbon dioxide, hydrogen and methane. The composition of the product gas varies according to feedstock, gasification temperature and gasifying medium.Biomass are those substances which contain cellulose, hemicellulose (which are long chain sugars) besides lignin which gives strength to biomass. On heating hemicellulose volatilizes first at temperature around 200^o C followed by cellulose which volatilizes at slightly higher temperature followed by pyrolysis of lignin at around 350^o C. In gasifier owing to short supply of oxygen and high temperature the products of volatile oxidation are reduced to carbon monoxide and hydrogen. Gasification involves the following steps

- Drying
- Pyrolysis or thermal decomposition
- Partial combustion of some gases, vapours, and char
- Gasification of decomposition products

1.2.1 Drying

The bagasse which is fed to the gasifier contains 15 percent of moisture. The moisture content of bagasse is reduced from 50 percent to 15 percent in the dryer and the remaining 15 percent moisture which is present in the bagasse first volatilises as the bagasse enters the bed of hot sand in the gasifier. This is the first step that takes place in the gasifier.

1.2.2 Pyrolysis or Thermal Decomposition

Based upon the heating rate and final temperature, pyrolysis can be of three types,

- Slow pyrolysis
- Fast pyrolysis
- Rapid pyrolysis

Slow pyrolysis has low final temperature and slow heating rate under such conditions the product yields more of oil and char. Fast pyrolysis has medium heating rate and high final temperature, the yield of product mainly contains more of bio oil and small amount of bio gas and char. Whereas in rapid pyrolysis the heating rate is in the range of 400-600 C/second and the final temperature is in the range of 850 - 1000 C, under such conditions the product mainly consists of gases and small amount of char. In gasification rapid pyrolysis is desirable owing to high yeild of gases.

The pyrolysis step in gasification can be represented by the following reaction.

Bagasse + heat → vapours + char

1.2.3 Partial combustion of some gases, vapours and char

Gasification reactions are endothermic in nature. The oxygen present in bagasse is released during the pyrolysis step in gasification this oxygen is utilized for combustion of some of the gases, vapours and char. Combustion is an undesirable action because it releases energy whereas gasification packs up energy in the product but a very small part of gases or vapours are still combusted to give heat for the gasification reaction.

1.2.4 Gasification of Decomposition products

The char which is obtained from the pyrolysis step is gasified using a gasifying medium like steam, air or oxygen to form carbon monoxide and hydrogen.Gasification process adds hydrogen to and strips carbon away from the feedstock to produce gases with higher hydrogen to carbon ratio. The reaction can be represented as

Char + gasifying medium → CO + H_2 (gases).

The above shown reaction is an overall representation of the gasification reaction but there are various reactions which actually take place during gasification. Based on the extent of the reaction the product yield varies. The gasification reactions are shown in section 1.3

1.3 Gasification reaction

The following are the typical gasification reaction taking place in the reactor. The extent of reaction and yield of products varies according to the reactions which are pronounced in that range of reactor temperature. In this work product composition is found assuming equilibrium model. This model gives the value of maximum yield of products in the reactor temperature range which is a good initial value. Later the reactor parameters will be optimized to get the maximum yield.

The main reactions are

- Boudouard reaction

$$C + CO_2 \leftrightarrow 2CO$$

- Water gas reaction

$$C + H_2O \leftrightarrow CO + H_2$$

- Hydro gasification

$$C + 2H_2 \leftrightarrow CH_4$$

- Shift reaction

$$CO + H_2O \leftrightarrow CO_2 + H_2$$

- Methanation reaction

$$2CO + 2H_2 \rightarrow CH_4 + CO_2$$

$$CO + 3H_2 \leftrightarrow CH_4 + H_2O$$

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$$

- **Steam reforming reaction**

$$CH_4 + H_2O \leftrightarrow CO + 3H_2$$

$$CH_4 + 0.5O_2 \rightarrow CO + 2H_2$$

1.4 Combustion Reaction

The various combustion reactions taking place in the reactor are

$$C + O_2 \leftrightarrow CO_2$$

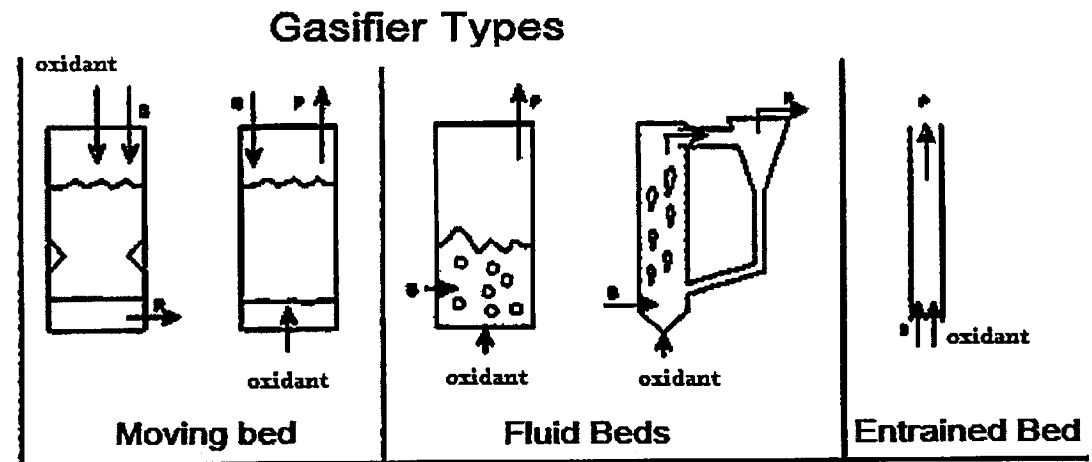
$$CO + 0.5O_2 \rightarrow CO_2$$

$$CH_4 + 2O_2 \rightarrow CO_2 + H_2O$$

$$H_2 + 0.5O_2 \rightarrow H_2O$$

1.5 Gasifier Reactor Types

The reactor system used throughout the world to gasify biomass are broadly classified as (i) Direct heating system and (ii)Indirect heating system. In direct heating system the heat required for the gasification process is supplied externally either by electrical heating or by burning some fuels like charcoal or by heating the gasifying medium, whereas in indirect heating method the heat required for gasification is supplied by burning the char formed as a result of gasification in a separate chamber and transferring this heat through a heat carrier to the gasification chamber. This method reduces the supply of energy from external source thereby increases the cycle efficiency of the system.



	Moving bed		Fluid Beds		Entrained Bed
	Co-Current	Counter Current	Dense	Circulating	
T^0C	700-1200	700-900	< 900	< 900	1500
tars control	low easy	Very high very easy	Intermediate Intermediate	Intermediate Intermediate	absent very complex
scale	< 5MW	< 20 MW	10 < MW < 100	20 < MW < ?	> 100 MW
feedstock	very critical	critical	Less critical	Less critical	very fine particles

Figure 1.1: Gasifier Types

In indirect heating system combustion and gasification takes place in separate chambers and thereby prevent dilution of product gas with nitrogen that enter with air required for combustion. Using steam as gasifying agent improve the heating value of the product gas thereby increasing the gasification efficiency.

1.5.1 Direct Heating System

Direct heating reactors used throughout the world are dominated by four configurations which are counter current moving bed, co current moving bed, entrained bed, and fluidised bed reactors. The moving bed or fixed bed reactors consists of a bed of biomass sitting on top of a moving grate and the oxidant is fed into the reactor from above or below. In this type char is burnt in the same reactor to give the heat of gasification.

Counter Current Moving Bed Gasifier

In this type of gasifiers the fuel is feed from the top and the oxidant is fed from the bottom. The oxidant is air or oxygen. The fuel gets dried up first, as it enters the reactor, then devolatalises and then the char formed is burnt to give heat of gasification. In this type of reactors the product gas does not reach high temperatures. The remaining ash material is removed from the bottom. The heat from product gas from devolatilisation is used for drying. Biomass is still wet before it enters the oxidation zone as it is dried by product gas at low temperature and also the product gas contains more of particulates. The main disadvantages with this type of reactors are, the tar in the product gas are not cracked as they don't reach high temperature, problem with gas cleaning, the heating value of the product gas is only around 4-6 MJ/Nm^3 when air is used as oxidant, product gas is diluted with nitrogen. There are chance of hot spots in the reactor which gives wide variation in product composition.

Co Current Moving Bed Gasifier

In this type of reactors fuel is fed from the top and the fuel moves downs as new fuel is added from the top. The oxidant is given from the top or at the mid way. The fuel is dried before it enters the oxidation zone through radiation and conduction mechanism. The temperature of the product gas reaches about 1200°C which helps in tar cracking, but still there is a small amount of tar remaining in the product gas. The product gas from this type of gasifiers have heating value of around 4-6

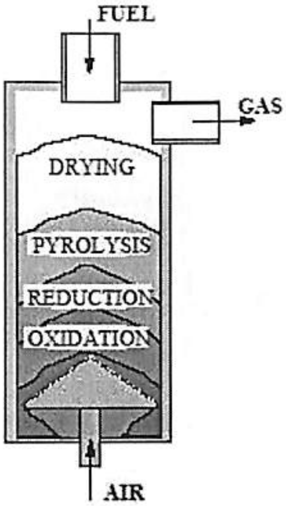


Figure 1.2: Counter current moving bed reactor

Reactor Type	Gas Comp Vol %					HHV(MJ/m^3)
	H_2	CO	CO_2	CH_4	N_2	
Fluidbed,air blown	9	14	20	7	50	5.4
Updraught	11	24	9	3	53	5.5
Downdraught,air	17	21	13	1	48	5.7
Downdraught,oxygen	32	48	15	23	0	10.4
FICFB	40	25	20	10	5	13

Table 1.2: Comparison of different types of gasifier Ref: Bridgewater(1995)

MJ/Nm^3. The disadvantages of this type of reactors are, the product gas contains large amount of particulates, though amount of tar of formed is less compared to updraft, the heating value of product gas is also less compared to that of a CFB gasifier.

Fluidised Bed Gasifier

Traditional circulating fluidised bed gasifier has the gasification bed maintained at fast fluidisation regime. This results in large flow rate of gasifying medium, when

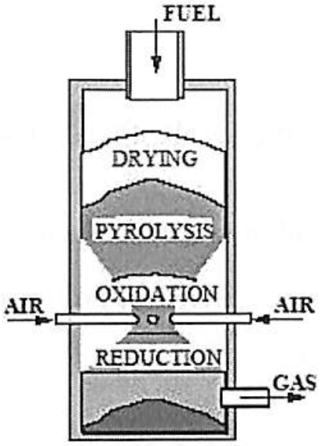


Figure 1.3: Co-current moving bed gasifier

air is used as gasifying medium results is dilution of product gas with large amount of nitrogen. When steam is used as gasifying medium large flow rate of steam is required to maintain fast fluidisation regime.

In this work two fluidised bed chambers are used to overcome the issue. The gasification section is maintained at bubbling fluidisation regime with steam and other chamber at fast fluidisation regime with air. The circulation of bed material between two chambers are maintained by balance of pressure between the two.

1.5.2 Indirect Heating System

In indirect heating system char is burnt and the heat is taken by a heat carrying medium to give the heat of gasification. There are various heat carrying medium available, in this work sand is used the heat carrying medium to carry heat from the riser to the gasifier. In indirect heating system combustion is separated from the gasification this helps to improve the gasification efficiency and thereby gives product gas with good heating value. Also dilution of product gas with nitrogen is avoided.

1.6 Circulating Fluidised Bed Process

At steady state the gasifier chamber has a bed of hot sand to which bagasse at 15% moisture content is added using a screw feeder.

