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SUITABLE SYNGAS FOR FISCHER TROPSCH PRODUCED FROM BAMBOO CULMS IN A NEWLY DESIGNED DOUBLE OPPOSITE PIPE FLUIDIZED BED GASIFIER

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ABSTRACT

The spread of biomass in each nation makes it the world's largest and most sustainable renewable energy resources. The bamboo culms (BC) are abundantly available in Nigeria, but untapped. However, adequate preparation and transformation of biomass into synthesis gas in a suitably designed biomass gasifers was vigorously pursued. A bubbling fluidized bed gasifier (BFBG) was designed using fluidization dynamic principle and standard procedure in the literature. The gasifier was attached with two opposite extended torispherical heads inclined at angle of 60 ^oC to the reactor to ensure high carbon conversion of raw material by inducing high degree of turbulence and perfect mixing, The bamboo gasification results shows that operating with an equivalence ratio of 0.36 and bed temperature of 870℃ gives a corresponding syngas volumetric composition of H_2, CO_2, CO and CH_4 as 28.14%, 14.61%, 8.43% and 1.12% respectively, with syngas yield of 3.19 m^3/kg. The H_2/CO ratio of 1.93 implies that it is suitable for Fischer Tropsch technology. The obtained results however showed that the BC (bamboo culms) should be utilized as an alternative energy source in the energy diversification plan in Nigeria through gasification.

Keywords: Bamboo Culms, Biomass, Bubbling Fluidized Bed, Design, Gasification, Synthesis Gas.

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1. Introduction

Biomass gasification is sustainable, cleaner and benign energy production process on renewable basis (Ramin *et al*., 2019). Gasification is an important energy process compared to other recent technology that still utilizes fossil fuels, and it is gaining interest as an energy generating technique due to its shift away from petroleum sources (Gonzalez *et al.,* 2017). Biomass are lignocellulosic feedstock which comprises cheap and non-food materials from plants and they are surplus and have not been well utilized all over the world.

Crude oil today is still the major income in Nigerian economy and there is need for immediate diversification to other promising raw materials. Bamboo as a lignocellulosic biomass assures better economy as the most viable alternative with over 100 products derived from it. Bamboo is a fast developing plants on earth with largely uncoordinated production and untapped potentials (Atanda, 2015). Finished goods from bamboo are barely available since the technicality is lacking. Even as large growth of bamboo are scattered in some parts of Nigeria, there has not been any in-depth research on its properties, availability, distribution and usage in Nigeria. Though, fractions of bamboo that are used contribute a meager portion of the economy in Nigeria. In this regard, bamboo can be transformed into various raw materials for both domestic and industrial usage if properly harnessed. Moreover, it will encourage the export market with huge potential income of about US$22.0 billion on yearly foreign exchange for Nigeria (Ige, 2016). It has been reported 1.70 million hectares of local bamboo of two species are found in Nigeria ((Ige, 2016). This local bamboo grows well in southern region and villages in Delta, Niger, Benue, Cross river, Enugu, Anambra etc. The average diameter of bamboo culms (BC) varies from 3.2 to 9.1 cm while the average height varied from 2.7 to 4.4 cm (RMRDC, 2004a). It also retains its natural habitat during and after harvesting (Anon, 2015e). The planned method of growing bamboo reduces environmental pollution, with 35% CO_2 reduction in the climate and invariably oxygen will be evolved (Atanda, 2015). One of the major areas that should be identified and pursued vigorously by both the private sector and government is the use of bamboo for gasification to produce syngas. In Nigeria, the yearly production of bamboo culms deposits high volumes of lignocellulosic biomass as waste, and clean energy can be generated when this waste is converted. The conversion of bamboo culms into syngas and other energetic gases can be achieved via thermo-chemical routes such as gasification. The process of gasification involves the conversion of low value fuels such as biomass to volatile products in form of gas at elevated temperatures (600 – 1300 ^{o}C) using compressed air, pure oxygen and steam as fluidizing agents. This process is the best method and most efficient in extracting the energy in biomass. Currently, it is the most sought method for utilizing waste (Wu and Chein, 2015). Gasification reactions take place in a well modified cylinder called gasifiers, usually categorized based on their contacting mode or type and position of fluid used. Consequently, three broad designs of gasifiers exist namely; fixed bed, fluidized bed (bubbling and circulating) and entrained flow gasifiers (Demirbas, 2008; Basu, 2010).

However, the most appropriate gasifier for biomass gasification is the bubbling bed gasifier (Basu, 2006). This is as a result of its capacity to take wide range of solids; perfect mixing, efficient heat exchange and controlled temperature, moderate oxygen and steam requirement, high cold gas efficiency and continuous operation which yields uniform products in the gasifier (McKendry, 2002).

Fedrick and David, 2010 designed a 60MW circulating fluidized bed gasification system using rice husk as feedstock. The drop in pressure within the siphon is greater than the pressure drop along the return leg. However, this encourages effective solid circulation in the gasifier. Dayanada and Syreepathi, 2012, designed bubbling bed gasifier using chicken litter, saw dust and rice husk as raw-material. They experienced low product gas composition which was attributed to the low carbon combustion efficiency. The design of a fluidized bed gasifier was achieved using rice husk with the purpose of investigating the influence of the equivalence ratio on gas power and volumetric yield (Ramirez *et al.,* 2007). Their results showed that there was a decrease in a definite flow of gas due to the small quantity of feedstock used. A double phase biomass-air steam fluidized bed gasifier which showed that an increase in biomass-steam ratio and velocity of the fluid increases the rate gas production was developed (Sadaka *et al.,*2002). In addition, opening the valve for more steam flow-rate above 0.25 kg/min reduces the gas production rate due to the reduction in bed temperature.

In a nutshell, it can be seen that thermochemical conversion of lignocellulosic biomass to synthesis gas via gasification is a viable means for conversion of the carbon and oxygen in biomass into intermediate products with C_1 (CO and CH_4). The first generation syngas can be further transformed into high hydrocarbon products by microbial activities or thermochemical process (Petrus and Noordermeer, 2006). Due to many obvious advantages of gasification products, it is necessary to integrate units or systems that can transform the lignocellulosic material into sustainable energy and promote agricultural development in all nations.

The aim of this work is to utilize bamboo culms in a predefined reactive condition and fabricated fluidized gasifier by thermochemical process for efficient production of syngas that can serve as suitable feedstock in Fischer-Tropsch technology.

