



Conversion of Saw Dust to Natural Gas Using Fluidized Bed Reactor

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Abstract- There is a tremendous need for Trinidad and Tobago to diversify its economy from dependent on natural gas. As the natural gas demand increases and the supply decreases many plants are forced to take drastic measures to remain profitable. The need for a cost effective, practical solution for the replacement of natural gas is required for the survival of Trinidad and Tobago's energy sector and by extension the country's economy.

This study designs a fluidized bed gasifier for the conversion of biomass to natural gas. The biomass used was sawdust (cedar wood) with the gasifying agent being air and the bed material being silica sand. Three design concepts were developed with one being selected to manufacture. The selected design utilised a direct combustion heat source to provide the required heat input for the process. The model developed produced an average of 5.33% hydrocarbon content of the product gas generated. This was achieved at an average bed temperature of 636°C. The cold gas efficiency of the reactor was calculated to be 12.96%. The thermal efficiency of the gasifier was calculated to be 74.84%.

Keywords- Fluidized Bed, Gasifier, Bed Temperature, Cold Gas Efficiency, Biomass

I. INTRODUCTION

There has been a steady decline in the supply of natural gas to downstream petrochemical industries in Trinidad and Tobago, production of ammonia decreased slightly by 1.5 percent in the 2017/2018 period, with output dipping from 3,739.3 thousand metric tonnes to 3,681.9 thousand metric tonnes. The fall in production was precipitated by a temporary production stoppage at the Caribbean Nitrogen Company's (CNC) plant in January 2018, consequent to a breakdown in negotiations over a new natural gas supply contract [1]. This problem has persisted for the past few years and resulted in industries being shut down and workers sent home.

Trinidad is very dependent on natural gas and any shortfall in production would result in a poor state for the country's economy. A practical way to solve Trinidad's dependency on natural gas is the use of gas produced from biomass. The

biogas can be used as a replacement for natural gas so many industries would not require major changes to their infrastructure to accommodate the change. One of the most efficient methods for biomass gasification is the use of fluidized bed reactors which has numerous advantages over traditional biomass gasification methods. The conversion of biomass to natural gas through the gasification process could be used as an alternative energy source to fill the future's energy demand gap [2, 3]. In addition, it would contribute to handling biomass waste effectively. It is projected that in the future, exploitation of different sources such as agricultural waste and forestry waste will be a significant contributor to the substitution of fossil fuel [3-5].

Biomass is categorized as all living and biological matter originating from Earth and is the most abundant carbon containing source on the planet [6]. Renewable energy is any energy resource that is renewed by nature and whose supply is not affected by the rate of consumption [7]. Biomass can be converted to energy through two different processes, biochemical or thermochemical. The major components of biomass are carbon, hydrogen, oxygen and moisture. Theoretically any biomass can be converted thermochemically to energy but biomass with moisture greater than 50% is not practically feasible [8].

This study will design and construct a fluidized bed reactor by utilizing the thermochemical process to produce natural gas in Trinidad and Tobago.

II. BIOMASS MATERIAL

Biomass can deliver nearly everything that fossil fuels provide, whether fuel or chemical feedstock. Additionally, it provides two important benefits that make it a viable feedstock for syngas production. First, it does not make any net contribution to the atmosphere when burnt; second, its use reduces dependence on non-renewables and often imported fossil fuel.

As shown in Table 1 different fuels have different energy densities. There are four types of Biomass that are usually considered for the process of gasification, the forms of biomass are:

Agricultural residues are generated after each harvesting cycle of commodity crops. A portion of the remaining stalks and biomass material left on the ground can be collected and used for conversion purposes.

Energy crops are produced solely or primarily for use as feedstock's in energy generation processes. Energy crops includes hybrid poplar, hybrid willow, and switch grass

Forestry residues are the biomass material remaining in forests that have been harvested for timber. Timber harvesting operations do not extract all biomass material, because only timber of certain quality is usable in processing facilities. Therefore, the residual material after a timber harvest is potentially available for energy generation purposes. Forestry residues are composed of logging residues, rough rotten salvageable dead wood, and excess small pole trees.

Urban wood waste/mill residues are waste woods from manufacturing operations that would otherwise be landfilled. The urban wood waste/mill residue category includes primary mill residues and urban wood such as pallets, construction waste, and demolition debris, which are not otherwise used.

TABLE I. C/H RATIO OF DIFFERENT FUELS [8]

Fuel	C/H mass ratio	Oxygen %	Energy density GJ/t
Anthracite	44	2.3	27.6
Bituminous coal	15	7.8	29
Lignite	10	11	9
Peat	10	35	7
Crude oil	9		42 (mineral oil)
Biomass/Cedar	7.6	40	20
Gasoline	6	0	46.8
Natural gas (CH ₄)	3	0	56
Syngas (CO:H ₂ in ratio 1:3)	2	0	24

A. Sawdust as a Biomass Material

A study conducted by the National Bioenergy Center investigated the use of varying biomass sizes in a fluidized bed reactor and the resulting products formed. The researchers used white oak as the biomass material and the particle diameters varied from 6, 13, 18 and 25 mm. It was discovered that as the particle size increases, a decrease in the formation of light gases, accompanied by an increase in the formation of char occurred. The gas produced consisted of CO, CO₂, CH₄ and H₂ [9].

B. Principles of Gasification of Biomass

Gasification of biomass occurs through the following steps:

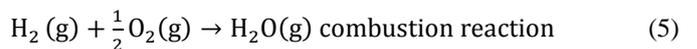
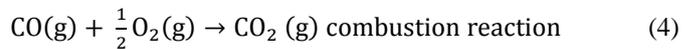
1) Drying:

Biomass contains a significant percentage of moisture and this moisture is evaporated in the drying process as represented in equation (1).



2) Pyrolysis and Combustion:

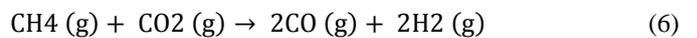
The removal of volatile substances, de-volatilization, of biomass results in the release of vapours mixtures. With additional heating of the biomass a flame is produced around the burning solid. This stage leads to char combustion which results in CO₂ release and ash formation as seen in equations (2, 3, 4 and 5).



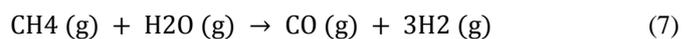
3) Gasification:

The solid char residue remaining after pyrolysis is converted into gas. Ash is also produced as a by-product in the process. However, this step will proceed only in an oxygen depleted environment. Most gasification reactions are endothermic (the water-gas shift reaction being an exception), therefore it is necessary to provide the required amount of heat to maintain the desired gasification temperature. The process is represented by equations (6, 7, 8, 9, 10 and 11).