In the gasifier, the temperature is at 850°C and at near atmospheric pressure. As the bagasse enters, first drying of bagasse takes place followed by devolatalisation or pyrolysis giving out vapours and char, then gasification of the char takes place to give carbon monoxide and hydrogen gas. The vapours formed during devolatalisation and the gases formed by char gasification undergo gas-gas reactions in the freeboard area above the bed in the gasifier giving out product gas. There is still char remaining after gasification which gets stuck to the sand particles. The sand along with this char is taken to the riser chamber through a chute at the bottom, to be burnt in the riser to produce the heat required in the gasifier. The sand acts as the heat carrier to transfer heat from the riser chamber to the gasifier chamber. The circulation rate of sand required to give the heat of gasification is calculated based on the energy balance. This circulation rate is used to calculate the elutriation rate in the riser.

In the riser chamber the char is burnt by passing air which helps to maintain fast fluidisation regime. The flow rate of air required to burn all the char from the gasifier is calculated and a 5% excess air is supplied. The velocity of the air is based on the flow rate and the area of the riser. The velocity should be sufficient to elutriate all the sand entering in the riser and maintain the circulation rate. The hot sand from the riser is collected at the top of the riser using cyclone and is sent back to

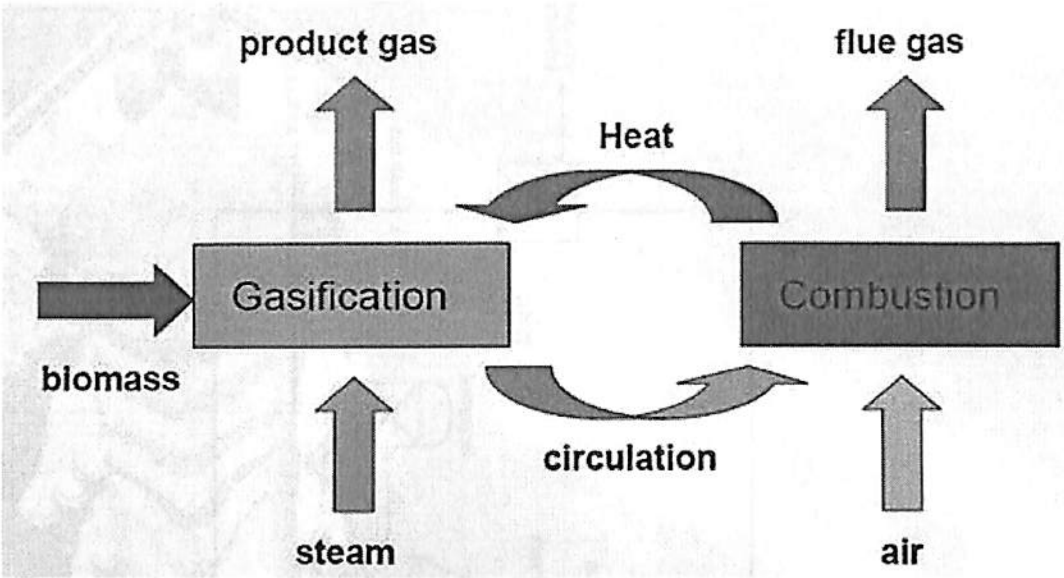


Figure 1.4: Indirect Gasification Concept

the gasifier. At steady state the process takes place continuously by difference in pressure between the two chambers.

When the heat released by combustion of char is not sufficient to meet the heat required for gasification of bagasse, 5-10% of the product gas is passed to the riser for combusting it to meet the heat requirement. During startup conditions the sand is added to the riser and air circulation is started and LPG burner is used at the bottom of the riser to heat the sand during the startup. As the temperature of sand rises slowly steam is passed and when balance of pressure is attained bagasse is added. Initially as the temperature of sand is low(say 500°C) which is the flame temperature of LPG, there will be formation of more char which is burnt to rise the temperature of sand to 850°C and when steady state condition is attained the yield of product gas is maximised and char yield is reduced.

The product gas from the gasifier is at around 850°C. It has to be cooled down

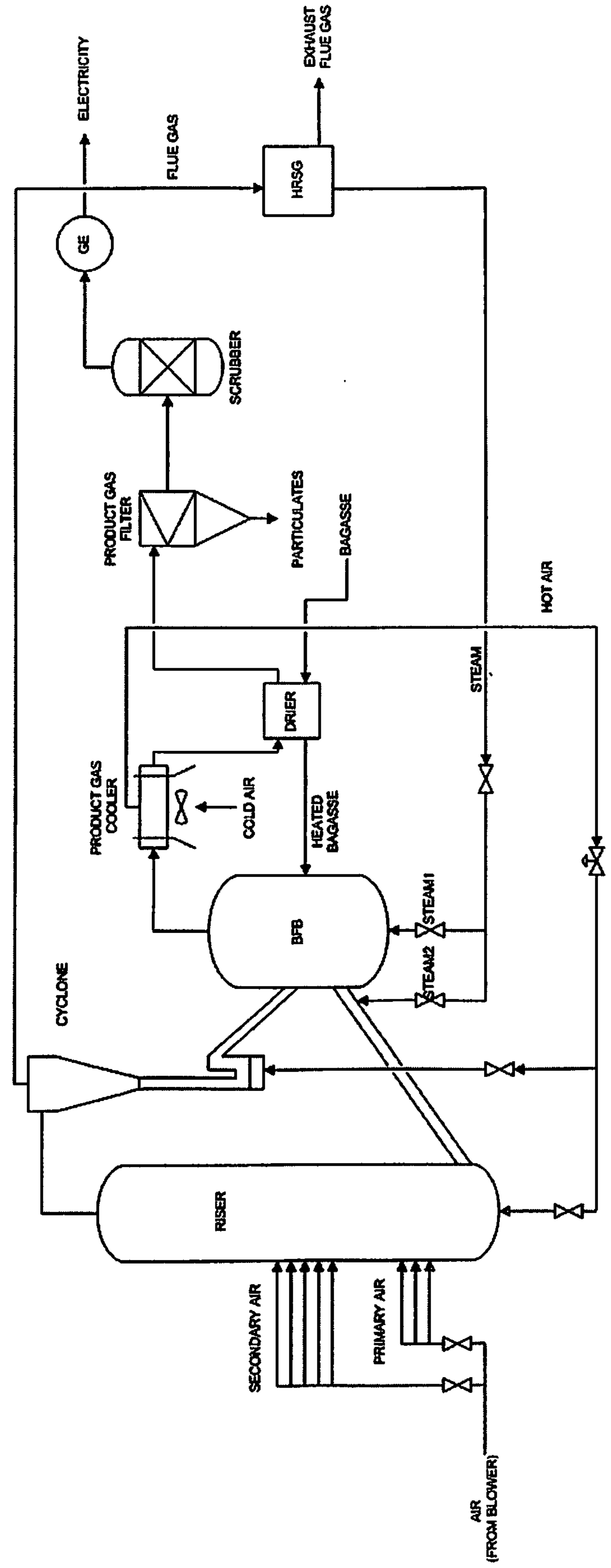


Figure 1.5: Process Flow Sheet

to 50°C for use in a gas engine. The product gas contains small amount of heavy hydrocarbons ('tars') which starts condensing as its temperature decreases. This tar has to be removed from the product gas before using it in gas engine. In order to avoid condensation of tar in the heat exchanger an air cooler is used to cool the product gas to 200°C. In air coolers the skin temperature of the tube is high which prevents condensation of tar in the tubes. Further cooling of the product gas is done in a scrubber. In the scrubber, water is used to scrub tar from the product gas and to cool the product gas down to 70°C. The hot air coming from the air cooler can be used in the riser chamber for combustion, which thereby reduces energy consumption.

The flue gas coming from the riser is at around 1000°C. The energy from this flue gas is used to generate steam at 350°C and 1.28 bar. This steam can be used for the process. The cycle efficiency of the system is thus improved.

The product gas from the gasifier has an heating value of around 14 MJ/m^3. The heating value of the gas can be improved using catalysts like dolomite crystals or olivine. The catalysts can be used along with the bed material to help crack the hydrocarbons in the bed and improve the efficiency of gasification.

Chapter 2

Literature Survey

Literature on biomass gasification for design and product gas estimation is reviewed from an engineering point of view. Design of a circulating fluidized bed gasifier for sugarcane bagasse has been done in this work and the product gas estimation was done by an equilibrium model.

2.1 Product Gas Estimation

The estimation of product gas composition can be done by non- stoichiometric equilibrium model or stoichiometric equilibrium model or kinetic model as shown in Basu 2010 [1]. For kinetic model the values of activation energy and pre exponential factor for bagasse at the operating conditions of the gasifier are necessary, though these values are available for rapid pyrolysis of bagasse as calculated by Zanzi [15],[5] for thermal degradation of cellulose, hemicellulose, lignin of bagasse [14], for pyrolysis of bagasse at high heating rates [13]. Pyrolysis characteistics of biomass and biomass components as suggested by Raveendran [12] gives an idea of the temperature ranges at which pyrolysis starts. The values for steam gasification of bagasse at 850°C are not available in literature. Therefore an equilibrium model as suggested in [8] is used in this work to estimate the product composition. To estimate the product gas composition, ultimate analysis and proximate analysis of the feed is necessary. In this work, these values are taken from literature as given by Kirubakaran [7].

2.2 Design of Gasifier

Design of circulating Fluidized Bed Gasifier involves sizing of gasifier, riser, connection chute and cyclone. The sizing is done based on fluid dynamics and hydrodynamics across the system. Hydrodynamics of dual fluidized bed gasifier system is analyzed and reported in two parts by Hofbauer.H. In this work hydrodynamic analysis is based on the literature by Hofbauer.H, B.leckner [10] [6].

Kaiser et al, studied the hydrodymics of a dual fluidized bed gasifier and have reported in two parts but his findings are specific to his experiments, in this work a attempt to make a genaralisied procedure for design based on properties feed is made. The first part describes about the riser with injection and the second part includes fluid dynamics for gasifier and pressure balance. In the first part for riser, the dimensions of riser are determined first based on fluidisation velocity and flow rate and these dimensions are used to calculate the pressure drop. The solids volume fraction in riser and gasifier column is calculated by applying the modified two phase theory as proposed from Johnsson, Andersson, and Leckner (1991) as given by Kunii [9].

The fluid dynamics of riser and gasifier includes determination of minimum fluidization velocity, terminal velocity, fluidization velocity, bed area, bubble volume fraction. In the gasifier properties of steam and the particles properties of sand are used. In the riser properties of air and sand are used. Bubble volume fraction, bubble rise velocity depends on the fluidisation velocity and particle size in the bed. B. Leckner [11] gives relation for bubble void fraction for various particle size. Based on bubble volume fraction the diameter of riser and gasifier are fixed.

Conection chute connects the gasifier column to the riser column. Kaiser et al considered the pressure drop in the connection chute as that similar to pressure drop in a pipe of similar dia, this assumption is not valid as the bed materials will flow in an irregular fashion in the chute. Hence in this work the pressure drop is calculated assuming that the bed materials will flow as beds because of the inclination. An expression suggested by Kunii [9] is used. A small amount of steam is injected in the chute to keep balance pressure and to maintain minimum fluidisation condition.

Chapter 3

Estimation Of Product Gas Composition

3.1 Equilibrium model

An equilibrium model predicts the product gas composition if the reactants are allowed to react in a fully mixed condition for an infinite period of time in order to reach equilibrium. However as an approximation we assume that equilibrium is achieved in the gasifer . This helps to find the maximum product that can be obtained if reaction proceeds to completion. Though only rough estimate is possible with equilibrium model, this can be a reasonable start for the design. As the reacion time is quite small this assumption is fairly reasonable. The products of gasification reactions mainly consist of methane, carbon dioxide, carbon monoxide, hydrogen, water vapour and char.