2. Materials and Method

2.1 Material analysis

The selected lignocellulosic biomass sample for this study is bamboo culms (BC) as shown in Figure 1 and was sourced along Delta steel express way Warri, Delta State, Nigeria. BC were chopped to a maximum of 30 mm in length, washed thoroughly with tap water to remove dirt's and sand. It was further sorted and arranged in a sack prior to preparation stage. The chopped and shredded samples were air dried over a period of 50 days.. The dried samples were milled with locally fabricated ball hammer machine and screened to different particle sizes (0.1 mm, 0.60 mm and 1.25 mm) using a mechanical shaker (Xiam 210) equipped with different mesh sizes of sieve plates as shown in Figure 2. The sieved material is as shown in Figure 3. Proximate analysis was carried out using experimental procedure described by Arpit and Kumar, 2017 todetermine the moisture content, fixed carbon content, ash content and volatile matter content of air-dried biomass samples ground to particle size below 1.0 mm. Determination of total carbon,nitrogen, oxygen and sulphur percentages in the biomass was carried out using energy dispersive x-ray spectroscopy (EDXS) at laboratoryof electron microscopy, North West University Potchefstroom campus, South Africa. The high heatingvalue (HHV) of the lignocellulosic biomass was determined with XRY-1A bomb calorimeter (Shanghai Changi China) using ASTM-D 4809. All the sieved material were stored at ambient temperature for use during operation. The bed material used for this study was sand sourced from Umudike Umuahia, Abia state, Nigeria. This sand was kilned at 1200 ^{o}C and screened to required size of 385 µm. The materials of construction were locally sourced such as

4 mm, 2 mm, and 1.5 mm thickness stainless steel pipes, 1.5 inch ball valve, 8 mm copper pipes, 18 mm cylindrical pipe dryer equipped with 32 g of silica gel pebbles, 8 mm and 10 mm pressure pipes, tedlar balls, angle bar etc.

Fig. 1. Freshly cut and sized samples of Bamboo culms

Fig. 2. Mechanical shaker (Xiam 210)

Fig. 3. Sieved BC material

3. Method

3.1. Design Method Consideration

The gasifier design was made using fluidization dynamic principle and standard procedure in literature (Ragnar, 2000; Perry's, 2002; Ramirez *et al*., 2007, Dayanadan and Syreepathi, 2012; Folayan, *et al.,* 2015) based on the physical properties of BC. Each of the sub-parts of the gasifier was designed as models.
The proposed BFBG systems for BC under study consist of reaction chamber, an air distribution plate and plenum aligned vertically as one unit.
For the design calculations, the following parameters; particle size dp (μm), particle density ρ (kg/m^3), sphericity ($\emptyset_s$) and porosity or voidage Ɛ(mf) of the bed materials and BC were determined. These values were used to deduce the hydrodynamic properties; minimum fluidization velocity U_{mf} (m/s), terminals velocity U_t (m/s) and fluidization velocity U_f (m/s) of the sand and biomass samples under study. The calculation procedure is as shown in Appendix A

3.1.1. Reaction Chamber

In bubbling fluidized bed gasification, a box with cross-section of circular cylinder is technically suited. Hence, a cylindrical stainless tube 4mm thickness with 0.063 m internal diameter was used for this study. Calculations were deduced for all components of the plant. It was made up of the reaction chamber attached with extended pipe 3 inches length by 1.5 inches internal diameter with torispherical head, air distribution plate and a plenum. Based on this, the gasifier height was determined.
The separation and lifting bed particles is known as fluidization (Natarajan *et al.,*1998; Sanchez, 1997; Kunni and Levenspiel, 1991; Lahijani and Zainal, 2011). Weight balance for a fluidized bed indicates that the ideal weight of the particle per unit area equal to the loss on fluid pressure across the bed particles. The relation can be expressed mathematically as shown in Equation 1 and 2 (Rhodes, 2008);

$$Pressure = \frac{Weight\ of\ particle - Upthrust\ on\ particle}{Bed\ cross\ sectional\ area} \qquad (1)$$

$$\Delta P = H(1 - \varepsilon)(\rho_p - \rho_f)g \qquad (2)$$

Where ρ_p *and* ρ_f represent bed particles density and fluid density respectively, H is the depth of bed, g is gravitational force and Ɛ is the voidage within the bed.
The pressure loss and fluid velocity share a relationship in a packed bed region as described by Equation 3 and 4 the Carman Kozeny equation for laminar and turbulent regime respectively, whereas equation 5 is the Ergun's equation for bubbling fluidized bed (Kunni and Levenspiel, 1991; Lim and Alimuddin, 2008; Rhodes, 2008; Basu, 2010).
For laminar regime (Carman-Kozeny equation):

$$\frac{(-\Delta P)}{H} = \frac{180\,\mu U}{x^2}\frac{(1-\varepsilon)^2}{\varepsilon} \tag{3}$$

For turbulent regime (Carman-Kozeny equation):

$$\frac{(-\Delta P)}{H} = \frac{1.75\,\rho_f U^2}{x}\frac{(1-\varepsilon)^2}{\varepsilon^3} \tag{4}$$

Erguns equation is given by the relation:

$$\frac{(-\Delta P)}{H} = \frac{150(1-\varepsilon)^2}{\varepsilon^3}\frac{\mu U}{x_{sv}^2} + \frac{1.75\,(1-\varepsilon)}{\varepsilon^3}\frac{\rho_f U_{mf}^2}{x_{sv}} \tag{5}$$

Substituting equation (2) into equation (5) and rearranging gives the following hydrodynamic parameters:

- The minimum fluidization velocity (U_{mf}):

  The minimum fluidization velocity(U_{mf}) is an important parameter to define the initial fluidization process (Sarker, 2012, Ramin, 2019). The particles bed will be deduced differently for the sand and the biomass materials using Equation 6.

$$U_{mf} = \frac{dp^2 * (\rho_p - \rho_f) * g}{150 * \mu_f} \times \frac{\varepsilon^3 * \emptyset^2}{1-\varepsilon} \tag{6}$$

- The terminal velocity of the particle (U_t):

  Equation 7 calculates the terminal velocity for the bed and particles within the range of Reynold's number $(0.4 < Re < 50)$ (Souza-Santos, 1996);

$$U_t = dp * \left[\frac{4\,(\rho_p - \rho_f)^2}{225 * \rho_f * \mu_f} * g^2\right]^{1/3} \tag{7}$$

- The fluidization velocity during the gasification (U_f) :

  The relationship connecting the expanded and minimum heights of the fluidized bedduring gasification process defined as superficial velocity of the gas was established (Chatterjee *et al.*, 1995); as shown in Equation 8.

$$\frac{H}{H_{mf}} = 1 + \left[\frac{10.978 * (U_f - U_{mf})^{0.738} * \rho_f^{0.378} * d_p^{1.006}}{U_{mf}^{0.937} * * \rho_f^{0.126}}\right] \tag{8}$$

The restriction suggested in Equation 9 for the bubbling fluidized bed is:

$$1.2 < \frac{H}{H_{mf}} < 1.4 \tag{9}$$

From Equation (9) 1.3 was chosen for the ratio $\frac{H}{H_{mf}}$ for the design, and substituting into Equation 8 gives Equation 10:

$$U_f = U_{mf} + \left[\frac{0.3 * U_{mf}^{0.937} * \rho_f^{0.126}}{10.978 * \rho_p^{0.376} * d_p^{1.006}}\right]^{1/0.738} \tag{10}$$