Dry reforming reaction (Endothermic)



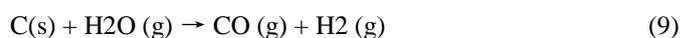
Wet reforming reaction (Endothermic)



Homogeneous water-gas shift reaction (Exothermic)



Heterogeneous water-gas reaction (Endothermic)



Boudouard reaction (Endothermic)



Methanation (hydrogasification) reaction (Exothermic)



Exothermic reactions through partial combustion in biomass gasification provide energy to the endothermic reactions and a steady state can be achieved where the gasifier maintains the gasification operation at certain required operating temperatures [10].

C. Product Gas Composition and Factors Affecting Gasification Products

The following are lists of expected products of the gasification of sawdust: Methane, Steam, Carbon Monoxide, Hydrogen and Carbon Dioxide. It is important to note that the nature of the products of biomass gasification in a fluidized bed reactor is influenced by many factors.

The following are the known factors that affect the nature of the products: type of biomass used; gasifying agent used;

moisture content and size of biomass particles; type, size and weight of bed material particles; gas velocity of the gasifying agent; biomass to gasifying agent ratio (comparable to air to fuel ratio in thermodynamics); presence of catalysts and amount of char in the bed.

D. Fluidized bed reactors

Fritz Winkler of Germany introduced gaseous products of combustion into the bottom of a crucible containing coke particles, creating the first demonstration of gasification of coal in a fluidized bed [11].

Gasification is perhaps the earliest commercial application of fluidized beds. The major motivation for use of this type of gas–solid contacting process is its excellent solid mixing capability. Crushed coal is fed into a bubbling fluidized bed of hot solids at 700°C to 950°C as shown in Figure 1. Steam, the major gasifying medium, is fed into the base of the fluidized bed through a sparger pipe-type of distributor. This fluidizes the raw coal along hot solids in the bed. Gasification products leave the bed from the top. The gas is cleaned and used. The burning of the char in the lower section of the bed provides the heat required for endothermic gasification reactions [12]. A major future role is envisaged for electricity production from biomass

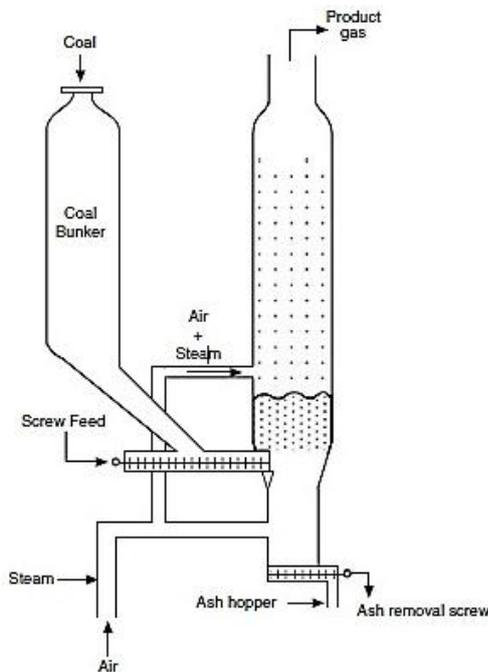
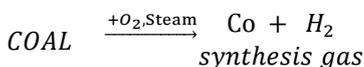


Figure 1. Schematic of gasifier [13].

Winkler's coal gasifier represents the first large-scale, commercially significant use of the fluidized bed [13]. The desired reaction, simply represented, is as follows



There are two main types of FBR's:

1) *Bubbling Fluidized Bed BFB:*

This type of reactor consists of a fluidizing grate through which the gasifying agent (gaseous form) is passed. The gas velocity is increased to a point where the drag force created is equal to the weight of the bed particles. This causes the bed material to develop bubbles thereby turning the bed material into a fluid state. The critical defining parameter of the BFB is the velocity at which the gasifying agent enters the fluidizing grate. Typical gas velocity values are 0.5-2 ms⁻¹ [11]. During the gasifying process, the bed material mostly stays in the reactor.

2) *Circulating Fluidized Bed CFB:*

The major difference between CFB's and BFB's is the velocity at which the air enters the reactor. CFB's use higher air velocities (typically 3-10 ms⁻¹). This increase in air velocity causes a large amount of the bed material to be displaced from the reactor in contrast to BFB's. The displaced bed material is mostly entrained in the resulting syngas and separation techniques are used to recover the bed material from the gas and circulate it back to the reactor bed.

III. DESIGN OF THE FLUIDIZED BED REACTOR

The schematic diagram of the fluidized bed reactor is shown in Figure 2. It utilizes a propane burner to supply the external heat required for the gasification process. The bed is heated directly by heating the bed material through an opening in the reactor body. Reactor temperature control is also challenging with this design.

The following is a list of important design specifications that will be considered for the design of this project;

- (a) The desired heating value of the product gas. Air will be used as the gasification medium for this project and the typical range of heating value is 3.7 MJ/Nm³ (Table 2)
- (b) The desired gas composition
- (c) The required power output of the gasifier

TABLE II. COMPARISON OF THE COMPOSITION OF PRODUCT GAS GENERATED FROM WOOD GASIFICATION FOR DIFFERENT GASIFIERS [14].

Property	Downdraft	Updraft	BFB	CFB
Tar (mg/Nm ³)	10-6000	10000-150000	Not defined	2000-30000
Particles (mg/Nm ³)	100-8000	100-3000	Not defined	8000-100000
LHV (MJ/Nm ³)	4.0-5.6	3.7-5.1	3.7-8.4	3.6-5.9
H ₂ (vol%)	15-21	10-14	5-16.3	15-22
CO (vol%)	10-22	15-20	9.9-22.4	13-15
CO ₂ (vol%)	11-13	8-10	9-19.4	13-15
CH ₄ (vol%)	1-5	2-3	2.2-6.2	2-4
CnHm (vol%)	0.5-2	Not defined	0.2-3.3	0.1-1.2
N ₂ (vol%)	Remaining	Remaining	41.6-61.6	Remaining

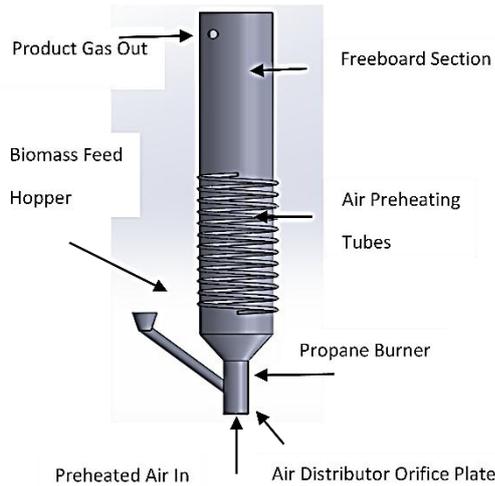


Figure 2. Showing design concept 3

A. Values for Design Parameters

1) Operating Temperature

Biomass from plants used as fuel in a BFB gasifier can be classified into two major groups;

Agricultural wastes

Ligno-cellulosic biomass. This is the non-starch fibrous part of plants.