3.1.1 Determination of chemical formula for bagasse

In order to write a reaction for the gasification it is necessary to estimate an empirical molecular formula for bagasse. This can be done from the ultimate analysis of bagasse which is shown in Table 3.1.

Component	Percent
Carbon	43.8
Hydrogen	5.8
Oxygen	47.1

Table 3.1: Ultimate Analysis of Bagasse

From the composition, number of moles of carbon, hydrogen and oxygen are found. The chemical formula for the compound is determined by taking carbon basis.

Compound	Weight	M.wt	Moles	Mole fraction	C Basis
C	0.438	12	0.0365	0.293	1
H	0.058	1	0.058	0.466	1.589
O	0.471	16	0.0294	0.236	0.806

Table 3.2: Molecular Formula

3.1.2 Gasification reaction

Overall gasification reaction for 100Kg of bagasse reacting with 30Kg of steam can be represented as

$$3.766CH_{1.59}O_{0.81}+1.66H_2O \rightarrow n_1C+n_2H_2+n_3CO+n_4CO_2+n_5H_2O+n_6CH_4 \quad (3.1)$$

Carbon balance

$$n_1 + n_3 + n_4 + n_6 = 3.766 \quad (3.2)$$

Hydrogen balance

$$2n_2 + 2n_5 + 4n_6 = 9.31 \quad (3.3)$$

Oxygen balance

$$n_3 + 2n_4 + n_5 = 4.71 \quad (3.4)$$

3.1.3 Equilibrium constants

The three main gasification reactions that takes place are Boudouard reaction, Water-gas reaction and Hydrogasification reaction. During steam gasification water-gas reaction will be the dominant reaction compared to other gasification reaction but when air is used as the gasifying agent boudouard reaction will be the dominant reaction. The equilibrium rate constants of the three reaction are taken from literature [4]. These constants are either directly available from graphs or can be calculated from the Gibbs free energy of the reaction using the following relation. In this work the values are taken from graphs for the temperature of 850°C.

$$K_e = exp(-\Delta G/RT) \tag{3.5}$$

where Δ G = Δ H - TΔ S.

Boudouard Reaction

$$C + CO_2 \rightarrow 2CO \ldots R(1)$$

$$K_e = \frac{y_{CO}^2 P}{y_{CO_2}} \tag{3.6}$$

Water Gas Reaction

$$C + H_2O \rightarrow CO + H_2 \ \ldots \ R(2)$$

$$K_e = \frac{y_{CO} y_{H_2} P}{y_{H_2O}} \tag{3.7}$$

Hydrogasification Reaction

$$C + 2H_2 \rightarrow CH_4 \ \ldots \ R(3)$$

$$K_e = \frac{y_{CH_4}}{y_{H_2}^2 P} \tag{3.8}$$

where y represents the mole fraction of species.

Substituting the values of equilibrium constants found at 850°C and the total pressure of the system is estimated as 118Kpa using pressure balance. Taking $n7$ as the total number of moles, the above set of equation becomes

$$\frac{n_3^2}{n_7 n_4} = 0.139 \tag{3.9}$$

$$\frac{n_3 n_2}{n_7 n_5} = 0.139 \tag{3.10}$$

$$\frac{n_6 n_7}{n_2^2} = 5.3 \tag{3.11}$$

$$n_1 + n_2 + n_3 + n_4 + n_5 + n_6 = n_7 \tag{3.12}$$

3.1.4 Productgas Composition

Solving equations 3.2 to 3.4 and 3.9 to 3.12 simultaneously we get the number of moles of the product gas. From the number of moles, volume percent of the product gas can be found. The solution of the above set of equations are found as

Compound	Moles
Carbon	0.4723
Hydrogen	1.1350
Carbon monoxide	1.0128
Carbon dioxide	1.1848
H_2O	1.3277
Methane	1.0961
Total no of moles	6.2287

Table 3.3: Product Gas Composition

Volume percent of the product gas on dry basis are shown in table 3.4. The heating value of the product gas with above composition is found as 12.01 MJ/m^3. The above represented composition is achievable only at theoretical conditions that

Gas	Vol Percent
Carbon monoxide	22.87
Carbon dioxide	26.75
Hydrogen	25.63
Methane	24.75

Table 3.4: Volume Percent of Product gas

is when all the gasification reaction proceed to completion and equilibrium is established but at operating conditions these values can be the maximum possible values. At operating conditions in the freeboard the vapours formed from pyrolysis reacts with the steam and the gases formed after gasification to give the final product yield. The final product will therefore have lesser heating value compared to the one found. However for design purpose the above estimated values can be used as an initial guess and later during optimisation of the plant it can be modified.

Chapter 4

Mass And Energy Balance

4.1 Mass Balance

4.1.1 Fuel Feed Rate

The objective is to design a gasifier to gasify 100 kg/hr of bagasse. The fuel feed rate for the system is taken as 100 kg/hr of bagasse.

4.1.2 Steam Flow Rate

The steam to fuel ratio for steam gasification of biomass in Fast Internal Circulating Fluidized Bed(FICFB) will be in the range of 0.1 to 0.8 as given in literature [3]. For the present system, value of 0.3 is taken as the steam to fuel ratio. From the following graph taken from literature it is found that 0.3 steam to fuel ratio yields product gas with good heating value. The steam flow rate can be optimised later during operation of the plant.

4.1.3 Product Gas Composition and Char Flow Rate

Flow rate of the product gas and flow rate of char for the equilibrium composition are found as shown in the Table 4.1

Product gas flow rate in dry basis is 100 kg/hr and flow rate of char after gasification is found to 7.59Kg/hr.

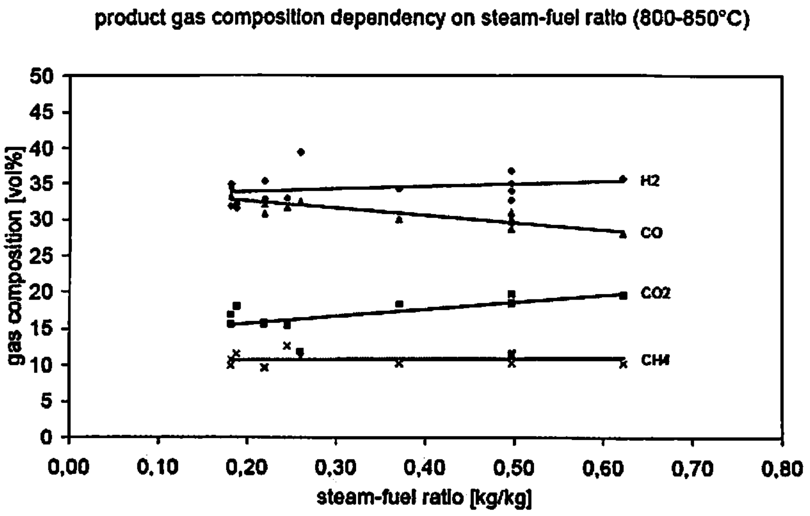


Figure 4.1: steam to fuel ratio (ref stiochiometric steam consumption, H.Hofbauer)

Compound	Molar flow rate kmol/hr	Mol% wet	Mass flow rate kg/hr
Hydrogen	1.14	18.22	2.24
Carbon monoxide	1.01	16.26	28.36
Carbon dioxide	1.18	19.02	52.13
Methane	1.1	17.6	17.54
H_2O	1.33	21.32	23.9
Total			124.2
Char	0.47	7.58	7.59

Table 4.1: Product flow rate

4.2 Energy Balance

4.2.1 Energy Required For Heating Bagasse

The energy required to heat bagasse from 150°C to 850°C will be approximately equal to the energy required for gasification.C_p of bagasse is given by the relation shown in equation 4.1.

$$C_p = 1112 + 4.85T \tag{4.1}$$

The energy required to heat 100 kg of bagasse is estimated as 54 kW, this energy should be supplied by the sand for gasification of bagasse. Refer to appendix for detailed calculation

4.2.2 Energy Required For Heating Air

In the riser chamber 160 kg/hr of air is required for burning 7.59 kg/hr of char and 13 m^3 of product gas.The equilibrium model predicts that 7.59 kg/hr of char is formed after gasification. This char is available for burning in riser. The heat taken by air/flue gas to heat up from 300°C to 1000°C is calculated as 34.75 kW.

$$\Delta H = \int mC_p dt \tag{4.2}$$

4.2.3 Energy Obtained By Burning Char

The amount of energy released by burning 7.59 kg/hr is found from the heating value of char. From literature [5] it is found that the heating value of char formed by steam gasification is 18.9 MJ/kg. The amount of energy released is found as 39.85 kW.

4.2.4 Energy Obtained by Burning of Product Gas

The amount of energy released by burning 13.5 m^3 of product gas. The heating value of the product gas is calculated as 13.29 MJ/m^3. The energy released by burning 13.5 % of product gas is calculated as 49.82 kW.

4.2.5 Circulation Rate of Sand

Sand from the riser should carry heat and should give the energy required for gasification in the gasifier. Therefore the mass of sand required for transfering 54 kW of energy is calculated as 1620 kg/hr and this circulation rate is maintained in the riser by adjusting the elutriation rate. The sand from the gasifier comes at 850°C and has to be heated to 1000°C to give heat of gasification. C_p of sand is found to be 0.8 from literature [2].

4.2.6 Energy Balance for riser

In the riser char and 13.5 % of product gas are burnt to heat the sand from 850°C to 1000°C. The air enters at 300°C and the flue gas from the riser leaves at 1000°C. The heat given by char burning and product gas burning should be equal to heat taken by sand and heat taken by flue gas.

Energy	Released	Taken
Char combustion	39.85	
Productgas combustion	49.82	
Heating 1620 kg/hr of sand		54
Heating 160 kg/hr of air		34.75
Total	89 kW	89 kW

Table 4.2: Energy balance for riser

4.2.7 Energy Required For Generating Steam

In process 30kg/hr of steam is required for gasification of char. This steam is generated from the hot flue gas which comes from the riser at 1000°C. The energy required for generating 30kg/hr of steam at 350°C and 1.28 bar is found to be 30 kW.

4.2.8 Energy Available in Hot Flue Gas

The flue gas from the riser comes at 1000°C which has to be cooled down to 200°C before being released into the atmosphere. The energy available for recovery from the flue gas is calculated as 34 kW. The C_p of flue gas is 0.96 KJ/kg K.The energy available from flue gas is sufficient to generate 30 kg/hr of steam at 350°C and 1.28 bar for the gasification process in the gasifier.

4.2.9 Energy Available in Hot Product Gas

The product gas comes from the gasifier at 850°C and has to be cooled down to 200°C before filtering and scrubbing. It is further cooled to 40°C before using it in the gas engine. The C_p of the product gas is shown in the Table 4.3.

Compound	a	b	c	d	C_P
H_2	29.11	-1.92E-03	4.00E-06	-8.70E-10	19341.02
CO	28.16	1.68E-03	5.37E-06	-3.60E-09	20134.93
CO_2	22.26	5.98E-02	-3.50E-05	7.47E-09	33076.09
CH_4	19.89	5.02E-02	1.27E-05	-1.10E-08	40291.15
H_2O	32.24	7.96E-05	1.32E-05	-4.55E-09	25016.63

Table 4.3: Heat capacity of product gas

The energy carried by 100 m^3 of product gas as sensible heat is found as 41 kW. An air cooler is used to cool the product gas from 850°C to 200°C using air as the cooling medium, the advantage of air cooler is that the skin temperature of the tube is high and it prevents condensation of tar in the tubes. The energy taken from the product gas to heat air is found as 22.65 kW. The outlet air is heated up to 300°C and a part of this hot air is used in the riser. In the air cooler the product gas is cooled to 450°C. The product gas at 450°C is used to preheat bagasse, heat given by product gas to preheat bagasse is 12.3 kW. After preheating bagasse, product gas is cooled to 200°C and the tar condensed is taken back to the gasifier to crack tars and improve efficiency.