- The overall height of the reaction chamber:

  The establishment of this parameter can be expressed as shown in Equation 11 (Kunni and Levenspiel, 1991):

$$H_t = TDH + H \tag{11}$$

Based on the calculations, an internal diameter of 0.063 m, gasifier height 0.13m and $H_{mf} = 0.09m$ were selected for the design of the reactor. Within a particular height, only small quantity of particles disengaged and falls back to the bed. This upward force generated by induced air balances the light particles up the bed. In other words, light particles are entrained above the height of the freeboard when the upward velocity of air is sufficiently high. At this length of the free-board is called the transport disengagement height (TDH). Based on the aim of reducing the slugging phenomena, deducing the threshold disengagement height (TDH) taking into account the superficial velocity,U_f, and the gravitational force, g, was made in agreement with empirical correlation as shown in Equation 12 (Folayan *et al.,*2015):

$$TDH = 1.08{U_f}^{1.2}6.71 - 1.21gU_f \tag{12}$$

- Thickness of Cylinder:

A cylindricalpressure vessel is classified as thick or thin walled based on its thickness to radius ratio. Thick walled cylinder was used because of the pressure within the enclosed vessel as explained by Equation 13 (Folayan *et al*., 2015):

$$t/_r \geq 1/_{10} \tag{13}$$

Where, t is the thickness of the cylinder and r is the radius.

- Insulation Thickness

In this study, a rock wool 12 mm thickness was used to insulate the cylinder. The primary aim of insulating the reactor is to conserve energy and minimize financial and thermal loses. In a cylindrical piece, increasing the insulation thickness increases the conduction resistance of the insulation layer but decreases the convection resistance of the surface.

3.1.2. Biomass Feeding Subsystem(Gravity Chute Hopper)

A gravity chute is a simple device where biomass feedstock is fed into the bed via a chute which functions with the help of gravity. Thissub-system was made of a stainless steel hopper for the biomass material storage screwed to a 1.5 inches ball valve attached to a 1.5 inches stainless steel pipe inclined at an angle of 40° into the reactor bed at a height of 0.0875 m from the reactor bed for injection of raw biomass into the gasifier. The feeder (the hopper base) will supply biomass by means of gravity to the bed at a specified rate with the aid of ball valve to avoid fuel accumulation that will cause system blockage.

3.1.3.Two Opposite Torispherical heads

The disparity between the bubbling and circulating fluidized bed is the return leg attached to the circulating bed that returns converted and unconverted bed materials back into the circulating bed. In this study two opposite extended pipe 3 inches length by 1.5 inches internal diameter pipe in form of a returned leg was attached to a bubbling bed gasifier at an angle of inclination of 60° to the fluidized bed at a height of 0.0875 m from the reactor bed. It was designed to create turbulence within the reactor bed and ensure perfect mixing of gas and solids.

3.1.4.Plenum Chamber

The design of plenum chamber helps to achieve initial uniform distribution of fluidized gas before passing the distributor plate. Various designs of plenum have been proposed at any location of gas entry into the fluidization column. But, the vertical gas entry was chosen for the design for faster uniform distribution as shown in Equation 14 and 15 (Litz, 1972).

$$D_{entry} > D_{plenum}/36 : \quad H_{plenum} = 3(D_{plenum} - D_{entry}) \tag{14}$$

$$D_{entry} < D_{plenum}/36 : \quad H_{plenum} = 100D_{entry} \tag{15}$$

3.1.5. Air Distribution Plate

The porous plate or plate distribution grate or straight hole orifice type plates (plate with vertically drilled holes) was chosen for bubbling fluidized bed design. This is due to the simplicity of the design, minimal operating constraints and the cost of fabrication. Combustion air was injected into the furnace via anairdistributor plate, which holds the bed materials and evenly spreads the fluidizing gas into the solid bed. Table 1 shows the input data representing the design of air distribution plate considered for the bed material.

Table 1
Design Specification for AirDistributor Plate

Parameter	Value
Fluidization Velocity ($\boldsymbol{m/s}$)	2.18
Minimum Fluidization velocity ($\boldsymbol{m/s}$)	0.39
Minimum fluidiation height ($\boldsymbol{m}$)	0.127
Particle density ($\boldsymbol{kg/m^3}$)	260
Bedporosity	0.48
Beddiameter ($\boldsymbol{m}$)	0.063
Number of perforation	96

3.1.6. Air Pre-heater

The exit air from the oiless compressor was connected to a pre-heater (a 30 turns coiled pipe embedded inside a 1 foot 2.5 inches internal diameter pipe) attached with a heater band rated 1000 watts to preheat the exit air from the compressor. A K-type thermocouple was mounted on the pre-heater to control the air temperature within the experimental range. During the preheating of the reactionchamber, a propane gas burner connected to the entrance of the plenum was selected. The combustion gas generated by the burner crosses the sand of the bed warming it up around 600℃. At this temperature the fluidized bed temperature will ensure the biomass self-ignition, giving start to the autonomy of the combustion and gasification reactions.

3.1.7. Cyclone Subsystem

Cyclone subsystem is an essential part of a fluidized bed gasifier, it collects particulate material during the gasification process. Cyclone has no moving parts, its easy fabrication and high efficiency makes it suitable for fluidized bed gasifier. Cyclone uses vortex means without filters to separate particulates from air or gas stream. The most common cyclone design consists of a top cylindrical part, bottom conical part and a tangential inlet. Usually a gas mixed with particulates passes via the tangential point close to top of the cyclone. The gas flow is forced into a downward spiral simply because of the cyclone shape and the tangential entry (Cooper and Alley, 1986). From the literature(Ashbee and Davis, 1992), a standard cyclone with dimensions and geometric relations shown by Stairmand was chosen. Table 2 shows the calculated dimension from the empirical co-relation.

Table 2
Cyclone dimensional values.