Agricultural biomass gasifies at a temperature of 500°C but ligno-cellulosic biomass gasifies at temperatures in the range of 500-900 °C. Sawdust was selected as the biomass and this is classified as a ligno-cellulosic type biomass. Therefore, a gasification operating temperature of 600°C was selected. Another important factor in deciding the gasification temperature was the cost and availability of materials required for the manufacture of the prototype.

2) Retention Time

This is the time the biomass is given to undergo the gasification process. The value selected influences the height of the BFB gasifier. A longer residence time translates to a larger gasifier.

3) Equivalence Ratio

Equivalence Ratio is the actual air fuel ratio to the stoichiometric air fuel ratio. For the gasification of biomass, the equivalence ratio lies between 0.2-0.3 [15].

4) Fluidization Velocity

The range of fluidizing velocity in a BFB gasifier depends on the mean size of the bed material particles. The range selected should be between the minimum fluidization velocity and the terminal velocity of the mean particle size. Particles are grouped in Geldart's powder classification and provides a range for the fluidization velocity.

The typical fluidization velocity of silica sand of mean diameter 1 mm is 1-2 m/s [15]. Larger particles require a higher fluidization velocity. The following assumptions were

made to produce a practical design and are paramount to the design:

The air flow through the bed is assumed to be well distributed

The reactor is not pressurized and operates at a pressure slightly higher than atmospheric pressure.

Table 3 shows the parameters used in the design of the BFB gasifier:

TABLE III. DESIGN PARAMETERS

Parameter	Symbol	Value	Comments
Retention Time	R_T	10 seconds	This value selected reduces the amount of tar being formed in the products [8, 20].
Bed Temperature	T_B	600°C	Ligno-cellulosic biomass (sawdust) prefers a gasification temperature of 800°C [15].
Outlet Temperature of the reactor	T_o	550°C	This value is based on the EMCM model of a BFB reactor that caters for heat loss through reactor [16].
Minimum Fluidization Voidage of the Bed	emf	0.7	Typical values range from 0.5-0.85 [15].
Average Particle Diameter of Bed Material (Silica Sand)	d_p	0.414 mm	[16]
Bed Fluidization Particle Density (Silica Sand)	p_p	2600 kg/m ³	[16]
Fluidization Velocity	U_g	1 m/s	The typical fluidization velocity for silica sand vary between 1.0 and 2.0 m/s [15].
Power Output	Q	7 KW	Value selected based on previous bench scale models designed
Lower Heating Value of the product gas	LHV_g	3.5 MJ/m ³	For BFB gasifier using air as the gasifying agent typical range is 3.5 to 7 MJ/Nm ³ [15].
Higher Heating Value of the biomass (wet basis)	HHV_d	20502 KJ/kg	Higher Heating Value of sawdust [15]

B. Process Parameters Design Calculations

The design of a BFB gasifier involves both process and hardware calculations. Process design is necessary since it gives the type and yield of product and it also gives important parameters required for the design of hardware i.e. process calculations give an estimate of values that can be used to size the gasifier.

1) Determination of Product Gas Flow Rate

The volume flowrate of the product gas, V_g , is given by the following equation:

$$V_g = \frac{Q}{LHV_g} \quad (12)$$

NOTE: V_g considers the gas produced by the biomass and the volume of air used in the gasification process.

$$V_g = \frac{7}{3500} = 0.002 \text{ m}^3/\text{s}$$

2) Determination of Biomass Feed Rate

The biomass feed rate, M_f , is given by the following equation:

$$M_f = \frac{Q}{LHV_{bm}\eta_{gas}} \quad (13)$$

The LHV_{bm} is given by the following equation:

$$LHV_{bm} = HHV_{daf} - (20300 \times H_{daf}) - (2260 \times M_{daf}) \quad (14)$$

Where,

H_{daf} is the hydrogen mass fraction in the fuel

M_{daf} is the moisture mass fraction in the fuel

HHV_{daf} is the higher heating value of the biomass on a moisture and ash free basis

$$HHV_{daf} = \left(HHV_d \times \frac{1-M}{1-ASH-M} \right) \quad (15)$$

where $HHV_d = 20502 \text{ KJ/kg}$

M = the moisture fraction on a raw fuel basis

ASH = the ash fraction on a raw fuel basis

$$HHV_{daf} = \left(20502 \times \frac{1-0.3}{1-0.01-0.3} \right) = 20799 \text{ KJ/kg}$$

$$LHV_{bm} = 20799 - (20300 \times 0.065) - (2260 \times 0.3) = 18802 \text{ KJ/kg}$$

Designing for an efficiency of 100%

$$M_f = \frac{7}{18802 \times 1} = 0.4 \text{ g/s}$$

3) Determination of Stoichiometric Air Required for Combustion

The dry air required for complete combustion of a unit weight of dry hydrocarbon, M_{da} , is given by:

$$M_{da} = \left[0.1153C + 0.3434 \left(H - \frac{O}{8} \right) + 0.0434S \right] \text{ kg/ kg of dry fuel} \quad (16)$$

Where,

C = % of carbon on a dry basis = 47.2

H = % of hydrogen on a dry basis = 6.5

O = % of oxygen on a dry basis = 45.4

S = % of Sulphur on a dry basis = 0

Substituting the values of C , H , O and S into the equation gives

$$M_{da} = 5.725 \text{ kg/ kg of dry fuel}$$

4) Determination of Flow Rate of Gasifying Agent (Air)

In the previous section we calculated the stoichiometric air required for combustion, M_{da} . However, for an air blown gasifier the actual amount of air required for gasification is given by:

$$M_{TH} = M_{da} \times ER \quad (17)$$

where ER is the equivalence ratio = 0.25

$$\text{Equivalence Ratio, } ER = \frac{\text{Actual Air Required}}{\text{Stoichiometric Air}} \quad (18)$$

For gasification $ER < 1$ and for combustion $ER > 1$

Using the values of ER and M_{da}

$$M_{TH} = 5.725 \times 0.25 = 1.43 \text{ kg/ kg of dry fuel}$$

C. Hardware Design

1) Determination of Minimum Inside Cross-Sectional Area of Gasifier and Minimum Inside Diameter

The inside cross-sectional area of the gasifier, A_b is given by the following equation:

$$A_b = \frac{V_g}{U_g} \quad (19)$$

where $V_g = 0.002 \text{ Nm}^3/\text{s}$

and $U_g = 1 \text{ m/s}$

Therefore,

$$\text{Bed Area, } A_B = \frac{0.002}{1} = 0.002 \text{ m}^2 = 2000 \text{ mm}^2$$

$$\text{Diameter of bed, } D_B = \sqrt{\frac{4A_b}{\pi}} = \sqrt{\frac{4 \times 2000}{\pi}} = 50.4 \text{ mm} \\ \approx 2 \text{ inches}$$