4.2.10 Available low grade energy

There is some part of energy generated by the system which is of low grade and is not economical to recover this energy from the system.

Product gas has to be cooled from 850°C, the energy available as sensible heat in product gas is 41 kW. The product gas is cooled using an air cooler, in the air cooler 22.65 kW is taken by air to cool product gas from 850°C to 450°C for using it in riser. The hot air from the air cooler comes at 300°C. The product gas at 450°C is futher cooled to 250°C by exchanging heat with incoming bagasse, 12.3 kW of energy is used for preheating bagasse and the remaining 6.05 kW energy is available as excess energy. From the energy taken by air for cooling product gas in air cooler, 12.65 kW is utilised in the system and the remaining 10 kW is available as excess energy. Total excess energy available is 16kW.

Energy	Available	Taken
Sensible heat in product gas	41 kW	
Preheating air		12.65 kW
Preheating bagasse		12.3 kW
Excess	16 kW	

Table 4.4: Available excess energy

4.2.11 Heating Value of Product Gas

The calorific value of the product gas, for the composition estimated by equilibrium model is found as 370kW refer to table 4.5. When this product gas is used in a gas engine with 50% efficiency we could generate 185 kW of electrical energy. Refer to appendix for detailed calculation.

Compound	Heating value MJ/m^3	volume m^3	Calorific value $MJ/100\ m^3$
Hydrogen	9.88	25.63	253.2
Carbon monoxide	11.57	22.87	264.6
Methane	32.79	24.75	811
Total			1329

Table 4.5: Heating value of product gas

4.3 Gasification Efficiency

The heating value of bagasse with 15% moisture content is found to be 16.29 MJ/kg, the efficiency of gasification can be found from the heating value of the product gas. Gasification efficiency is the ratio of output energy to input energy.

$$\eta = \frac{370}{452.5} = 81.7\%. \tag{4.3}$$

Chapter 5

Design

5.1 Fluid Dynamics in Gasifier

Properties of Sand

Particle dia	120 μ m
Density	2650 Kg/m^3
Sphericity	0.8

Table 5.1: Properties of sand

Properties of Steam at 350C and 128.9 Kpa

Density	0.4493 Kg/m^3
Viscosity	2.23*10^{-6}Kg/ms

Table 5.2: Properties of steam

Properties of Bagasse

Particle Size	500 μ m
Particle density	111 Kg/m^3
Sphericity	0.7

Table 5.3: Properties of Bagasse

5.1.1 Estimation Of Minimum Fluidisation Velocity

The minimum velocity of the fluidising agent to fluidise the bed materials is found using the equation given below, which is a function of density of the bed materials density of the fluidising agent, viscosity of the fluidising agent and the voidage at minimum fluidised condition. The voidage at minimum fluidised condition is a function of Reynolds number and Archimides number.

$$U_{mf} = \frac{d_p^2 \left(\rho_p - \rho_g\right) g}{\mu} \frac{\epsilon_{mf}^3 \phi^2}{1 - \epsilon_{mf}} \tag{5.1}$$

Where ϵ_{mf} represents the voidage at minimum fluidisation and is obtained from the root of the Ergun equation shown below

$$\epsilon_{mf}^3 + \frac{150Re}{\phi^2 Ar}\epsilon_{mf} - 150Re + 1.75\frac{Re^2\phi}{\phi^2 Ar} = 0 \tag{5.2}$$

Re represents the Reynolds number for the particle and is obtained using the following equation C.3

$$Re = \left(27.2^2 + 0.048Ar\right)^{0.5} - 27.2 \tag{5.3}$$

Ar represents the Archimides number

$$Ar = \frac{d_p^3 \rho_g \left(\rho_p - \rho_g\right) g}{\mu^2} \tag{5.4}$$

5.1.2 Estimation Of Terminal Velocity Of Particle

The terminal velocity of the particle is estimated using the following equation. The terminal velocity of the particle is the maximum value of velocity for the fluidising agent at this velocity, the particle in the bed will start escaping from the gasifier i.e beyond this range of velocity the fluidising regime becomes fast fluidisation regime.

$$U_t = d_p \left[\frac{4(\rho_p - \rho_g)^2 g^2}{225\rho_g \mu}\right]^{1/3} \tag{5.5}$$

5.1.3 Fluidization Velocity

Fluidization velocity of the gasifier has a minimum value of minimum fluidisation velocity, which is lowest velocity to keep the bed in fluidised state and the maximum value is the terminal velocity. At the terminal velocity the bed becomes pneumatic transport bed. The bed in the gasifier is to be maintained at bubbling fluidised state and so the value of fluidisation velocity should lie between minimum fluidisation velocity and terminal velocity. The fluidisation velocity also affects the bubble dia in the bed. The fluidisation velocity is adjusted in such a way that the bubble dia in the bed for the bed height doesn't reach the dia of the gasifier, because at such condition slugging takes place which is undesirable. The bubble grows as it rises the bed.

5.2 Bed Area

Bed area is found from the volumetric flow rate of the fluidising agent and the fluidising velocity. In this work the volumetric flow rate of steam is taken as $0.03\text{m}^3/\text{s}$. The flow rate of steam can be fixed by taking the value of ratio of steam flowrate to biomass flowrate between 0.2 and 0.8, taken from literature [3]. In this work the value is taken as 0.3 which gives good yield of product gases. As the flow rate of steam increases the cracking of hydrocarbons starts in the freeboard space and also shift reaction takes place which increases the carbon dioxide composition. Steam generation also consumes energy and hence optimised values should be chosen. Excess steam in the product gas also causes condensation problem in handling the product gas downstream.

$$A_b = \frac{V_g}{U} \tag{5.6}$$

where, V_g is the Volumetric flow rate of gas

U is the velocity of the fluidising velocity.

5.3 Bubble Diameter and Bubble Rise Velocity

Bubble dia depends on the area of the bed and height of bed. From the area of the bed the dia of the bed can be calculated. The fluidising velocity of steam is adjusted in such a way that diameter of the bed is greater than the diameter of the bubble. The height bed sufficient for the gasification reaction is first assumed and is recalculated from the balance of solids in the bed.

$$d_b = 0.52\,(U - U_{mf})^{0.4} \times \left(h + 4\sqrt{A_b}\right)^{0.8} \times g^{-0.2} \tag{5.7}$$

From the diameter of the bubble the rising velocity of the bubble can be calculated using the following relation for bubble rise velocity

$$U_{br} = 0.711\,(gd_b)^{0.5} \tag{5.8}$$

5.4 Bubble Void Fraction

Bubble void fraction is the fraction in the bed occupied by the bubble, its a function of bubble dia and the difference between minimum fluidisation velocity and fluidising velocity.

$$\epsilon_b = \left[1 + \left(\frac{1.3}{f}\right) \times (U - U_{mf})^{-0.8}\right]^{-1} \tag{5.9}$$

$$f = [0.26 + 0.7 \times exp\,(-3.3d_p) \times (0.15 + (U - U_{mf}))]^{-1/3} \tag{5.10}$$

5.5 Gasifier Void Fraction

The void fraction of the gasifier is dependent on the voidfraction in the bubble phase and in the emulsion phase. It is assumed that the emulsion phase void fraction is equal to the void fraction at minimum fluidised condition.

$$\epsilon = \epsilon_b + (1 - \epsilon_b)\,\epsilon_e \tag{5.11}$$

$$\epsilon_e = \epsilon_{mf} \tag{5.12}$$

5.6 Pressure Drop In Gasifier

Pressure drop across the gasifier bed is obtained from the density of the particle in the bed multiplied by the space occupied by the material,height and acceleration due to gravity. From the pressure drop across the bed the pressure in the gasifier can be found. The Pressure at the gaifier outlet is used for balancing the pressure across the system and is also used to determine the composition of the product gas.

$$\Delta P_{gasifier} = (1 - \epsilon)\,\rho_g h g \tag{5.13}$$

5.6.1 Mass Of Solids In Gasifier

The average density of the bed is found using the fluidising velocity and voidage, from the average density of particle the mass of solids present in the bed for a particular height can be found. The height of the bed in the gasifier is found by iteratively to

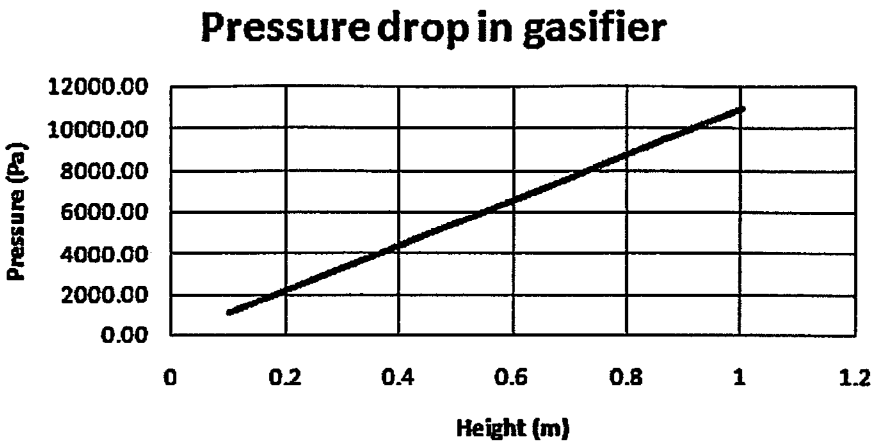


Figure 5.1: Gasifier pressure drop

match the pressure balance and the solids balance in the system.

$$\rho_{avg} = \rho_g \left(\frac{G_s}{G_g}\right)\left(\frac{U}{U_p}\right) \tag{5.14}$$

$$U_p = U - U_{mf} \tag{5.15}$$

$$G_s = \rho_p U_p (1 - \epsilon) \tag{5.16}$$

$$G_g = \rho_g U \tag{5.17}$$

Mass of solids in the bed

$$M_{gas} = (1 - \epsilon)\, \rho_{avg} A h \tag{5.18}$$

5.7 Connection Chute

The connection chute is located at the bottom of the gasifier section which connects to the riser. After gasification the sand along with char flows through the chute at minimum fluidised condition. The connection chute is a pipe, so the diameter of the chute is taken from the available standard dimensions of the pipe. The chute is kept at an inclination angle of 30°. As the angle of inclination increases the circulation rate increases. The pressure drop across the chute is found ref [9] from the expression for pressure drop through a moving bed downflow with the length resolved to its vertical component for the angle of inclination.

$$\Delta P_{chute} = \left[150\frac{(1-\epsilon_{mf})^2}{\epsilon_{mf}^2}\left(\frac{\mu\Delta u}{(\phi_s d_p)^2}\right) + 1.75\left(\frac{1-\epsilon_{mf}}{\epsilon_{mf}}\right)\left(\frac{\rho_g(\Delta u)^2}{\phi_s d_p}\right)\right] Lsin\theta \quad (5.19)$$

$$\Delta u = u_g + u_s \quad (5.20)$$

5.8 Riser

Riser is the combustion chamber, where the char from the gasifier is burnt using air. Air serves as the fluidising medium and it maitains the bed in fast fluidised condition. The flow rate of air required for complete combustion of char is calculated and from the volumetric flow rate of air the riser diameter is fixed. The air in the riser enters at three different locations, bottom, primary inlet and secondary inlet. The air should elutriate the particles entering the gasifier to the top of the riser. Hence the height of the riser is first fixed and the circulation rate for that height is calculated. There is a diffuser kept in the riser to reduces the velocity of the solids so that erosion at the surface of the riser is reduced.