Sub-parts parameter	Value
BodyDiameter (m)	0.08
Height of inlet(m)	0.04
Width of inlet(m)	0.04
Diameter of Gas Exit (m)	0.04
Length of Vortex Finder (m)	0.04
Length of Body (m)	0.12
Length of Cone (m)	0.2
Diameter of Dust Outlet (m)	0.03
Stainless steel thickness (mm)	2

4. Fabrication of Bubbling Fluidized Bed Gasifier

The pilot bubbling fluidized bed gasifier plant was fabricated at Dykes fabrication company along Umuahia Ikot-Ekpene express way and was set-up for experiment at Engineering workshop NationalRoot and CropResearch Institute (NRCRI) Umudike, Abia state, Nigeria. The main elements of this experiment includes a well-insulated bubblingfluidized bed gasifier system with internal diameter of 0.0635 m, height of the reactor bed is 0.127 m and the total height of the gasifier is 0.678 m. Two opposite 3 inches pipes with torispherical heads were attached and inclined at an angle of 60° at a height of 0.0875 m from the reactor bed. A feeding system that comprises a gravity chute feed hopper with 1.5 inches ball valve attached at the reactor bed at an angle of 40° to the horizontal and height of 0.0889 m was used for proper delivery of the feedstock. A stainless steel steam generator having two injection points on the gasifier at 0.1524 m and 0.3302 m from the reactor bed was equipped with an internal and external heater. An air supply unit (24 litres oiless compressor with working pressure of 8 bar) which injects the required inlet air was attached with a heater that increases the temperature of the outlet air. A propane gas was used as a source of combustion. Two flow meters (rotametre) for air and gas were used to regulate the flow of the fluid. Cyclone, ash bottle and dry filter packed with 32 g silica pebbles, a condenser unit, a gas sampling and collecting unit with 8 mm pressure pipe attached with a pin deflator and internal valve tedlar balls were all incorporated for gas cleaning and sampling. 5 type K-thermocouples and 2 pressure gauges were used to maintain temperature and pressure within the system. The schematic diagrams of the set up in 2D and 3D views are shown in Figures 4 and 5 respectively. The top and side views are shown in Appendix B. The live set-up of the pilot plant before and after lagging are shown in Figures 6 and 7 respectively. The product ejected out of the gasifierwas made dust free and cleaned by passing it through a gas cleaning and cooling system before collecting it with a tedlar internal valve ball, before injecting into a gasanalyzer. The hot product gas from the fluidized bed gasifier will be made to pass through a cyclone to remove the larger particles. After passing through the cyclone, the gas will still contain dust particles and tar and hence, it will be further cleaned and dried by passing it through a dry filter column packed with silica gel. It is then cooled by passing it through a condensing unit. The dry, cleaned and cool synthesis gas was analyzed offline with gas chromatographic machine where it measures the composition of synthesis gas CO and H_2 presence in the fuel gas.

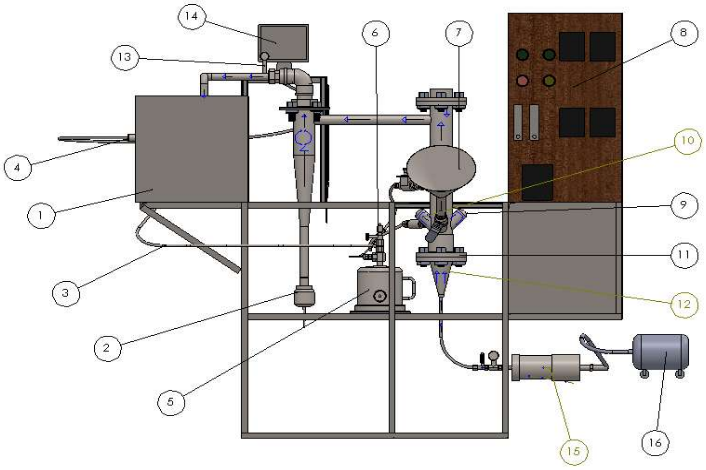

Figure 4: Schematic 2D diagram of a Pilot plant Bubbling fluidized bed gasifier.

(1) Condenser unit (2) Ash bottle (3) Hot water exit from condenser (4) Syngas exit (5) Steam generator (6) Steam pressure indicator (7) Feed Hooper (8) Integrated control panel (9) 3" protruded torispherical head (10) Steam injection point 1 (11) 2 $^1/_2$ " stainless steel flange (12) Plenum (13) Exit temperature thermometer (14) Gas Analyser (15) Air preheater (16) Compressor

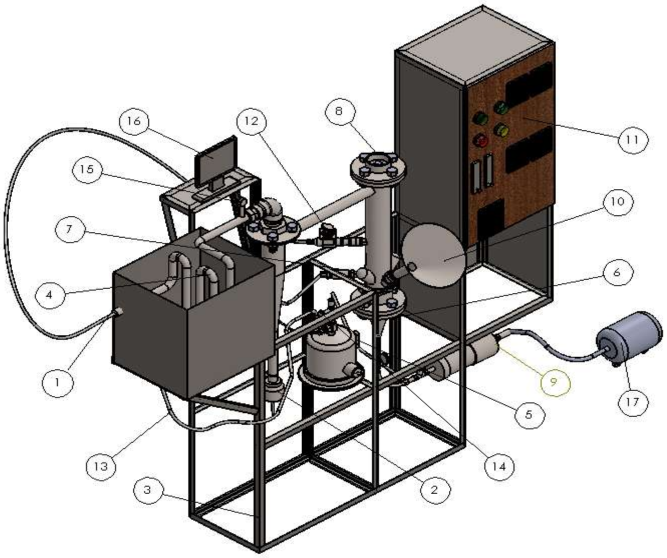

Fig.5. Schematic 3D diagram of a Pilot plant Bubbling fluidized bed gasifier.

(1) Syngas exit (2) Ash bottle (3) Metallic frame (4) Condenser unit (5) Steam generator (6) Plenum (7) 3" protruded torispherical head (8) $2\,{}^{1}/_{2}$ " stainless steel flange (9) Air preheater (10) Feed Hooper Steam injection point 1 (11) Integrated control panel (12) Top steam injection valve (13) Hot water exit from condenser (14) Steam pressure indicator (15) Exit temperature thermometer (16) Gas Analyser (17) Compressor