2) Determination of Reactor Bed Volume and Bed Height

The bed volume is given by the following equation:

$$V_{bed} = \frac{\dot{C} \times R_T}{\rho_p \times x_{char} \times (1-ef)} \quad (20)$$

Where,

\dot{C} is the char feed rate entering the bed = $0.47 \times 5 \times 10^{-4} \text{ kg/s}$

x_{char} is the fraction of reacting char in the bed = $4\% = 0.04$

$$V_{bed} = \frac{0.47 \times 5 \times 10^{-4} \times 10}{2600 \times 0.04 \times (1 - 0.7)} = 7.53 \times 10^{-5} \text{ m}^3$$

$$\text{Bed Height, } H_B = \frac{V_{bed}}{A_b} = \frac{7.53 \times 10^{-5}}{0.002} = 38 \text{ mm}$$

3) Determination of Reactor Volume and Freeboard Height of Gasifier

The overall volume of the reactor, V_r , is given by the following equation:

$$V_r = V_g \times R_T \\ V_r = 0.0003 \times 10 = 0.003 \text{ m}^3 \quad (21)$$

$\text{Freeboard Volume} = \text{Reactor Volume} - \text{Bed Volume}$

$$\text{Freeboard Volume} = 0.003 - 7.53 \times 10^{-5} = 2.992 \times 10^{-3} \text{ m}^3$$

Using a freeboard diameter of 4 inches (0.1016 m) we can get the freeboard height by the following equation:

$$\text{Freeboard Height} = \frac{\text{Freeboard Volume}}{\text{Freeboard Area}} \quad (22)$$

$$\text{Freeboard Height} = \frac{2.992 \times 10^{-3}}{\frac{\pi \times 0.1016^2}{4}} = 0.369 = 15 \text{ inches}$$

4) Design of Air Distributor

There are many different types of distributor designs for a BFB gasifier. The distributor type chosen for this design was based on ease of manufacture and cost to manufacture. The distributor type selected was a single plate orifice type distributor.

a) Determination of Bed Pressure Drop and Pressure Drop across the Distributor Plate

The first step in the design of the air distributor is the calculation of the bed pressure drop.

The bed pressure drop is given by:

$$\Delta P_B = \rho_p \times (1 - \epsilon) \times H_B \times g \quad (23)$$

Where,

Density of silica sand, $\rho_p = 2600 \text{ kg/m}^3$

Height of the bed, $H_B = 0.025 \text{ m}$

Gravity, $g = 9.82 \text{ m/s}^2$

Bed voidage, ϵ , is given by:

$$\epsilon = \frac{U_g + 1}{U_g + 2} \quad (24)$$

$$\epsilon = \frac{1+1}{1+2} = 0.67$$

$$\therefore \Delta P_B = 2600 \times (1 - 0.67) \times 0.025 \times 9.82 = 211 \text{ Pa}$$

Now the pressure drop across the distributor plate is given by:

$$\frac{\Delta P_D}{\Delta P_B} = 0.15 \rightarrow 0.30 \quad (25)$$

Therefore,

$$\Delta P_D = 0.3 \times 211 = 64 \text{ Pa}$$

b) Determination of The Air Velocity Through the Grid Holes

The air velocity, U_o , through an orifice is given by:

$$U_o = C_D \left[\frac{2\Delta P_D}{\rho_{air}} \right]^{0.5} \quad (26)$$

Where,

ρ_{air} = density of air at $600^\circ\text{C} = 0.4 \text{ kg/m}^3$

$$\therefore U_o = 0.8 \left[\frac{2 \times 64}{0.4} \right]^{0.5} = 14.31 \frac{\text{m}}{\text{s}}$$

c) Determination of Air Flow Rate Below the Distributor Plate

The air flow rate below the plate is given by:

$$V_g = U_g \times \frac{\pi}{4} B_D^2 \quad (27)$$

$$V_g = 1 \times \frac{\pi}{4} \times 0.05^2 = 0.002 \text{ m}^3/\text{s}$$

d) Determination of the Number of Grid Holes Required on the Plate

The number of grid holes required, N is given by the following equation:

$$N = \frac{4 \times V_g}{U_o \times \pi \times d_h^2} \quad (28)$$

Where, d_h is the diameter of an individual orifice on the plate

The hole density, N_D , is given by:

$$N_D = \frac{4N}{\pi B_D^2} \quad (29)$$

Substituting for N

$$N_D = \frac{4}{\pi B_D^2} \times \frac{4 \times V_g}{U_o \times \pi \times d_h^2} = \frac{4}{\pi \cdot 0.05^2} \times \frac{4 \times 0.002}{14.31 \times \pi \times d_h^2} = \frac{0.09063}{d_h^2}$$

Using a triangular pitch as shown in Figure 3:

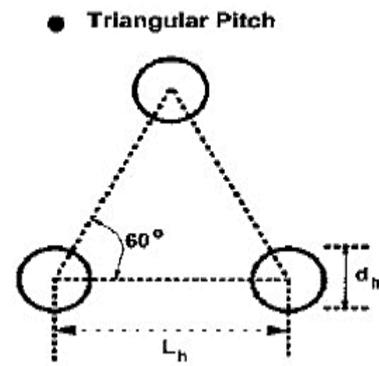


Figure 3. Triangular Pitch Arrangement [17]

Where,

$$L_h = \frac{1}{\sqrt{N_D \sin 60}}$$

Using a hole diameter, d_h of 2 mm we get the following combination as seen in Table 4.

TABLE IV. COMBINATIONS FOR A HOLE DIAMETER OF 2 MM

Hole Diameter, d_h /mm	Number of holes, N	Hole Density, N_D per m ²	L_h / mm
2	44	22658	7

Therefore, the 2-inch plate type air distributor will have 44, 2mm holes drilled 7mm apart.

D. Determination of Heat Input Required to Achieve Desired Bed Temperature

The operating range for the gasification process is $500^\circ\text{C} - 900^\circ\text{C}$ and the maximum service temperature of the material chosen for the reactor construction, 316 Stainless Steel, is 1696.73°F (924.85°C) (SSINA 2018) as shown in Figure 3. For this design, the maximum temperature seen by the stainless-

steel material will be 900°C. The reactor consists of two major pipe sections;

1. A two-inch diameter stainless steel pipe that is four inches in length (Bed Reactor section)
2. A four-inch diameter stainless steel pipe that is fifteen inches in length (Freeboard Section)

Due to the availability of material a 5000W Nichrome coiled wire will be used to achieve the desired bed and freeboard temperatures. The Nichrome coils are 100 cm long and has an outer diameter of 7.7mm. The Nichrome coils are purely resistive elements and cannot be wrapped directly on the pipe. This would result in current flowing through the reactor creating a major electrical shock hazard. Therefore, a ceramic sleeve will be inserted over the pipes. The ceramic material will provide a means of electrically insulating the pipe and allow good heat dissipation since ceramic has a relatively good thermal conductivity as an insulator. One coil will be used for each pipe.