5.8.1 Fluid-dynamics Of Riser

Minimum fluidisation velocity of the bed using air at 300° as the fluidising medium is found using equation 5.1. The voidage at minimum fluidised condition is found from the Archemedes number and Reynolds number for air using equations 5.2, 5.3 and 5.4. The terminal velocity of the bed materials are found using equation 5.5. In the riser, the bed is to be maintained at Pneumatic transport bed condition, for such condition the velocity of solids in the bed should be more than the terminal velocity. In the riser air enters at three different inlets, the velocity at the secondary inlet should be more to carry away the particles out of the riser to the cyclone seperator. The circulation rate of sand depends on the velocity of air at different inlets.

5.8.2 Pressure Drop In Riser

The pressure drop in riser is caused by the hydrostatic pressure drop due to the weight of the bed materials and frictional pressure drop due to moving bed materials in pnematic transport bed. The hydrostatic pressure drop is calculated by the distribution of voidage at different heights in the bed. The density of bed at different heights are calculated and from that the mass of materials at different height and elutriation rate of bed materials the frictional pressure drop in the bed is calculated.

5.8.3 Voidage At Different Height

The voidage at different height are calculated from the following equation. The voidage at dense zone is that voidage before the entry on primary air into the chamber. The voidage at infinity is calculated from the elutriation rate of the bed mate-

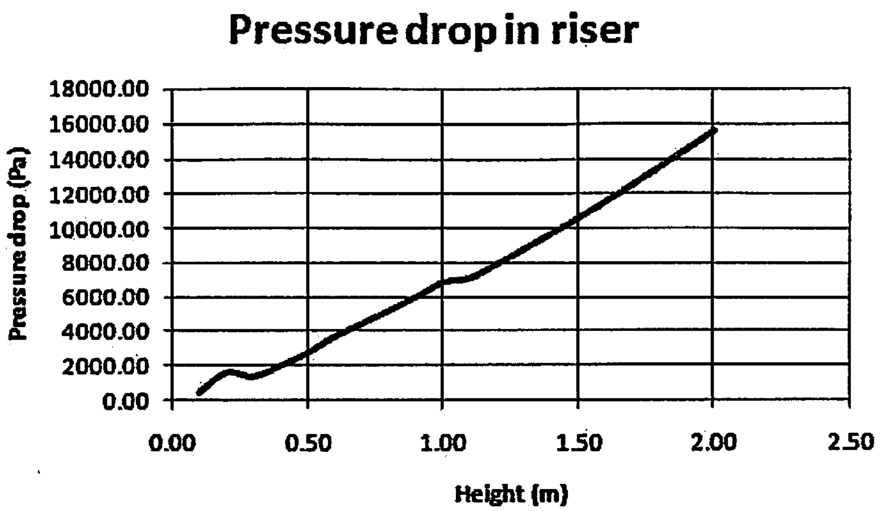


Figure 5.2: Riser pressure drop

rials.

$$\epsilon = [exp\left(-a\left(h_i - h_d\right)\right)\left(\epsilon_d - \epsilon_\infty\right)] + \epsilon_\infty \tag{5.21}$$

$$a = \frac{0.88 - 420d_p}{(U - U_t)^2 D^{0.6}} \tag{5.22}$$

$$\epsilon_\infty = \frac{K_i}{\rho_s(U - U_t)} \tag{5.23}$$

$$K_i = \rho_g U_G \times 23.7exp\left(\frac{-5.4U_t}{U}\right) \tag{5.24}$$

5.8.4 Solids Velocity In Bed

Velocity of the solids in the bed will be the velocity of the gas flowing through the chamber. The velocity of air is determined by the volumetric flow rate of air divided by the area available for the gas flow. The velocity differs at different air inlets because of change in flowrate of air in different sections.

$$U_s = \frac{V_g}{A\epsilon} \tag{5.25}$$

5.8.5 Average Riser Voidage At Different Zones

The average voidage of bed materials in denze zone, splash zone and transport zone are determined from the velocity of air at different inlets in the riser. ϵ_b gives the bubble void fraction in the bed and from this bubble void fraction the void fraction of the riser is determined at different zones.

$$\epsilon_b = \left[1 + \frac{1.3}{f}(U - U_{mf})^{(-0.8)}\right]^{-1} \tag{5.26}$$

$$f = [0.26 + 0.7 \times exp\,(-3.3d_p)]\,[0.15 + (U - U_{mf})]^{-1/3} \tag{5.27}$$

$$\epsilon = (1 - \epsilon_b)\epsilon_{mf} + \epsilon_b \tag{5.28}$$

5.8.6 Hydrostatic Pressure Drop

The voidage of bed at different height are determined and using the density of bed material i.e. sand, the hydrostatic pressure drop along the bed is determined.

$$\Delta P_{hyd} = (1 - \epsilon)\rho_s gh \tag{5.29}$$

5.8.7 Frictional Pressure Drop

The solid friction pressure drop coefficient is calculated based on the velocity of the gas. The average density of the bed at different height are calculated based on the solid velocity and gas velocity at different heights.

$$\Delta P_{fr} = \frac{2 f_s \rho_{avg} U^2 h}{D} \tag{5.30}$$

$$f_s = 0.05/U_g \tag{5.31}$$

$$G_s = U_s \rho_s (1 - \epsilon) \tag{5.32}$$

$$G_g = \rho_g U_g \tag{5.33}$$

$$\rho_{avg} = \rho_g \left(\frac{G_s}{G_g}\right) \left(\frac{U_g}{U_s}\right) \tag{5.34}$$

5.8.8 Pressure Drop

The difference between the hydrostatic pressure drop and friction pressure drop gives the total presure drop across the bed in the riser. Inlet pressure minus the pressure drop in the riser will give the operating pressure of the riser.

$$\Delta P_{riser} = \Delta P_{hyd} - \Delta P_{fr} \tag{5.35}$$

5.9 Cyclone

The diameter of the cyclone is decided based on the flow rate of air and solids and the velocity at the inlet duct of the cyclone. From the diameter of the cyclone necessary for the separation of the solid particle the height of cylindrical section, height of conical section and other dimensions of the cyclone are determined.

Diameter of cyclone

$$D_c = 1000\sqrt{10 A_i} \tag{5.36}$$

Width of inlet duct

$$B_c = D_C/4 \tag{5.37}$$

Diameter of gas outlet pipe

$$D_e = D_c/2 \tag{5.38}$$

Height of inlet duct

$$H_c = D_C/2 \tag{5.39}$$

Height of cylindrical section

$$L_c = 2D_c \tag{5.40}$$

Depth of exit gas pipe below inlet duct

$$S_c = D_c/8 \tag{5.41}$$

Height of conical section

$$Z_c = 2D_c \tag{5.42}$$

Diameter of solids outlet pipe

$$J_c = 2D_c \tag{5.43}$$

Minimum Particle dia

$$d_{min} = 120\mu m \tag{5.44}$$

Cut size of particle to be seperated

$$d_p = \sqrt{\frac{9\mu L}{\pi N_t V_i(\rho_s - \rho_g)}} \tag{5.45}$$

Particle capture

$$\eta = \frac{1}{1 + (d_p/d_{min}^2} 100 \tag{5.46}$$

5.9.1 Pressure Drop In Cyclone

The pressure drop across the cyclone is determined using the following relation where friction factor is taken as 0.005 and C_e is 2Kg/Kg.

Pressure Drop In Seperation Compartment

Pressure drop in seperation compartment

$$\Delta P_f = f_w \left(\frac{Ar}{0.9Q}\right) \left(\frac{\rho_g}{2}\right) \left(\frac{U_a}{U_i}\right)^{3/2} \tag{5.47}$$

$$f_w = f(1 + 2\sqrt{c_e}) \tag{5.48}$$

Tangential velociy

$$U_a = \left[\frac{V_e \times \left(\frac{r_e}{r_a}\right)}{\alpha}\right] \tag{5.49}$$

$$r_a = \frac{D_c}{2} \tag{5.50}$$

$$r_e = r_a - \left(\frac{L}{2}\right) \tag{5.51}$$

Tangential velocity at r$_i$

$$U_i = \left[\frac{U_a \left(\frac{r_a}{r_i}\right)}{1 + \left(\frac{f_w}{2}\right) \left(\frac{Ar}{Q}\right) U_a \sqrt{\frac{A_r}{r_i}}}\right] \tag{5.52}$$

$$r_i = \frac{D_c}{2} \tag{5.53}$$

Pressure Drop In Gas Outlet Pipe

$$\Delta P_e = \left[2 + 3\left(\frac{U_i}{V_i}\right)^{4/3} + \left(\frac{U_i}{V_i}\right)^2 \left(\frac{\rho_g}{2}\right) \left(\frac{V_i}{2}\right)\right] \tag{5.54}$$

$$V_i = \frac{Q}{\pi r_i^2} \tag{5.55}$$

Total Pressure Drop

The total pressure drop across the cyclone is sum of pressure drop in seperation compartment and pressure drop in gas outlet pipe

$$\Delta P_{cyc} = \Delta P_f + \Delta P_e \tag{5.56}$$

5.10 Seal Loop

The main task of the seal loop is to avoid gas mixing of combustion air with product gas. The downcomer and the standpipe of the seal loop are modelled as bubbling fluidized beds.The horizontal connection between these parts is modelled by using the correlation for friction force. The height of the bed at the side of the stationary fluidized zone is as high as the standpipe. The height of the bed at the side of the combustion zone was calculated to fullfil the pressure balance of the seal loop.

Pressure drop in the horizontal is found using the following correlation.

$$\Delta P_{horizontal} = [\rho_s(1-\epsilon_{mf}) + (\rho_g \epsilon_{mf})] \times \left[(9.8 sin\theta) + \left(\frac{2 f_s U_s^2}{D}\right)\right] L \qquad (5.57)$$

Where f_s is taken as 0.05 for particles of the size $120\mu m$ as given in [?] for sand particles.

The difference between the pressure in the riser and the gasifier is balanced by the seal loop by adjusting the height of the downcomer which matches the height of the bed in the gasifier.

$$P_{riser} - P_{gasifier} = \Delta P_{SLsta} - \Delta P_{SLdow} - \Delta P_{SLhor} \qquad (5.58)$$

5.10.1 Pressure Balance Around The System

The difference in operating pressure between the riser and gasifier should be equal to sum of gasifier pressure drop and connection pressure drop minus riser pressure drop and cyclone pressure drop.

$$P_{riser} - P_{gasifier} = \Delta P_{gas} + \Delta P_{con} - \Delta P_{Riser} - \Delta P_{cyc} \qquad (5.59)$$

Chapter 6

Results and Discussion

6.1 Results

The design of a'Circulating Fluidised Bed Gasifier' was carried out and the dimensions of the gasifier, riser, connection chute and the seal loop were found as shown in section 6.1.1

Requirements

1. Basis 100Kg/hr feed of bagasse
2. Sand avg particle dia 120μ m
3. Bagasse particle dia 500μ m
4. Steam at 350°C and 128.9 Kpa
5. Air at 300°C and 137 Kpa

6.1.1 Dimensions

Gasifier chamber

Diameter - 0.38m
Height - 3m
Material of construction - mild steel with an insulation of refractory layer of cement
Structure - Cylinder
Steam inlet - Bottom

Riser chamber

Diameter below diffuser - 0.15m
Diameter above diffuser - 0.17m
Height of riser - 2m
Material of construction - mild steel with an insulation of refractory layer of cement
Structure - Cylinder with diffuser at a height of 1m
Air inlet - At three different position
- bottom, primary and secondary
LPG burner(for start up) - At the bottom of riser

Connection chute

Diameter - 0.07m
Length - 0.6 m
Angle of inclination - 35°
Steam inlet - Using distributors all through the length of chute