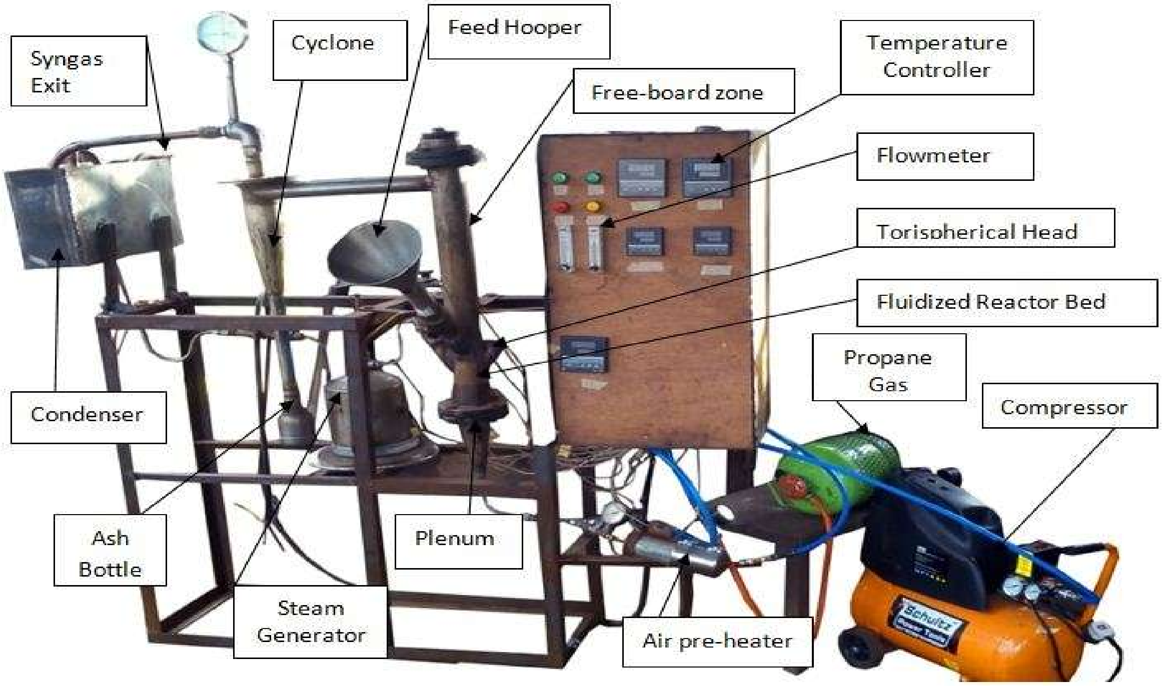

Fig. 6. Live set-up of the Pilot plant Bubbling fluidized bed gasifier before lagging

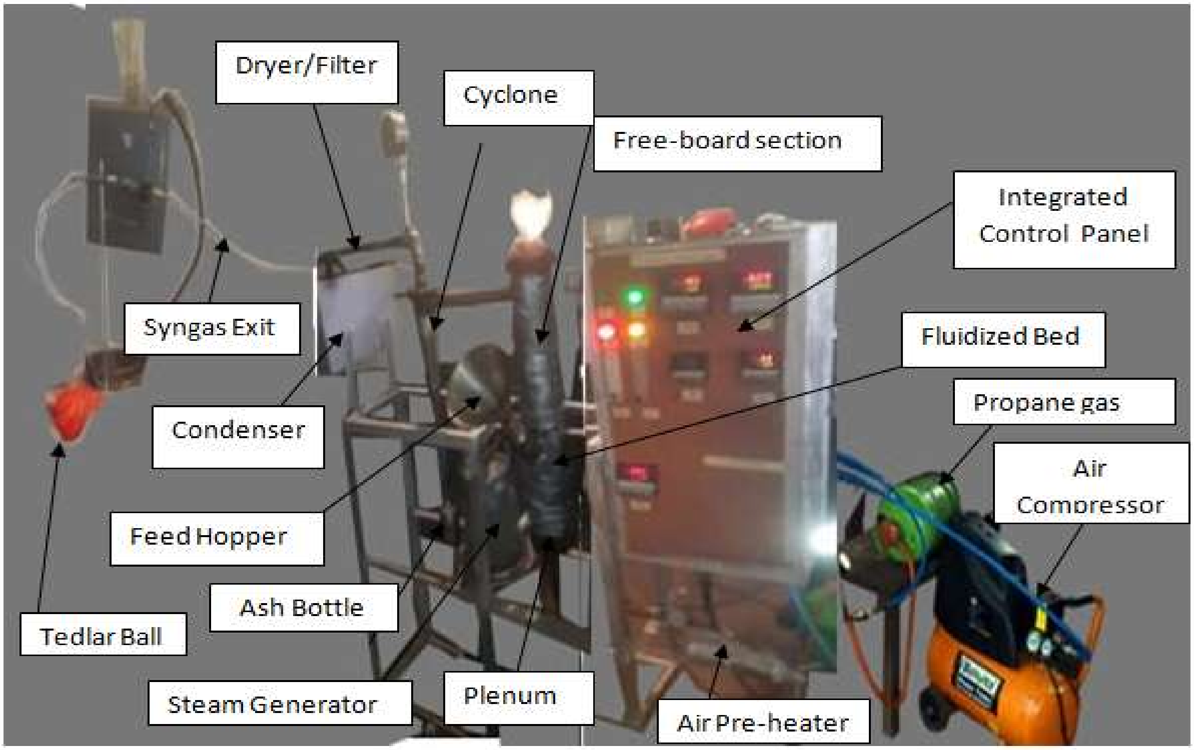

Fig. 7. Live set-up of the Pilot plant Bubbling fluidized bed gasifier after lagging

5. Experimental Procedure

5.1. Start-Up

A Pilot plant bubbling fluidized bed gasifier was designed, fabricated and set up to carry out the experimental investigation. In this study, the preheating of the gasifier reactor for start-up takes 50 minutes by blowing air over sand with 2.5 inches (1000 watts) heater band wrapped round the reactor. Sand (1.0 µm, 300 g) an inert bed material, was placed into the gasifier via a gravity chute feed hopper. Feeding of sand and biomass material into the reactor bed was done manually. The sand is supported by a distributing plate with aperture lesser than the sand particles. From the control panel, the heater band circuit breaker wasturned on to preheat the fluidized bed reactor, meanwhile the compressor was turned on and air was introduced tothe gasifier bed via the distributing plate. The air flow rate being adjusted with flow meter to keep the sand fluidized. Also, the air pre-heater was switched on, which is varied from 25-65˚C with aid of a k-type thermocouple. The fluidizing sand was monitored 10 minutes to ensure that the fluidized bed has become stable and the materials were completely dry. The propane gas valve was opened gradually and combustion at the reactor bed took place by igniting from the top with subsequent increase in temperature of the bed to 650 ℃. Increase in temperature was as a result of the heat released from the combustion of propane and air mixture. The propane gas supply was locked as soon as combustion can be maintained by the biomass material alone (auto-ignition). The steam generator and the super-heater were turnedon for the preparation of the steam for the experiment. Feeding of the biomass starts at reactor bed temperature of 650℃. The produced steam temperature reached 130℃ and was superheated to a temperature of 250℃ before the stainless steel valve was opened to inject superheatedsteam into the bubbling fluidized bed gasifier. The plant wasoperated at this condition for 30 minutes to ensure that a steady-state condition was reached in the fluidized bed reactor. Experimental runs were conducted under steady state by allowing enough time for the system to stabilize each time certain parameters were changed. The average of first gas sample were collected after 3 minutes of feeding, and the gas samples was collected and trapped inside a tedlar ball equipped with internal valve prior to offline gas analysis. The biomass feed rate of 0.14 g/sec was used for every experiment.

5.2. Steady State Operation

The plant was allowed to operate for 30 minutes to ensure that all the units (electrical, instruments and equipments) are in stable conditions. At these conditions, there is no observable change displayed on the control panel or attached instruments at every set point made for control. Experimental runs were conducted under steady state by allowing enough time for the system to stabilize each time certain parameters were changed. At constant temperature and produced syngas flow, steady state was maintained. The temperature controllers, pressure gauge, relief valve and the flow meters were all operating in a stable condition.