Fourier's Law for temperature distribution in a cylinder is given by:

$$q_r = \frac{2\pi Lk(T_{s,1}-T_{s,2})}{\ln\frac{r_2}{r_1}} \quad (30)$$

Where,

q_r is the rate of heat transfer=5000W (Assuming no heat losses to the insulation)

The thermal conductivity of alumina ceramic ranges 12-38 W/mK [18]. For our purposes we will use the minimum value as a worst-case scenario

k is the thermal conductivity of ceramic (alumina)=12 W/mK

r_2 and r_1 are the outer and inner radii of the ceramic sleeve respectively

$T_{s,1}$ is the outer surface temperature of the ceramic sleeve

Assuming a temperature difference of 50°C between the nichrome coil and the ceramic surface due to the contact resistance, $T_{s,1} = 950^\circ\text{C}$

$T_{s,2}$ is the inner surface temperature of the ceramic sleeve

L is the length of the ceramic sleeve

For the ceramic sleeve for the 2-inch Schedule 40 stainless steel pipe

The ceramic sleeve has the same OD as the pipe and a wall thickness of 0.125"

$$r_2 = 2.5" = 0.0333375 \text{ m}$$

Solving for $T_{s,2}$ we get

$$T_{s,2} = 950 - \frac{5000 \times \ln\frac{0.0333375}{0.0301625}}{2\pi \times 0.1016 \times 12} = 884.7^\circ\text{C}$$

$$T_{s,2} = 884.7^\circ\text{C}$$

We can now solve for the inside temperature of the stainless-steel pipe

Assuming zero contact resistance between the ceramic sleeve we get that

$$T_{s,1} \text{ is the outer surface temperature of the 2" pipe} \\ = 884.7^\circ\text{C}$$

Using Fourier's equation

For the 2-inch Schedule 40 stainless steel pipe

$$r_2 = 3 \text{ cm and } r_1 = 2.63 \text{ cm}$$

Solving for $T_{s,2}$ we get

$$T_{s,2} = 884.7 - \frac{5000 \times \ln\frac{0.03}{0.0263}}{2\pi \times 0.1016 \times 15} = 816^\circ\text{C}$$

We get $T_{s,2} = 816^\circ\text{C}$ which is acceptable

E. Determination of Heat Input Required to Preheat Gasifying Agent (Air)

The air entering the grate must be preheated close to the bed temperature. Failure to do so will cause heat to be transferred to the air and result in a lower bed temperature where the gasification process will not be able to take place.

The air needs to be heated from 40°C to 600°C. This would require the design of an air preheating chamber.

The flowrate of air needed to be supplied to the gasifier, $V_g = 0.002 \text{ m}^3/\text{s}$. Air exits the compressor at 40°C.

Mass flow rate of air entering the air preheat chamber:

$$M_{air,preheat} = \rho_{air @ 40^\circ\text{C}} \times V_g$$

$$M_{air,preheat} = \rho_{air @ 40^\circ\text{C}} \times V_g = 1.127 \text{ kg/m}^3 \times 0.002 \text{ m}^3/\text{s} = 0.002254 \text{ kg/s}$$

The average air temperature inside the air preheat chamber

$$T_{average} = \frac{40+600}{2} = 320^\circ\text{C}$$

Heat required to preheat air from 40°C to 600°C:

$$q = M_{air,preheat} \times C_{p \text{ air @ } 320^\circ\text{C}} \times \Delta T \quad (32)$$

$$q = 0.002254 \times 1049.518 \times 600 - 40 = 1325 \text{ W}$$

F. Estimation of Heat Transfer Coefficient and Determination of Preheater Dimensions

The air preheater chosen for the project was based mainly on availability of materials and cost. The air preheater shown in Figure 4, will be a 0.5-inch diameter tubing coated with a 0.125" thick alumina ceramic coating to which a nichrome coil will be tightly wrapped on the outside. The tubing has OD 0.5" and ID is 0.402". The ceramic sleeve has outside diameter of 0.625" and internal diameter of 0.5".

The average gas velocity in the preheat chamber, $U_{air \text{ preheat}}$, is given by:

$$U_{air \text{ preheat}} = \frac{\text{Air flowrate inside the tube}}{\text{Cross Sectional Area of tube}} = \frac{V_g}{A} \quad (33)$$

Where,

$$V_g = 0.002 \text{ m}^3/\text{s}$$

$$A = \frac{\pi \times 0.0102^2}{4} \quad (\text{for } 0.402'' \text{ ID tubing})$$

$$\therefore U_{\text{air preheat}} = \frac{0.002}{\frac{\pi \times 0.0102^2}{4}} = 24.48 \text{ m/s}$$

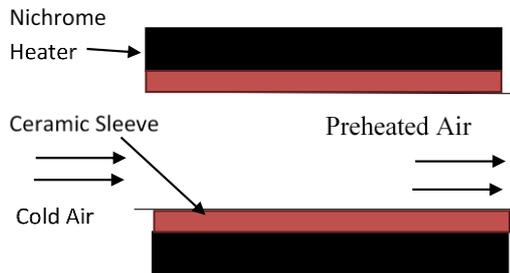


Figure 4. Inlet Air Tubing

Reynold's number of the gas flow at average air temperature:

$$Re = \frac{U_{\text{air preheat}} \times \rho_{\text{air}} \times D_{\text{preheater}}}{\mu_{\text{air}}} \quad (34)$$

Where,

density of air, $\rho_{\text{air}} = 0.5979 \text{ kg/m}^3$

dynamic viscosity of air, $\mu_{\text{air}} = 2.99348 \times 10^{-5} \text{ kg/ms}$

$$Re = \frac{24.48 \times 0.5979 \times 0.0102}{2.99348 \times 10^{-5}} = 4987$$

$Re > 2300 \therefore$ the flow is regarded as turbulent

Nusselt number of the flow is given by the following equation for turbulent flow:

$$Nu = 0.023 Re^{0.8} Pr^{0.4} \quad (35)$$

Where, $Pr = 0.68$

$$Nu = 0.023 \times 4987^{0.8} \times 0.68^{0.4} = 17.91$$

We can calculate the estimated heat transfer coefficient of the air flowing in the tube by the following equation:

$$Nu = \frac{hD}{k} \quad (36)$$

$$h = \frac{17.91 \times 0.0461844}{0.0127} = 81.09 \text{ W/m}^2\text{k}$$

The length of the tubing can be calculated by the following equation:

$$q = \frac{\Delta T}{Ah} \quad (37)$$

Where, $A = \pi DL$

Using an initial estimate of the inside temperature of the stainless-steel tubing as 800°C based on previous calculation

$$L = \frac{q}{\Delta T \times h \times \pi D} = \frac{1325}{(800-40) \times 81.09 \times \pi \times 0.0102} = 0.67 \text{ metres}$$

We can now calculate the correct value of the inside surface temperature of the stainless-steel tubing by first calculating the inside surface temperature of the ceramic sleeve.