Cyclone

Cylindrical section height - 0.71m
Conical section height - 0.71m
Diameter of cyclone - 0.35m
Solids outlet dia - 0.08m
Inlet dia - 0.17m

Stand pipe

Height - 0.4m

Diameter - 0.08m

Horizontal

Length - 0.15m

Diameter - 0.06m

Down comer

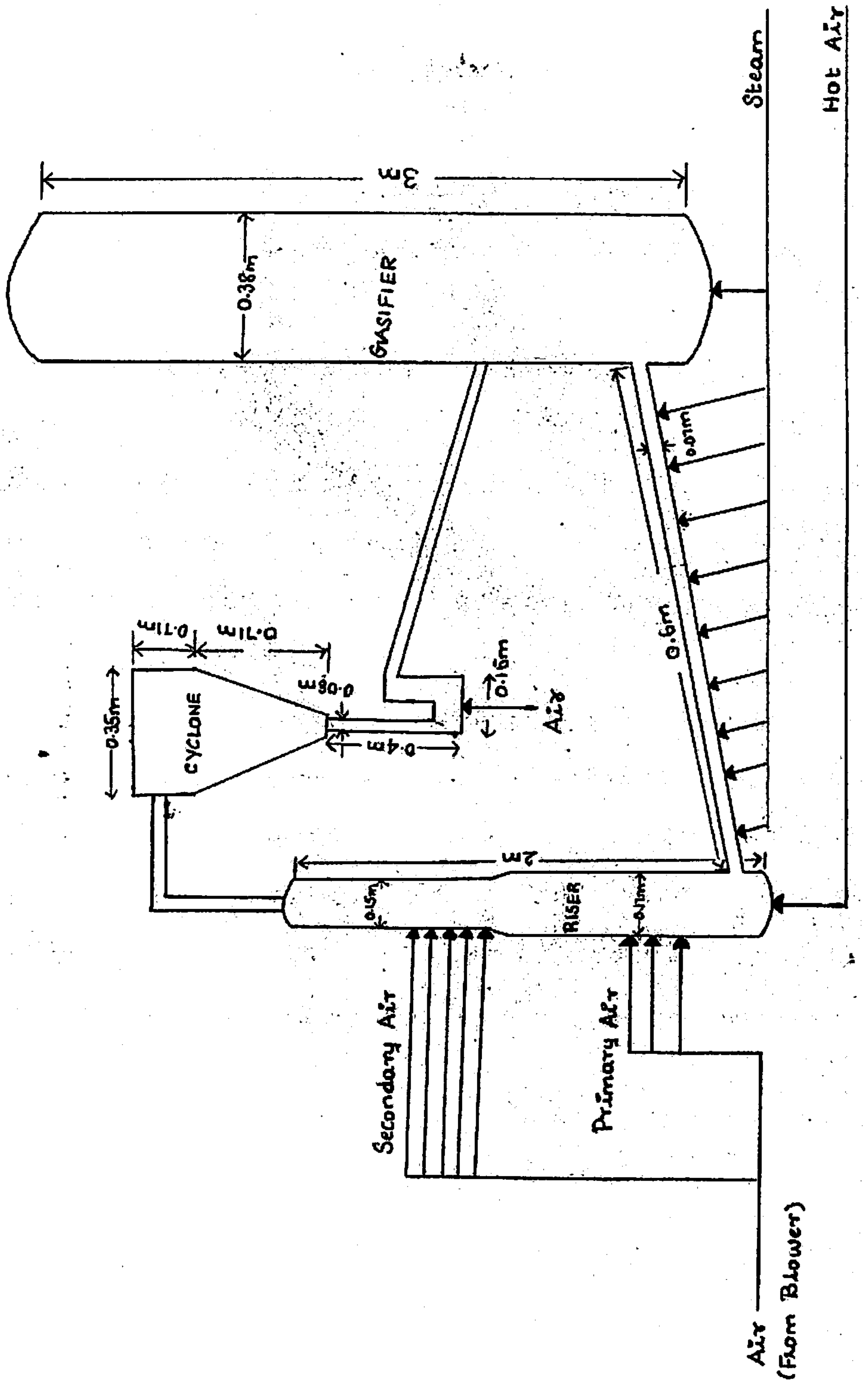
Height - 0.06m

6.2 Discussion

In 'Circulating Fluidised Bed Gasifier', steam is injected at the connection chute to maintain minimum fluidised condition in the chute. The steam in the chute prevents entry of air to the gasifier section. The designed gasifier has advantage of varying parameters, such as bed temperature and steam to fuel ratio. The bed temperature is maintained by the circulation of sand between the riser and the gasifier. The circulation rate can be varied by changing the air velocity at different inlets to the riser. The temperature of sand in the riser is controlled by char flowrate to the riser and additional fuel input to the riser.

The circulation rate of sand is calculated by heat balance. The composition of product gas is affected by bed temperature and steam to fuel ratio. The designed system can be optimised to get product gas with good heating value during operation. The bed material considered for design is sand but additional catalyst like olivine or dolomite crystals can be added to bed materials to improve cracking of tar and theryby improve product gas quality.

Air is injected at the syphon to fluidize the sand in the horizontal, standpipe and downcomer. The bed materials from the syphon are returned to the gasifier just


3m
0.38m
GASIFIER
Steam
Hot Air
0.07m
0.6m
0.16m
Air
0.11m
0.11m
0.08m
0.35m
CYCLONE
0.4m
2m
0.15m
RISER
0.1m
Secondary Air
Primary Air
Air
(From Blower)

above the bed. The fuel fed above the bed using a screw feeder.

The product gas from the gasifier comes at around 850°C. The heavy hydrocarbons present in the product gas are in gaseous form at this temperature but when the product gas is cooled these heavy hydrocarbon starts condensing and forms tar. To avoid condensation of tar in the heat exchanger an air cooler is used, the advantage of air cooler over water is the skin temperature of the tube in air cooler is high, this helps to avoid condensation of tar in the tube. The hot air that comes from the air cooler is used in the riser for combustion. This improves the energy efficiency around the system. The flow rate of air required for cooling is a function of product gas outlet temperature. Hence in this system the product gas is cooled upto 400°C in air cooler so that the hot air from air cooler is sufficient to be used in the riser. Further cooling of product gas is done in a chamber where the hot product gas is passed over the feed bagasse to preheat bagasse.

The product gas after preheating bagasse comes at around 200°C. The product gas carries some particulates as it is passed through bagasse, these particulates are removed in scrubber. The advantage of using hot product gas for bagasse preheating is that the tar which condenses by cooling is taken back to the gasifier along with the feed, where cracking is done with steam in the gasifier. This helps to improve product gas composition. Further cooling of product gas from 200°C to 50° is done in the scrubber.

6.2.1 Startup Considerations

During startup of gasifier, the bed materials is fed to the riser. The riser is provided with an LPG burner at the bottom. Air compressor or the blower is started and the sand is slowly heated up by burning LPG and the slowly the flow rate anf belovity of air in the riser is increased to carry the sand particles away from the riser. The flame temperature of LPG is not sufficient to reach 850°C. As the sand enters the gasifier section the steam is injected to maintain the flow of sand between the two chambers. As the temperature of sand reaches around 600° by LPG burning, Bagasse is fed to the gasifier. As the bed temperature is less more amount of char is formed which is burnt in the riser using air, to increase the temperature. When the sand reaches 850°C steady state is attained.

Amount of char formed during the reaction is estimated from the oxygen used for combustion in the riser. An oxygen analyser is used to analyse the flue gas. A gas chromotograph is used to analyse the product gas.

Chapter 7

Future Work

In this work an equilibrium model is used to determine the product gas composition which is a good approximation. Yet, for the estimation of the exact composition a kinetic model is necessary. Kinetic model requires experimental data to estimate preexponential factor and Arhenius energy.

A CFD modeling can be carried out to determine the exact heat distribution in the system. The heat from the product gas and flue gas has to be recovered in an economical way and hence possible economic options for the same has to be analyzed. Tar production in the system has to be minimized, possible catalyst for in-bed tar reduction can be studied.

In this work product gas after preheating bagasse comes at around 250°C. Possible filter to withstand such temperature and prevent clogging by tar condensation has to be looked for. In this work ash removal from the system is done from the cyclone and possible ways to remove them from the bed is yet to be analyzed. Attrition of sand takes place in the bed, as a result the size of sand reduces and gets away from the cyclone. Thereby reduces the amount of sand in the bed, in order to overcome this loss, additional sand has to be added to the bed. Possible ways to feed sand into the bed to maintain continuous operation is yet to be analyzed.

Bibliography

[1] Prabir Basu. *Biomass Gasification and Pyrolysis.* Elsevier, 2010.

[2] Jock Brown. Biomass gasification: Fast internal circulating fluidised bed gasifier characterisation and comparison. *University of Canterbury*, 2006.

[3] H.Hofbauer and R.Rauch. Stiochiometric water consumption of steam gasification by the ficfb gasification process. *Fuel and environmental technology.*

[4] J.M.Smith, H.C. Van Ness, and M.M.Abbott. *Introduction to Chemical Engineeing Thermodynamics.* McGraw HIll, 2001.

[5] Rajeev jorapur and Anil K. Rajvanshi. Sugarcane leaf bagasse gasifier for industrial heating purpose. *Biomass and Bioenergy*, 13:141–146, 1997.

[6] S. Kaiser, G. Loffler, K. Bosch, and H. Hofbauer. Hydrodynamics of a dual fluidized bed gasifier. part ii: simulation of solid circulation rate, pressure loop and stability. *Chemical Engineering Science*, 58:4215 4223, 2003.

[7] V. Kirubakaran, V. Sivaramakrishnan, R. Nalini, T. Sekar, M. Premalathae, and P. Subramanian. A review on gasification of biomass. *Renewable and Sustainable Energy Reviews*, 13:179186, 2009.

[8] Vigneswaran KUMARAN. *Design of a low cost fluidized bed gasifier for sawdust gasification in rural china.* PhD thesis, The University of Melbourne, 2007.

[9] Daizo Kunii and Octave Levenspiel. *Fluidisation Engineering.* Elsevier, 2005.

[10] G. Loffler, S. Kaiser, K. Bosch, and H. Hofbauer. Hydrodynamics of a dual fluidized-bed gasifierpart i: simulation of a riser with gas injection and diffuser. *Chemical Engineering Science*, 58:4197 4213, 2003.

[11] A. Go mez Barea and B. Leckner. Modeling of biomass gasification in fluidized bed. *Progress in Energy and Combustion Science*, 36:444509, 2010.

[12] K. Raveendran, Anuradda Ganesh, and Kartic C. Khilar. Pyrolysis characteristics of biomass and biomass components. *Fuel*, 75, No-8:987–998, 1996.

[13] J. F. Stubington and S. Aiman. Pyrolysis kinetics of bagasse at high heating rates. *Energy & Fuels*, 8:194–203, 1994.

[14] Gabor Varhegyif and Michael J. Antal. Kinetics of the thermal decomposition of cellulose, hemicellulose, and sugar cane bagasse. *Energy & Fuels*, 3:329–335, 1989.

[15] Rolando Zanzi, Krister Sjstrm, and Emilia Bjrnbom. Rapid pyrolysis of bagasse at high temperature. *Proceedings of Third Asia-Pacific International Symposium on Combustion and Energy Utilization*, 1:211–215, 1995.