5.3. Shutdown Procedure

At the completion of sampling and data recording, the heat supply (heater band), the air supply unit, steam supply unit, and main circuit breaker (MCB) were shut down in that order. Maintenance was followed as the need arises.

5.4. Analysis of Synthesis Gas

Duringthe steady state operation, the collection and analysis of the gas samples started. The design of the gassampling unit ensured that the gas sampling line was free of tar and particles. The gas samples were collected from the gas sampling line using an $8mm$ pressure pipe attached with a deflator and tedlar balls equipped with internal valve. Samples were collected at different experimental runs with corresponding process conditions as designed by the experiment. The gases were analyzed off-line to determine the synthesis gas composition using a Gas Chromatographic (GC) machine. The gas analyzer (BUCK Scientific Model 910, Analtrace Environmental Consultant and Laboratories, Warri, Delta State) was equipped with a thermal conductivity detector (TCD) and flame ionization detector (FID), and was used to perform the analysis. A GC column (RESTEK MXT-1, USA) of 30 m length, 0.53 mm ID and 10 μm df, connected with an injection loop of 500 μl were selected for the chromatographic analysis. The GC was calibrated using helium as a carrier gas at 20 psi and standard gases (mixture containing CO, CO_2, H_2 and CH_4 in nitrogen). The temperature of the oven was set to 50 ℃, the heater temperature (detector) was set at 200 ℃ and the injector temperature at 100 ℃.

However, the gas compositions were determined by the help of retention time and its concentration which was calculated based on the area under the related peak.

6.1.Equivalence ratio

The equivalence ratio (ER) is among the key variable in a gasification set-up that maintains the efficiency of the system. Practically, it defines the bed and freeboard temperatures, the qualities of the syngas and efficiency of the system.

$$ER = \frac{(R_{A/F})r}{(R_{A/F})s} \qquad (22)$$

Where, $(R_{A/F})r$ is the real or actual air to fuel ratio, $(R_{A/F})s$ is the stoichiometric air to fuel ratio. The real air to fuel ratio is calculated from the expression in Equation 23.

$$(R_{A/F})r = \frac{m_{air}}{3.547 * m_{BC}} \qquad (23)$$

Where, m_{BC} is the mass of the bamboo culms.
The air mass flow rate m_{air} is calculated as:

$$m_{air} = Volumetric\ flowrate * Density\ of\ air \qquad (24)$$

$$(R_{A/F})s = 8.89(\%C + 0.375 * \%S) + 26.5 * \%H - 3.3 * \%O \qquad (25)$$

6. Material and Energy Balance

The calculation of balance in mass and energy were carried out using the results of elemental and proximate analysis. Invariably, concentrations of the synthesis gas generated during the experiment were hydrogen (H_2), carbon-monoxide (CO), carbon(IV)oxide (CO_2), methane (CH_4) and nitrogen (N_2) (Shabudeen, 2010). The CO_2, H_2O and N_2 compositions in the syngas is a factor of the fuel chemical composition, the amount of air in the reaction and the condition of the dryer packed with 32 grams of silica gel pebbles. Thus, adequate global reaction for the gasification operation was presented as follows:

$$CH_{0.32}O_{0.70}N_{0.02}S_{0.0003} + hH_2O + a(O_2 + 3.76N_2) \rightarrow x_1CO + x_2CO_2 + x_3H_2 + x_4CH_4 + x_5H_2O + x_6N_2 \qquad (26)$$

In this regard, the molar balance for each element in equation (26) gives the global gasification reaction coefficients in Equation (27).

$$CH_{0.32}O_{0.70}N_{0.02}S_{0.0003} + H_2O + (O_2 + 3.76N_2) \rightarrow CO + CO_2 + 1.54H_2 + CH_4 + 0.7H_2O + 3.8N_2 \qquad (27)$$

The mass and energy balance summary table as shown in Table 3.

Table 3
Mass and Energy Balance Summary Table

Material Balance	
Material Parameter	Value (kg)
Total N_2 supplied by air and BC	0.0598
Total oxygen flow to gasifier with dry air, steam, BC moisture content and wet air	0.627
Total H_2 in flow to the gasifier with BC, steam and moisture	0.3069
Total H_2 associated with H_2 and CH_4 in Syngas	0.1519
H_2 associated with H_2S as waste	0.000025
Moisture in syngas in form steam	0.01132
Steam associated with hydrogen in syngas	0.1019
Oxygen associated with CO and CO_2 in dry syngas	0.922
Oxygen associated with steam in gas	0.820
Total carbon associated with CO, CO_2, and CH_4	0.3564
Energy Balance	
Energy Parameter	Value (MJ/kg)
Energy output with CO	5.168
Energy output with H_2	10.05
Energy output with CH_4	1.247
Energy Input HHV BC	17.132

7. Results and discussion

7.1 Biomass characteristics

The proximate, ultimate analysis and calorific value of BC are shown in Table 4. The low percentage of nitrogen and sulfur indicates BC is an environmentally friendly fuel. The calculated molecular formula for BC is $CH_{0.32}O_{0.70}N_{0.02}S_{0.0003}$. The sample contains a high percentage of volatile matter (78.43%), low moisture content (3.60%), ash content (3.20%) and fixed carbon below (14.77%). The volatile matter to fixed carbon in biomass and coal was greater in biomass by 4.0 than coal that is 1.0(Kumar *et al* 2009). This suggests that gas phase combustion will be a predominant feature of the thermo chemical conversion of fuel (Kumar *et al* 2009; Wang *et al.,* 2009).

Table 4: Analysis of Bamboo Culms

Proximate Analysis (%)	Value
Moisture content (MC)	3.60
Volatile Matter (VM)	78.43
Ash content	3.20
Fixed carbon (FC)	14.77
Ultimate Analysis (%)	
Carbon	50.64
Hydrogen	1.34
Oxygen	47.06
Nitrogen	0.92
Sulphur	0.04
Higher Heating Value (HHV)	
Calorific value (MJ/kg)	21.221

Table 5 shows the physical properties of BC used for design of the bubbling fluidized bed gasifier. The density of the gasifying medium is 1.171 kg/m^3 at 25 ˚C. From the particle size and the density of the crushed particles of BC and sand, the type of fluidization expected can be determined (Kummi and Levenspiel, 1991; Nyakuma *et al*., 2012). The result of both BC and sand can be classified as Group B particles according to Geldert classification criteria (Geldart, 1973). Solids classified in Group B particle ranges in particle sizes in the range of 40 µm < dp < 500 µm and the fluidization occurs with vigorous bubbling action. Also, bubbles appear when gas velocity exceeds the minimum fluidization velocity.