Using Fourier's equation

$$q_r = \frac{2\pi Lk(T_{s,1} - T_{s,2})}{\ln \frac{r_2}{r_1}}$$

Where,

q_r is the rate of heat transfer = 5000 W (Assuming no heat losses to the insulation)

k is the thermal conductivity of alumina ceramic = 12 W/mK

$r_2 = 0.009525 \text{ m}$

$r_1 = 0.00635 \text{ m}$

$T_{s,1}$ is the outer surface temperature of the ceramic sleeve = 950°C

$T_{s,2}$ is the inner surface temperature of the ceramic sleeve

L is the length of the ceramic sleeve = 0.67 m

Solving for $T_{s,2}$ we get

$$T_{s,2} = 950 - \frac{5000 \times \ln \frac{0.009525}{0.00635}}{2\pi \times 0.67 \times 12} = 909.9^\circ\text{C}$$

We can now solve for the inside temperature of the stainless-steel tubing

Assuming zero contact resistance between the ceramic sleeve we get that

$T_{s,1}$ is the outer surface temperature of the stainless steel tubing = 909.9°C

Using Fourier's equation

For the stainless-steel tubing

$r_2 = 0.00635 \text{ m}$

$r_1 = 0.0051054 \text{ m}$

Solving for $T_{s,2}$ we get

$$T_{s,2} = 909.9 - \frac{5000 \times \ln \frac{0.00635}{0.0051054}}{2\pi \times 0.67 \times 15} = 892.6^\circ\text{C}$$

We get $T_{s,2} = 892.6^\circ\text{C}$ which is acceptable

G. Quantifying the amount of heat supplied to the reactor

The volume of acetylene used during the experiment is found by:

Step 1: Calculate the volume of acetylene, V_2 , at atmospheric pressure, P_2 , using the general gas equation:

$$\frac{P_1 V_1}{T_1} = \frac{P_2 V_2}{T_2}$$

Assuming a constant temperature

$$V_2 = \frac{P_1 V_1}{P_2}$$

where

V_1 is the volume of the cylinder = 0.076 m^3

P_1 is the pressure of the gas inside the cylinder

Initially $P_1 = 1740 \text{ psi} = 11997 \text{ KPa}$

$$\therefore V_2 = \frac{11997 \times 0.076}{101.3} = 9 \text{ m}^3$$

The initial volume of the acetylene gas at atmospheric pressure = 9 m^3

After the experiment was concluded, $P_1 = 1635 \text{ psi} = 11273 \text{ kPa}$

$$\therefore V_2 = \frac{11273 \times 0.076}{101.3} = 8.46 \text{ m}^3$$

The final volume of the acetylene gas at atmospheric pressure = 8.46 m^3

The volume of gas used = 0.54 m^3

The number of moles of acetylene is given by the ideal gas equation:

$$PV=nRT$$

Number of moles of acetylene combusted,

$$n = \frac{101300 \times 0.54}{8.342 \times 303} = 21.6 \text{ mole}$$

The energy released during the combustion of acetylene is 1216 KJ/mol

Assuming all the acetylene is combusted then the heat supplied = 26266 KJ

The duration of the experiment was approximately 1 hour therefore, the power supplied = $262666000/3600 = 7296 \text{ W}$

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IV. METHODOLOGY

The operating temperature of the gasifier was the most important factor in determining a suitable material for the reactor body shown in Figure 5.

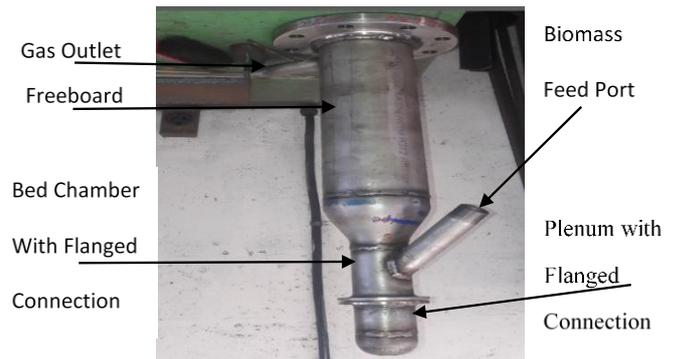
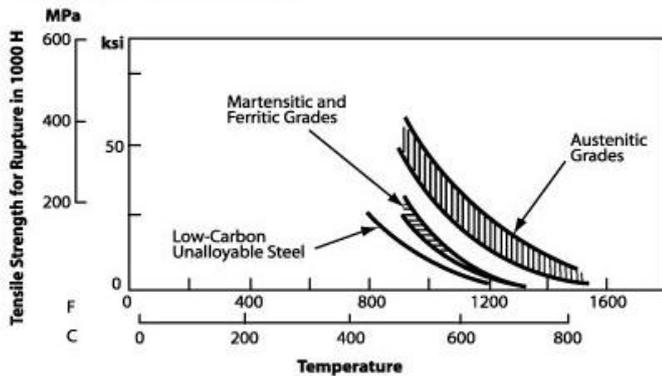


Figure 5. Partial Assembly of Prototype

Availability and cost of materials were also considered. 316 Stainless steel was chosen as the material for the reactor because of its high maximum operating temperature 927°C . At this temperature the material can still be operating within safe limits i.e. its strength characteristics are not adversely affected [19] as shown in Figure 6.



General comparison of the hot-strength characteristics of austenitic, martensitic and ferritic stainless steels with those of low-carbon unalloyed steel and semi-austenitic precipitation and transform ion-hardening steels.

Figure 6. Hot Strength Characteristics of Different Steel Alloys (Source: SSINA 2018).

A. Operating and Testing Procedure

Before the circuit is energized ensure that all electrical connections are secure and electrically insulated as shown in Figure 7.



Figure 7. Test Setup of Prototype

Check for any continuity issues with the coils and the reactor body using a multi-meter and address any problems as they arise. Once all the safety checks are done proceed with the operating/testing procedure as follows:

1) Pour the silica sand in the bed up to the fill mark indicated

- 2) Turn on the compressor and adjust the regulator to the desired pressure
- 3) Ensure that the bed is fluidizing uniformly at the pressure set by the regulator. Slight adjustments might need to be made to ensure the fluidization is uniform. Measure the velocity of the air by breaking the air hose tubing connection. Record the value, reconnect the joint and bolt the blind flange to the top reactor flange.