Appendix A

Energy Balance

Engergy Required To Heat Bagasse

Energy required to heat 100 kg bagasse from 150°C to 850°C is approximately the heat required for gasification. The enthalpy is calculated using the followin relation.

$$\Delta H = \int mC_p dt \tag{A.1}$$

$$C_p = 1112 + 4.85T \tag{A.2}$$

$$\Delta H = 100 \times [1112(850-150) + \frac{4.85}{2}(850-150)^2]$$

$$\Delta H = 196665000 J/hr$$

$$\Delta H = 196665/3600 = 54KW$$

This energy has to be supplied from the riser chamber through sand. The energy released in riser chamber by combustion of char and part of product gas should be sufficient to provide energy to heat sand from 850°C to 1000°C and to heat air/flue gas from 300 to 1000°C.

Energy Required For Air Heating

The energy required for heating 160 kg of air from 300°C to 1000°C is calculated as shown below. This much of energy is being carried by air from the riser chamber.

$$\begin{aligned}\Delta H &= 160 \times 1.117 \times (1000-300) \\ &= 34.75KW\end{aligned} \tag{A.3}$$

Sand Circulation Rate

The gasification energy has to be supplied by sand. The circulation rate of sand is determined from this energy as shown below.

$$\Delta H = 54KW$$

$$\begin{aligned} 54 &= massofsand \times 0.8 \times (1000 - 850) \\ sandCirculationrate &= \frac{54}{0.8(1000 - 850)} \\ &= 1620kg/hr \end{aligned}$$

Energy Released By Combustion

The energy released by combustion of 7.59 kg/hr of char and 13.5 m^3 of product gas should give 54 + 34.75 KW. The heating value of char is 5.25 KW/kg and that of product gas is 3.69 KW/m^3 ref table A.1. Energy Released By Char Energy released by combustion of 7.59 Kg/hr of char is calculated as shown below.

Compound	Heating value MJ/m^3	volume m^3	Calorific value MJ/100 m^3
Hydrogen	9.88	25.63	253.2
Carbon monoxide	11.57	22.87	264.6
Methane	32.79	24.75	811
Total			1329

Table A.1: Heating value of product gas

$$\Delta H = 7.59 \times 5.25$$

$$\Delta H = 39.85KW$$

Energy Released By Product Gas Energy released by combustion of 13.5 m^3 of product gas is calculated as shown below.

$$\Delta H = 13.5 \times 3.69$$

$$\Delta H = 49.82KW$$

Energy Balance For Riser

Energy	Released	Taken
Char combustion	39.85	
Productgas combustion	49.82	
Heating 1620 kg/hr of sand		54
Heating 160 kg/hr of air		34.75
Total	89 KW	89 KW

Table A.2: Energy balance for riser

Energy As Sensible Heat In Flue Gas

The flue gas from the riser comes at 1000°C and it has sensible heat which can be recovered. The energy available in the form of sensible heat is calculated as shown below.

$$\Delta H = 160 \times 0.96 \times (1000 - 200)$$

$$\Delta H = 34.13KW$$

Energy As Sensible Heat in Product Gas

The product gas comes from the gasifier at 850°C and this gas has 41 KW of energy which can be recovered. The values are tabulated in table A.3 This energy from product gas is partially used to generate hot air for riser and partially for bagasse preheating and the rest is nonrecoverable low grade energy.

Compound	Flow rate	$\int_{50}^{850} C_p$
Hydrogen	1.14	26996.17
Carbon monoxide	1.01	24798.40
Carbon dioxide	1.18	46232.77
Methane	1.1	51118.38
Total		149145.73

Table A.3: Sensible heat in product gas
Coverting to KW we get 41 KW

Compound	Flow rate	$\int_{200}^{450} Cp$
Hydrogen	1.14	8328.35
Carbon monoxide	1.01	7652.84
Carbon dioxide	1.18	13853.88
Methane	1.1	14322.07
Total		44157.16
		12.26 KW

Table A.4: Available energy to preheat bagasse

Energy Required to Heat Air from 40°C to 300°C

The energy required to heat 160 kg/hr of air from 40°C to 300°C is caculated as shown below

$$\Delta H = 160 \times 1.021 \times (300 - 40)$$

$$\Delta H = 42473600 = 12KW$$

Energy Available to Preheat Bagasse

The product gas at 450°C is used to preheat bagasse. The product gas is cooled upto 200°C and energy available for recovery is calculated from C_p of product gas is shown in table A.4.

Energy	Available	Taken
Sensible heat in product gas	41 KW	
Preheating air		12.65 KW
Preheating bagasse		12.3 KW
Excess	16 KW	

Table A.5: Available excess energy

Available low Grade energy

The energy available as sensible heat in product gas is 41 KW, 22.65 KW is used to preheat air from 40°C to 300°C for using it in riser. 12.3 KW of energy is used for preheating bagasse and the remaining 6.05 KW energy is available as excess energy. From the energy taken for heating air, only 12.65 is utilised and the remaining 10 KW is available as excess energy. Total excess energy available is 16KW.

Appendix B

Gasifier Design

In Gasifier steam is used as the fluidizing medium to maintain bubbling fluidization regime. Steam at 350°C and 1.28 bar is used. The properties are calculated correspondingly.

Minimum Fluidization Velocity

$$U_{mf} = \frac{d_p^2\left(\rho_p - \rho_g\right)g}{\mu}\frac{\epsilon_{mf}^3\phi^2}{1-\epsilon_{mf}} \tag{B.1}$$

Where ϵ_{mf} represents the voidage at minimum fluidisation and is obtained from the root of the Ergun equation shown below

$$\epsilon_{mf}^3 + \frac{150Re}{\phi^2 Ar}\epsilon_{mf} - 150Re + 1.75\frac{Re^2\phi}{\phi^2 Ar} = 0 \tag{B.2}$$

Re represents the Reynolds number for the particle and is obtained using the following equation C.3

$$Re = \left(27.2^2 + 0.048Ar\right)^{0.5} - 27.2 \tag{B.3}$$

Ar represents the Archimides number

$$Ar = \frac{d_p^3\rho_g\left(\rho_p - \rho_g\right)g}{\mu^2} \tag{B.4}$$

The voidage at minimum fluidisation is calculated from the root of Ergun equation by using trail and error method.

Minimum Velocity For Sand

In the bed, sand is heavier then bagasse and the fluidization velocity should correspond to fluidization of sand hence minimum fluidisation velocity is calculated for properties of sand.

Ar = 4050.7

Re = 2.9

$\epsilon_{mf} = 0.462$

$U_{mf} = 0.12ms^{-1}$

Terminal velocity

Terminal velocity for both sand and bagasse are calculated. This value represents the velocity at which the particle will elope from the bed and leave the gasifier. The fluidisation velocity should be less than this value for maintaining bubbling fluidisation regime.

$$U_t = d_p \left[\frac{4 * (\rho_p - \rho_g)^2 * g^2}{225 * \rho_g * \mu} \right]^{1/3} \tag{B.4}$$

$$U_{tsand} = 2.7$$

$$U_{tbagasse} = 1.4$$

Fluidisation Velocity

The fluidisation velocity for the gasifier is fixed between the minimum fluidisation velocity and terminal velocity. This value is then adjusted from calculating the diameter of bed from bed area. As fluidisation velocity increases bed diameter decreases and bubble diamet increases. Initially a bed height is fixed and bubble dia at this height is calculated and corresponding fluidisation velocity is fixed. Later this bed height is adjusted from the mass of solids in the bed.

$$U = 0.30ms^{-1}$$

Bed Area

Bed area is calculated from volumetric flowrate of steam and fluidisation velocity. The volumetric flow rate of steam is fixed based on the steam to fuel ratio, which is taken as 0.3 for this system. Hence the volumetric flow rate of steam is 30 kg/hr for 100kg/hr feed of bagasse. Taking fluidization velocity as 0.3 ms^{-1}. The bed area is calculated as 0.11 m^2

$$A_b = \frac{V_g}{U_g} \tag{B.5}$$

Mass flow rate = 30/3600 = 8.33*10^{-3} kg/s

Density = 0.2 kg/m^3

Volumetric flow rate = 0.03 kg/s

$$V_g = 0.03kgs^{-1}$$

$$U_g = 0.3ms^{-1}$$

$$A_b = 0.11m^2$$

Bed Diameter

Bed diameter is calculated from the bed area. The bed area is found as 0.11 m^2.

$A_b = 0.11m^2$

$\pi \times r^2 = 0.11$

$r = 0.19m$

$D = 0.38m$

Bubble Diameter

Bubble diameter is calculated from the bed area and bed height. The fluidisation velocity is adjusted till the bubble diamter is less than the diameter of the bed. Height of bed is taken as 1 m, which has been verified later fromthe mass of bed solids and correspoding pressure drop accross the bed.

$$d_b = 0.52\,(U - U_{mf})^{0.4} \times \left(h + 4\sqrt{A_b}\right)^{0.8} \times g^{-0.2} \tag{B.5}$$

$$d_b = 0.34m$$

For fluidisation velocity of 0.3 m/s, the buuble diameter is less than the bed diameter, hence this velocity is fixed as the fluidization velocity for maintaining bubbling fluidization regime in the bed. Bubble rise velocity is calculated, this velocity is taken as the velocity with which the bed materials in core moves.

$$U_{br} = 0.711 * (g * d_b)^{0.5}$$

$$U_{br} = 1.29ms^{-1}$$

The bubble rise velocity is less than the terminal velocity of the particles hence the estimation of fluidisation velocity is validated.

Bubble Void Fraction

Bubble void fraction represents the area occupied by the bubble in the bed. It is calculated from the fluidisation velocity and minimum fluidization velocity.

$$\epsilon_b = \left[1 + \left(\frac{1.3}{f}\right)(U - U_{mf})^{-0.8}\right]^{-1} \tag{B.7}$$

$$f = [0.26 + 0.7 \times exp(-3.3d_p) \times (0.15 + (U - U_{mf}))]^{-1/3} \tag{B.8}$$

$$f = 1.47$$

$$\epsilon_b = 0.22$$

Gasifier Void Fraction

The void fraction in the bed is calculated from the region occupied by bubble.

$$\epsilon = \epsilon_b + (1 - \epsilon_b) * \epsilon_e \tag{B.9}$$

$$\epsilon_e = \epsilon_{mf} \tag{B.10}$$

$$\epsilon_e = 0.467$$

$$\epsilon = 0.58$$

Pressure Drop Across The Bed

The pressure drop is calculated using the voidage and density of bed materials at different heights.

$$\Delta P_{gasifier} = (1 - \epsilon)\,\rho_g hg \tag{B.11}$$

The values of pressure drop at different height from the bottom to top is shown in table

The inlet pressure of the steam minus the pressure drop across the bed will give the pressure in the gasifer. The pressure in the gasifier is used for balance of pressure across the system.

Height(m)	Pressure Drop (Pa)
0.1	1090.18
0.2	2180.35
0.3	3270.53
0.4	4360.70
0.5	5450.88
0.6	6541.05
0.7	7631.23
0.8	8721.41
0.9	9811.58
1	10901.76

Table B.1: Pressure Drop in gasifier

Appendix C

Riser

In riser air is used as fluidization medium to maintain fast fluidisation regime. Air at 300°C is used. The riser is at 1000°C. Properties of air are calculated correspondingly.