Table 5: Physical properties of Bamboo Culms

Property	Sand	BC
Particle size dp (µm)	385	373
Bulk density ρ (kg/m^3)	2600	260
Sphericity $\emptyset_s$	0.66	0.64
Porosity ε_{mf}	0.42	0.48
Density of gasifying medium ρ_a (kg/m^3) = 1.171		

The sand and BC particles are categorized into group B according to Geldert's classification for coarse particles. From Table 6 it can be observed that the terminal velocity U_t is lesser than the superficial velocity U_f for fine particles. The fluidizing velocity for materials in a fluidizing bed is usually 0.5 to 1.7 m/secs (Geldart, 1973; Nyakuma *et al*., 2012).

Table 6

Hydrodynamic Properties for Sand and BC

Property	Sand	BC
Minimum fluidization velocity U_{mf} (m/s)	0.04	0.39
Terminals velocity U_t (m/s)	0.11	0.07
Fluidization velocity U_f (m/s)	0.15	2.18
Overall height of the reaction chamber (m)	0.678	

7.2 Combined effect of Equivalence Ratio and extended torispherical head.

The performance of the bubbling fluidized bed gasifiers depends mainly on the equivalence ratio (ER) being used. Moreover, the extended torispherical head induced more reaction as a result of the rebound collision within the reactor bed. The lower limit of the range is determined by the minimal amount of air required to oxidize the fuel and generate enough heat to maintain the gasification endothermic reactions. Very small values of ER reduced the reaction temperature and energy liberation necessary to maintain the reduction reactions. On the other hand, high equivalence ratio caused increase in the reaction temperature because greater amount of oxygen favors the combustion phase. Moreover, higher ER means a higher gasification temperature which can accelerate the gasification process and improve the syngas quality (Sharmina *et al.,* 2014). The oxidation reaction for CO and CO_2 production is $C + {}^{1}/_{2}\, O_2 \rightarrow CO$ and $C + O_2 \rightarrow CO_2$ respectively. Based on the oxidation reactions CO evolution consumes more C for the same amount of O_2. Therefore, an optimum ER of 0.36 delivers a quality syngas that suits Fischer-Tropsch technology. The results obtained from the developed gasifier were tested for BC. The attached two opposite 3 inches stainless steel by length inclined at angle of 60 degrees to the reactor with torispherical head could be the reason for high carbon conversion rate when compared to those in the literature without the extended pipe. The comparison of different gases obtained from the developed bubbling fluidized bed gasifier is made with results of the other authors to validate the proposed design as presented in Table 7.

Table 7
Comparison of BC experimental results with results of other biomass from different authors

Parameters	Current study (BC)	Guangul et al. (2012) Oil Palm Frond	Dayananda and Sreepathi (2012) Rice Husk	Arena et al.(2007) Saw Dust	Barriga (2002) Rice Husk
Bed Temperature (℃)	873	985	NA	NA	NA
Air Temperature (℃)	43.26	36	NA	NA	NA
Steam injection height (m)	0.33	NA	NA	NA	NA
Steam Biomass ratio	0.31	NA	NA	NA	NA
Equivalence ratio	0.36	0.27	0.21	0.34	0.40
H_2 (%)	28.14	8.47	2.00	2.8	4.39
CO (%)	14.61	22.78	9.4	9.6	14.26
CO_2 (%)	8.43	11.81	18.2	18.2	NA
CH_4 (%)	1.12	2.02	0.45	9.7	3.29
$H_2/CO\ \ Ratio$	1.93	0.37	0.21	0.29	0.31
Syngas yield(Nm^3/kg)	3.19	1.91	NA	NA	1.88
Carbon Conversion Efficiency (%)	81.62	74.4	NA	NA	NA
High Heating Value (MJ/Nm^3)	5.88732	4.66	NA	NA	NA

8. Conclusion

The design, operation and applicability of a bubbling fluidized bed gasifier in Nigerian using a local bamboo culm have been carried out. The syngas H_2/CO ratio obtained shows that the product is suitable as a raw material to Fischer-Tropsch synthesis technology. The two 3 inches pipe with torisperical head attached opposite to each other at an angle of 60 degrees to the reactor bed contributed to the high raw material conversion. The successful operation of the gasifier and the result obtained strengthens the research towards the development of clean source of energy and usage of agricultural waste in Nigeria with the use of fluidized bed gasification technology.

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APPENDIX A
Sample Calculation For The Hydrodynamic Properties

- Minimum fluidization velocity $\left(U_{mf}\right)$:

$$U_{mf(sand)} = \frac{(385*10^{-6})^2*(2600-1.171)*9.81}{150*4.014*10^{-5}} \times \frac{0.42^3*0.66^2}{1-0.42} = 0.04m/s$$

$$U_{mf(BC)} = \frac{(373*10^{-6})^2*(260-1.171)*9.81}{150*4.014*10^{-5}} \times \frac{0.48^3*0.64^2}{1-0.48} = 0.39m/s$$

- Terminal velocity of the particle (U_t):

$$U_{t(sand)} = 385 * 10^{-6} * \left[\frac{4\,(2600-1.171)^2}{225 * 1.171 * 4.014 * 10^{-5}} * 9.81^2\right]^{1/3} = 0.11\,m/s$$

$$U_{t(BC)} = 373 * 10^{-6} * \left[\frac{4\,(260-1.171)^2}{225 * 1.171 * 4.014 * 10^{-5}} * 9.81^2\right]^{1/3} = 0.07\,m/s$$

- Fluidization velocity during the gasification $\left(U_f\right)$:
- $U_{f(sand)} = 0.04 + \left[\frac{0.3*0.04^{0.937}*1.171^{0.1\ 26}}{10.978*2600^{0.376}*385*10^{-6^{1.006}}}\right]^{1/0.738} = 0.15\,m/s$

- $U_{f(BC)} = 0.24 + \left[\frac{0.3*0.24^{0.937}*1.171^{0.1\ 26}}{10.978*260^{0.376}*373*10^{-6^{1.006}}}\right]^{1/0.738} = 2.18\,m/s$

- Overall height of the reaction chamber:

$$H_t = 0.55 + (0.064 \times 2) = 0.678m$$

Where TDH = 0.55

APPENDIX B **Material and Energy Balance Calculation**

Data generated from empirical relation:
Basis: 1kg of BC
Chemical formular for BC
Steam supply is 0.15kg/kg of BC
Dry air supplied 2.1078kg/kg of BC
Specifichumidity of air 0.01kg of moisture per kg of dry air.
For Nitrogen:
Air contains 79% N2 by mass
So, N2 from air = $0.79 * 2.1078 = 1.665\ kg\ N_2/kg\ BC$
Total N_2 supplied by air and fuel which carry 0.92% N_2 :

$$= 1.665 + 0.0092 = 1.674\,kg\ N_2/kg\ BC$$

$$= \frac{1.674}{28} = 0.0598\,kgmol/kg\ BC$$

Since cleaned gas contains 47.7% N_2
Hence,
Amount of syngas produced:

$$= \frac{0.0598}{0.477} = 0.125\ kgmol\ N_2gas/kg\ BC$$

For Oxygen:
Oxygen supplied to gasifier is (with air) ie dry air:

$$= 0.21 * 2.1078 = 0.443\,kg\ O_2/kg\ BC$$

Oxygen associated with steam supply:

$$= \frac{16}{18} * 0.15 = 0.133\, kg\ O_2/kg\ BC$$

Oxygen with moisture in air supply (wet air):

$$= \frac{16}{18} * \frac{0.01\ kg\ of\ moisture}{kg\ of\ dry\ air} * 2.1078 = 0.0187\ \ kg/kg\ BC$$