4) Calculate the flowrate of the air by using the following formula:

$$\text{Flowrate of air} = \text{Velocity of the air} \times \text{Inside cross sectional area of tubing}$$

$$\text{Flowrate of air} = \text{Velocity of the air} \times 8.171 \times 10^{-5} \text{ m}^3/\text{s}$$

- 5) Record the pressures of the gases in both the acetylene and oxygen tanks. Also record the ambient temperature and pressure.
- 6) Adjust the regulator pressure on the acetylene and oxygen and light the torch. This should be done by trained professionals to minimize the hazard potential
- 7) Apply heat to the exposed area of the reactor (bed chamber)
- 8) Monitor the outside surface temperature of the stainless-steel pipe so that the temperature does not exceed 900 °C. This should produce a bed temperature >600°C. Open the gate valve and add the required amount of sawdust based on the calculation below

$$\text{mass flow rate of biomass} =$$

$$\frac{\text{flowrate of air} \times \text{density of air at } 600^\circ\text{C}}{1.43}$$

$$\text{mass flow rate of biomass} = \frac{\text{flowrate of air} \times 0.4}{1.43} \text{ kg/s}$$

- 9) Using the gas analyser test the product gas by record readings at 1-minute intervals. For a batch process, add biomass at 1-minute intervals. The quantity of biomass to be added is calculated by multiplying the value in step 8 by 60 seconds.
- 10) Repeat step 9 until at least 10 readings are obtained.
- 11) Once a substantial methane/hydrocarbon level is shown on the gas analyser then light the Bunsen burner and observe the condition of the flame being produced.
- 12) Record all observations
- 13) After testing is completed, cordon off the area and allow the reactor to cool properly before attempting any maintenance work.

B. Results and Discussion

Table 5 shows the data that was collected and used in the analysis section of the study.

TABLE V. TEST DATA

Parameter	Value
Acetylene Cylinder pressure before experiment	1740 psi
Acetylene Cylinder pressure after experiment	1635 psi
Flowrate of the gasifying agent (air)	0.002 m ³ /s

The gas analyser used measured the following gases

- Carbon Monoxide, CO
- Carbon Dioxide, CO₂
- Oxygen, O₂
- Hydrocarbon, C_xH_y
- Nitrogen Monoxide, NO
- Air

Table 6 shows the composition of the product gas produced during the gasification process at the respective bed temperatures. Readings were taken at two-minute intervals. Figure 8 shows the percentage of hydrocarbons produced at the varying bed temperatures.

TABLE VI. PERCENTAGE COMPOSITION OF PRODUCT GAS PRODUCED FROM SAWDUST GASIFIER

Readings	Bed Temp°C	% O ₂	% CO	% CO ₂	% Hydrocarbon C _x H _y	% Air	% NO
1	626	7.5	2.93	7.4	2.58	79	0.59
2	627	8.4	9.55	8.5	3.98	69	0.57
3	643	10.5	15.20	10.7	6.17	57	0.43
4	652	12.1	17.63	12.4	8.47	49	0.40
5	635	9.7	13.54	9.0	5.34	62	0.42
6	637	9.7	14.33	8.9	5.64	61	0.43
7	633	9.7	10.92	9.9	5.09	63	0.44
8	635	9.6	13.52	9.1	5.33	62	0.42
9	637	9.7	14.29	9.2	5.63	61	0.42
10	632	9.3	12.97	9.2	5.09	63	0.44
Average	636	9.6	12.49	9.4	5.33	63	0.46

At an average bed temperature of 636°C during the gasification of sawdust the following gases were produced: 9.6 % of oxygen, 12.49 % of carbon monoxide, 9.4% of carbon dioxide, 5.33 % of hydrocarbon, 63 % of air and 0.46 % of Nitrogen oxide.