Minimum Fluidization Velocity

$$U_{mf} = \frac{d_p^2\left(\rho_p - \rho_g\right)g}{\mu}\frac{\epsilon_{mf}^3\phi^2}{1-\epsilon_{mf}} \tag{C.1}$$

Where ϵ_{mf} represents the voidage at minimum fluidisation and is obtained from the root of the Ergun equation shown below

$$\epsilon_{mf}^3 + \frac{150Re}{\phi^2 Ar}\epsilon_{mf} - 150Re + 1.75\frac{Re^2\phi}{\phi^2 Ar} = 0 \tag{C.2}$$

Re represents the Reynolds number for the particle and is obtained using the following equation C.3

$$Re = \left(27.2^2 + 0.048Ar\right)^{0.5} - 27.2 \tag{C.3}$$

Ar represents the Archimides number

$$Ar = \frac{d_p^3\rho_g\left(\rho_p - \rho_g\right)g}{\mu^2} \tag{C.4}$$

$$Ar = 1428.09$$

$$Re = 1.05$$

$$\epsilon_{mf} = 0.46$$

$$U_{mf} = 0.06ms^{-1}$$

Terminal Velocity

$$U_t = d_p \left[\frac{4 * (\rho_p - \rho_g)^2 * g^2}{225 * \rho_g * \mu}\right]^{1/3} \tag{C.5}$$

$$U_t = 1.9ms^{-1}$$

Fluidization velocity

In riser fast fluidization regime is to be maintained, hence the velocity of bed materials should be more than the terminal velocity. In the riser air enters at three different inlet. The average velocity is taken to calculate the bed area and the bed diameter. The average fluidisation velocity in the bed is taken as 4 ms^{-1}.

Bed Area

The flow rate of air required to burn char and part of product gas is calculated. Bed area is calculated taking this flow rate and average fluidization velocity.

$$A_b = \frac{V_g}{U_g} \tag{C.6}$$

$$A_b = 0.02m^2$$

$$U_g = 4ms^{-1}$$

$$V_g = 0.07kg/s$$

Bed Diameter

Bed area is used to calculate the diameter of the bed which will be the diameter of the riser. $A_b = 0.02m^2$

$\pi \times r^2 = 0.02$

$r = 0.07m$

$D = 0.15m$

Air Flow Rate at Different Inlet

In the riser the ratio of bottom to total air is taken as 0.1, ratio of primary to total is 0.3 and ratio of secondary to total is 0.6. Less flow rate is given at the bottom to avoid dilution of product gas wit nitrogen as there is chance of air entering the gasifier if the flow rate is high at bottom. The flow rate is high at secondary as it has to lift the particles to the top of the riser.

C.1 Pressure Drop

Voidage at Different Height

Voidage at different height are calculated using the following formula and the results are tabulated.

$$\epsilon = \left[exp\left(-a\left(h_i - h_d\right)\right) * \left({}_d - \epsilon_\infty\right)\right] + \epsilon_\infty \tag{C.6}$$

$$a = \frac{0.88 - 420d_p}{(U - U_t)^2 D^{0.6}} \tag{C.7}$$

$$\epsilon_\infty = \frac{K_i}{\rho_s(U - U_t)} \tag{C.8}$$

$$K_i = \rho_g U_G \times 23.7exp\left(\frac{-5.4U_t}{U}\right) \tag{C.9}$$

Hydrostatic Pressure Drop

The pressure drop across the bed is due to hydrostatic pressure drop and that due frictional pressure drop. The hydrostatic pressure drop in the bed at different heights are calculated using the density of solids and the voidage at different heights. The value are tabulated.

$$\Delta P_{hyd} = (1 - \epsilon)\rho_s g h \tag{C.10}$$

Height	Gas velocity	ϵ_∞	a	k_i	voidage
0.10	0.59	-8.48E-09	2.01	2.55E-05	0.82
0.20	0.59	-8.48E-09	2.01	2.55E-05	0.67
0.30	1.98	1.15E-03	0.28	9.34	0.65
0.40	1.98	1.15E-03	0.28	9.34	0.63
0.50	1.98	1.15E-03	0.28	9.34	0.62
0.60	4.85	1.73E-03	0.11	22.52	0.64
0.70	4.85	1.73E-03	0.11	22.52	0.64
0.80	4.85	1.73E-03	0.11	22.52	0.63
0.90	4.85	1.73E-03	0.11	22.52	0.62
1.00	4.85	1.73E-03	0.11	22.52	0.62
1.10	3.78	1.73E-03	0.10	22.52	0.61
1.20	3.78	1.73E-03	0.10	22.52	0.61
1.30	3.78	1.73E-03	0.10	22.52	0.60
1.40	3.78	1.73E-03	0.10	22.52	0.60
1.50	3.78	1.73E-03	0.10	22.52	0.59
1.60	3.78	1.73E-03	0.10	22.52	0.58
1.70	3.78	1.73E-03	0.10	22.52	0.58
1.80	3.78	1.73E-03	0.10	22.52	0.57
1.90	3.78	1.73E-03	0.10	22.52	0.57
2.00	3.78	1.73E-03	0.10	22.52	0.56

Table C.1: Voidage

Density of sand 2650kg/m^3

Height	Voidage	Pressure Drop
0.10	0.82	467.75
0.20	0.67	1709.57
0.30	0.65	2706.31
0.40	0.63	3792.55
0.50	0.62	4964.60
0.60	0.64	5566.50
0.70	0.64	6618.56
0.80	0.63	7704.61
0.90	0.62	8824.12
1.00	0.62	9976.54
1.10	0.61	11039.70
1.20	0.61	12232.06
1.30	0.60	13453.84
1.40	0.60	14704.62
1.50	0.59	15983.94
1.60	0.58	17291.38
1.70	0.58	18626.50
1.80	0.57	19988.89
1.90	0.57	21378.12
2.00	0.56	22793.79

Table C.2: Hydrostatic Pressure Drop

Frictional Pressure Drop

Frictional pressure drop is calculated from the average particle density in the bed. Average bed density is calculated from the velocity and flowrate of air at different inlets.

$$\Delta P_{fr} = \frac{2 f_s \rho_{avg} U^2 h}{D} \tag{C.11}$$

$$f_s = 0.05/U_g \tag{C.12}$$

$$G_s = U_s \rho_s (1 - \epsilon) \tag{C.13}$$

$$G_g = \rho_g U_g \tag{C.14}$$

$$\rho_{avg} = \rho_g \left(\frac{G_s}{G_g}\right) \left(\frac{U_g}{U_s}\right) \tag{C.15}$$

Where U is air velocity at inlet, U_s is the velocity of solids in the bed which is equal to the the gas velocity in the bed, which is calculated using the following relation.

$$U_s = \frac{V_g}{A\epsilon} \tag{C.16}$$

The values of gas velocity, average density of bed at different heights and the frictional pressure drop are tabulated in table C.3.

Pressure Drop

Pressure drop across the bed is the difference between the hydrostatic pressure drop and frictional pressure drop. The values of pressure drop across the bed are tabulated in table C.4. The inlet pressure minus the drop through the bed which is 15572 Pa will give the pressure in the bed.

Gas velocity	fs	Gg	Gs	ρ_{avg}	ΔP_{fr}
0.59	0.09	0.51	512.63	872.23	63.32
0.59	0.09	0.51	512.63	872.23	126.64
1.98	0.03	3.19	1077.96	543.35	1369.39
1.98	0.03	3.19	1077.96	543.35	1825.86
1.98	0.03	3.19	1077.96	543.35	2282.32
4.85	0.01	4.37	2390.92	493.47	1911.64
4.85	0.01	4.37	2390.92	493.47	2230.25
4.85	0.01	4.37	2390.92	493.47	2548.85
4.85	0.01	4.37	2390.92	493.47	2867.46
4.85	0.01	4.37	2390.92	493.47	3186.07
3.77	0.01	4.37	1861.44	493.47	3971.96
3.77	0.01	4.37	1861.44	493.47	4333.05
3.77	0.01	4.37	1861.44	493.47	4694.14
3.77	0.01	4.37	1861.44	493.47	5055.23
3.77	0.01	4.37	1861.44	493.47	5416.31
3.77	0.01	4.37	1861.44	493.47	5777.40
3.77	0.01	4.37	1861.44	493.47	6138.49
3.77	0.01	4.37	1861.44	493.47	6499.58
3.77	0.01	4.37	1861.44	493.47	6860.66
3.77	0.01	4.37	1861.44	493.47	7221.75

Table C.3: Frictional pressure drop

ΔP_{hyd}	ΔP_{fr}	ΔP_{riser}
467.75	63.32	404.43
1709.57	126.64	1582.93
2706.31	1369.39	1336.92
3792.55	1825.86	1966.69
4964.60	2282.32	2682.27
5566.50	1911.64	3654.86
6618.56	2230.25	4388.31
7704.61	2548.85	5155.76
8824.12	2867.46	5956.66
9976.54	3186.07	6790.47
11039.70	3971.96	7067.74
12232.06	4333.05	7899.01
13453.84	4694.14	8759.71
14704.62	5055.23	9649.39
15983.94	5416.31	10567.63
17291.38	5777.40	11513.98
18626.50	6138.49	12488.01
19988.89	6499.58	13489.31
21378.12	6860.66	14517.46
22793.79	7221.75	15572.04

Table C.4: Riser Pressure drop

Appendix D

Pressure Balance

D.1 Pressure Drop in Connection Chute

The connection chute is a tube that connects the gasifier with the riser. The bed materials flow through the connection as moving bed at minimum voidage condition. The expression for pressure drop is given in equation D.1

$$\Delta P_{chute} = \left[150\frac{(1-\epsilon_{mf})^2}{\epsilon_{mf}^2}\left(\frac{\mu\Delta u}{(\phi_s d_p)^2} + 1.75\left(\frac{1-\epsilon_{mf}}{\epsilon_{mf}}\right)\left(\frac{\rho_g(\Delta u)^2}{\phi_s d_p}\right)\right]Lsin\theta \quad (D.1)$$

$$\Delta u = u_g + u_s \quad (D.2)$$

The velocity of solids is calculated as 0.475 ms^{-1}. The pressure drop across the connection of length 0.6 m and dia of 0.07m is calculated as 8373 Pa. The connection is inclined at an angle of 35^o.

D.2 Pressure Drop In Cyclone

The pressure drop across the cyclone is calculated as 273 Pa.

D.3 Seal Loop

Seal loop consists of a standpipe, horizontal and a downcomer to balance the pressure between the riser and the gasifier. The pressure drop in each section are calculated as shown. The standpipe, horizontal and downcomer are maintained at minimum fluidised condition using air at syphon. The height is adjusted such that the doncomer and the bed are at same height. The pressure drop across the standpipe of length 0.4 m is found as 4807 Pa. The pressure drop across the horizontal of length 0.15m and dia 0.06m is found to be 261 Pa. The pressure drop across the downcomer of height 0.05m is calculated as 1117 Pa. The pressure drop across the seal loop should be equal to the difference between the outlet pressure in riser and gasifier.

$$P_{riser} - P_{gasifier} = \Delta P_{SLsta} - \Delta P_{SLdow} - \Delta P_{SLhor} \tag{D.3}$$

$$\begin{aligned} P_{riser} &= 121KPa \\ P_{gasifier} &= 118KPa \\ P_{riser} - P_{gasifier} &= 3429Pa \\ \Delta P_{SLsta} &= 4807Pa \\ \Delta P_{SLdow} &= 1117Pa \\ \Delta P_{SLhor} &= 261Pa \\ \Delta P_{SLsta} - \Delta P_{SLdow} - \Delta P_{SLhor} &= 3429Pa \end{aligned}$$

D.4 Pressure Balance Across The System

The difference in operating pressure between the riser and gasifier should be equal to sum of gasifier pressure drop and connection pressure drop minus riser pressure drop and cyclone pressure drop.

$$P_{riser} - P_{gasifier} = \Delta P_{gas} + \Delta P_{con} - \Delta P_{Riser} - \Delta P_{cyc} \tag{D.4}$$

$$\Delta P_{gas} = 10901Pa$$

$$\Delta P_{con} = 8373Pa$$

$$\Delta P_{riser} = 15572Pa$$

$$\Delta P_{cyc} = 273Pa$$

$$\Delta P_{gas} + \Delta P_{con} - \Delta P_{Riser} - \Delta P_{cyc} = 3429Pa$$