Oxygen with moisture content in fuel (BC):

$$= \frac{16}{18} * 0.036 = 0.032\ \ kg/kg\ BC$$

Total oxygen flow to gasifier with dry air, steam, BC(MC), water:

$$= 0.443 + 0.133 + 0.032 + 0.0187 = 0.627\ \ kg\ O_2/kg\ BC$$

For hydrogen:

Total H_2 in flow to the gasifier with BC, steam, moisture content of BC and moisture in air:

$$H_2\ with\ steam\ supply = \frac{2}{18} * 0.15 = 0.017$$

$$H_2\ with\ moisture\ in\ BC = \frac{2}{18} * 0.036 = 0.004$$

$$H_2\ with\ moisture\ in\ BC = \frac{2}{18} * 0.01 * 2.1078 = 0.0023$$

$$H_2\ dry\ with\ air\ inflow = (\frac{2}{18} * 2.1078) + \frac{1.34}{100} = 0.028$$

H_2 associated with H_2 and CH_4 in syngas:

$$= (0.2814 * 2) + \ (0.0112 * 4) * 0.125 = 0.07595\ \ kgmol/kgBC$$
$$= 0.07595 * 2 = 0.1519 kg/kg\ BC$$

For Sulphur (Hydrogen associated with H_2S as waste):

Assuming all S converted to H_2S and removed by the gas cleaning system. Hence, hydrogen associated with H_2S in the raw syngas is found to be;

$$= 0.0004 * \frac{2}{32} = 0.000025\ \ kg/kg\ BC$$

Total hydrogen in the uncleaned dry syngas including that in H_2S:

$$= 0.07595598 + 0.0000025 = 0.07598\, kg/kg\ BC$$

To find the moisture in the syngas, the hydrogen in dry gas is deducted from the total hydrogen in flow obtained earlier using the hydrogen balance;

$$= Hydrogen\ in\ flow - Hydrogen\ in\ dry\ gas$$
$$= 0.0873 - 0.07598 = 0.01132\ \ kg/kg\ BC$$

Steam associated with this hydrogen in the gas:

$$= 0.01132 * \frac{18}{2} = 0.1019\, kg/kg\ BC$$

For oxygen:

Oxygen associated with CO and CO_2 in dry gas which have each I mole of oxygen respectively.

$$\{(0.1461 * 1) + \ (0.0843 * 1)\} * 0.125 = 0.0288\ \ kgmol/kg\ BC$$
$$= 0.0288 * 32 = 0.922\, kg/kg\ BC$$

Oxygen associated with the steam in the gas;

$$= 0.922 * \frac{16}{18} = 0.820\, kg/kg\ BC$$

Total oxygen in gas:

$$= 0.922 + 0.820 = \ 1.742\ kg/kg\ BC$$

Note: Here we note that this slightly more than the oxygen in flow of 0.820 $kg/kg\ BC$ calculated earlier. This must be due to measurement errors in the given data on fuel syngas composition.

For carbon:

Total carbon associated with $CO, CO_2$and CH_4 in dry gas whose production rate has been computed earlier as 0.125 $kgmol/kg\ BC$ is:

$$= (0.1461 + 0.08 + 0.0112) * 0.125 = 0.0297\ \ kgmol/kg\ BC$$
$$= 0.0297 * 12 = 0.3564\ \ kg\ of\ C/kg\ BC$$

Carbon conversion efficiency:

$$= \frac{0.3564}{0.5064} * 100 = 70.38\%$$

Energy Balance:

Carbon dioxide does not combust, which means it does not react with oxygen to produce water and carbon dioxide. Therefore, it has no enthalpy of combustion.

Components	Heat of Combustion($MJ/kgmol$)
CO	282.99
H_2	285.84
CH_4	890.36

Energy output with CO:

$$= 0.1461 * 0.125 \ (kgmol/kg\ BC) * 282.99 = 5.168 \ MJ/kg\ BC$$

Energy output with H_2:

$$= 0.2814 * 0.125\ (kgmol/kg\ BC) * 285.84 = 10.05\ MJ/kg\ BC$$

Energy output with CH_4:

$$= 0.0112 * 0.125\ (kgmol/kg\ BC) * 890.36 = 1.247\ MJ/kg\ BC$$

Total energy output:

$$= 5.168 + 10.05 + 1.247 = 16.465 \ MJ/kg\ BC$$

Total energy input(Channiwals/parrikh):

$$HHV_B = 0.3419C + 1.1783H + 0.1005S - 0.1034O - 0.015N - 0.0211A\ (MJ/kg)$$

$$= (0.3419 * 50.64) + (1.1783 * 1.34) + (0.1005 * 0.04) - (0.1034 * 20.31) - (0.015 * 0.92) - (0.0211 * 3.20) = 17.132\ MJ/kg\ BC$$

Cold gas efficiency of gasifier :

$$= \left(\frac{16.465}{17.132}\right) * 100 = 96.11\%$$

Equivalence Ratio:

The equivalence ratio of the gasification process is one of the most important parameter for the adjustment of the operating conditions.

$$ER = \frac{(R_{A/F})r}{(R_{A/F})s}$$

Where; the air fuel real is calculated from the expression:

$$(R_{A/F})r = \frac{m_{air}}{3.547 * m_{BC}}$$

Where 3.547 is calculated according to stoichiometery ratio of air to biomass for unit of biomass involved in complete combustion with presence of excess air.

The air mass flow rate m_{air} is calculated as:

$$m_{air} = Volumetric\ flowrate * Density\ of\ air$$

Density of air = 1.171 kg/m^3

Actual volumetric flowrate = 3.60 m^3/h

$$m_{air} = 3.60\, m^3/h * 1.171\, kg/m^3 = 4.22 \ kg/hr$$

Therefore;

$$(R_{A/F})r = \frac{4.22 \ kg/hr}{3.547 * 1} = 1.1897$$

$$(R_{A/F})s = 8.89(\%C + 0.375 * \%S) + 26.5 * \%H - 3.3 * \%O$$

$$(R_{A/F})s = 8.89(50.64 + 0.375 * 0.04) + 26.5 * 1.34 - 3.3 * 47.06 = 330.53$$

$$= \frac{330.53}{100} = 3.305$$

Therefore;

$$ER = \frac{1.1897}{3.305} = 0.359$$