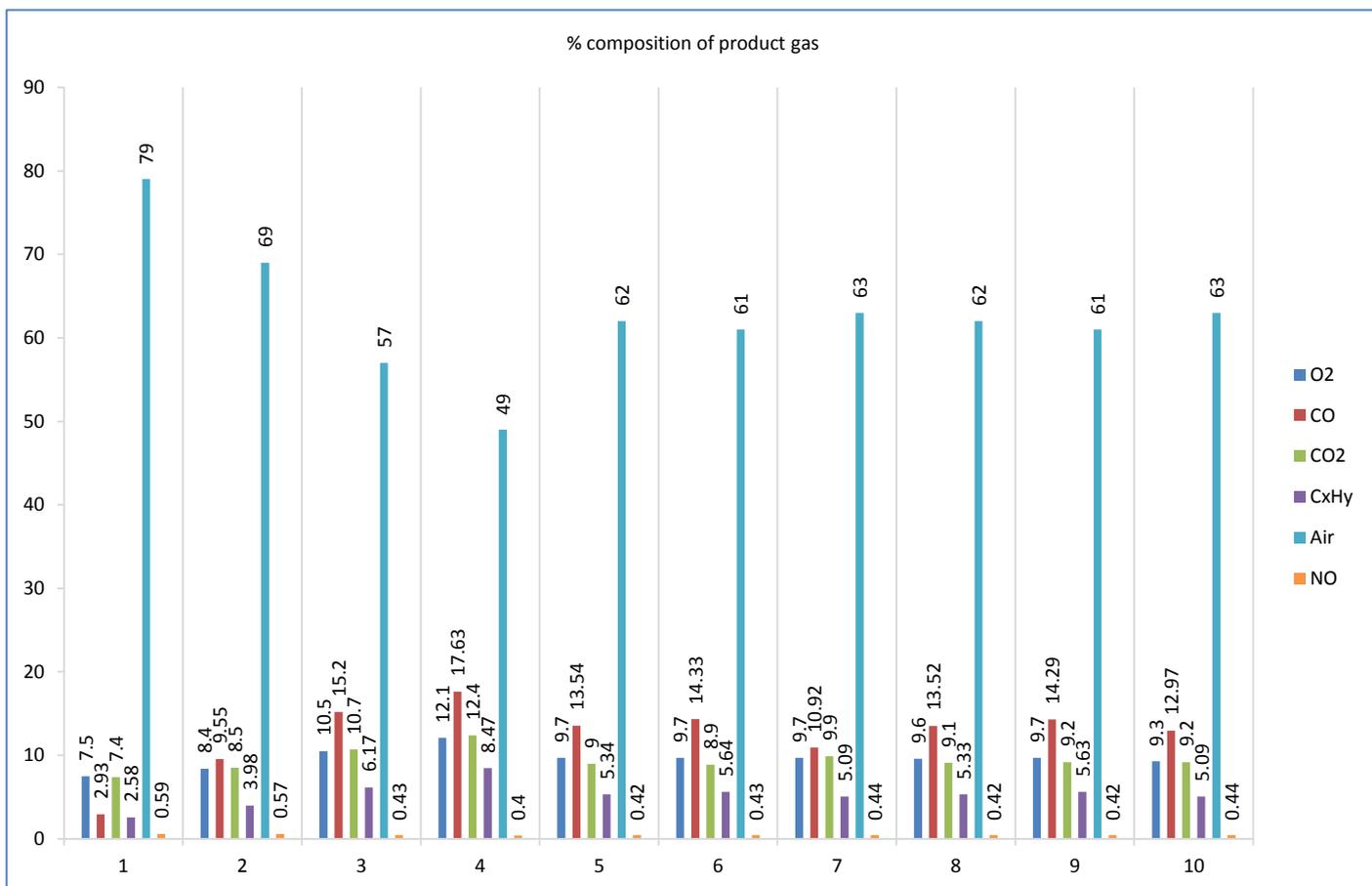


Figure 8. Percentage Composition of Product Gas

TABLE VII. LOWER HEATING VALUES FOR GASES FROM BIOMASS GASIFICATION [19]

Component	H2	CO	CH4	C2H6	C2H4	C2H2	Company	
HHV	12.745	12.633	39.819	70.293	63.414		TU Wien	
	12.769	12.622	39.781	69.693	63	58.059	NREL	
	12.753	12.626	39.721	69.595	62.952		ECN	
	12.766	12.641	39.847	70.402	63.998	58.975	DMT	
	12.761	12.634	39.747	69.636	62.989	58.039	Carbona	
	12.76	12.617	39.663	69.511	63.042	57.934	Univ. Sherbrock	
	12.758	12.631	39.739				Vattenfall	
	12.761	12.634	39.75	69.642	62.994	58.022	Nykomb	
	LHV	10.783	12.633	35.883	64.345	59.457		TU Wien
		10.788	12.622	35.814	64.748	59.036	56.078	NREL
10.789		12.626	35.796	63.704	59.024		ECN	
10.8		12.6	35.8	64.71		56.03	Verenum	
10.757		12.641	35.787	64.333	59.938	56.924	DMT	
10.8		12.6	35.8	64.71		56.03	Carbona	
10.793			35.81				Vattenfall	
10.8		12.634	35.823	63.756	59.07	56.06	Nykomb	
10.797		12.635	35.821	63.749	59.068		TPS	
10.789		12.630	35.812	64.744	59.033	56.088	Bioelectrica	

The cold gas efficiency of the gasifier is given by the following equation:

$$\text{Cold Gas Efficiency} = \frac{V_g q_g}{M_b C_b} \quad (38)$$

Where,

V_g is the gas generation rate of the product gas. For bench scale models the gas generation rate can be approximated to the flowrate of incoming air [15].

$$\therefore V_g = 0.002 \frac{m^3}{s}$$

q_g is the lower heating value of the product gas = 2730 KJ/m³

M_b is the rate of biomass addition = 5.6×10^{-4} kg/s

C_b is the heating value of the biomass = 75208 KJ/m³

$$\text{Cold Gas Efficiency} = \frac{0.002 \times 2730}{5.6 \times 10^{-4} \times 75208} = 12.96\%$$

The amount of heat available in the product gas = 0.002 X 2730 KW = 5460 W

The amount of heat supplied during the testing phase was 7296 W. See calculation in Appendix B.

Therefore, the thermal efficiency of the reactor,

$$= \frac{5460}{7296} \times 100 = 74.84\%$$

An important observation is that the percentage of hydrocarbons produced varied as the process endured. This could have been avoided by a constant supply of biomass according to the mass to air fuel ratio calculated before. An

important trend observed was an increase in bed temperature resulted in an increase in the percentage of hydrocarbons produced. The cold gas efficiency was calculated to be 12.96%. This efficiency represents the efficiency of the conversion of the biomass into gas.

As stated previously there are many factors that affect the products of gasification but from the results of the experiment the most apparent is bed temperature. This low value for cold gas efficiency obtained supports the theory that the optimum temperature for biomass gasification in a fluidized bed is 800°C. Another factor that could have contributed to the poor cold gas efficiency is the amount of moisture present in the sawdust. A notable observation during testing was that the bed temperatures dropped on average by 50°C when the sawdust was added.

The magnitude of the associated temperature drop could be reduced by drying the biomass beforehand. This would promote higher bed temperatures and a greater hydrocarbon (methane) production. Most commercial size fluidized bed gasifiers maintain a bed temperature of 800°C. The cold gas efficiency of these gasifiers is typically high (approximately 80%) [15]. The thermal efficiency of the gasifier considers the heat input entering the system and the heat leaving the system and this was calculated to be 74.84%. This value can be improved by minimizing heat loss to the environment through radiation and convection. However, it should be noted that the heating method from using the oxy-acetylene flame was more efficient than the heating coils since the overall power drawn from the coils would have been more than that supplied by the torch. Electricity is a high-quality energy source compared to that of the heat from combustion. Therefore, in terms of cost and savings, direct combustion heating is more feasible.

V. CONCLUSION

The primary pollutants released from the burning of fossil fuels are carbon monoxide (CO), carbon dioxide (CO₂), sulfur dioxide (SO₂), nitrogen oxides of the chemical form NO_x (primarily nitrogen dioxide, or NO₂), nitric oxide (N₂O), various hydrocarbons (methane, CH₄, being one such example) and substances collectively termed volatile organic compounds. Some of these are hazardous in their native forms; others are especially damaging only after they combine with other otherwise benign reagents in the atmosphere.

By far the most concerning and talked-about of these compounds is CO₂. Because carbon accounts from between 60 to 90 percent of the mass of the fossil fuels burned, CO₂ is the principal product of the combustion of fossil fuels worldwide.

This study theoretically and experimentally analysed a working bubbling fluidized bed gasifier using sawdust as the biomass, air as the gasifying agent and silica sand as the bed material was successfully carried out.

9.6 % of oxygen, 12.49 % of carbon monoxide, 9.4% of carbon dioxide, 5.33 % of hydrocarbon, 63 % of air and 0.46 % of Nitrogen oxide were produced from the experimentation.

The cold gas efficiency of the gasifier was calculated to be 12.96%. This figure represented the carbon conversion of the biomass into gas. Direct heating through combustion proved an adequate and better means of heating the gasifier compared to the electrical heating coils. However, the means of direct heating needs to be improved significantly to minimise heat wastage. Trinidad and Tobago is facing a natural gas shortage and gas produced from biomass can be a viable option. Biomass gasification can be beneficial in securing the sustainability of the country's energy sector and economy.

However, further work in this area needs to be done before biomass gasification can be considered a suitable alternative for natural gas.

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