Development of an Engineering Scale Coal wastes Gasifier for Heat Applications in Tanzania

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Abstract
The present work intended to design, fabricate and test an engineering scale Fluidized Bed Gasifier as an experimental and Test Rig for investigating the effects of reactor temperature and steam to coal wastes and biomass. Fluidized bed gasification (FBG) is an emerging energy conversion technology for solid and liquid fuels, well suited to low-grade fuels and waste materials. However, due to heterogeneous solid composition of coal wastes and biomass, unique gasification process model in a predetermined gasifier is required. The development of the gasifier was done in which 1kg of feedstock (at 15% moisture content) produces 2.0-2.5 Nm$^3$ of product gas, with approximately 1.027kg/Nm$^3$ cool and clean gas density. The produced gas has an energy content of approximately 5-6 MJ/m$^3$ that is more than the solid coal waste/biomass energy content. The feedstock consumption of the gasifier is about 1kg for 1.2 kWe electrical output. Analysis of the effect of height above the grate to product gas, flame temperature and flame stability of the gas has been performed. Results illustrate the potential of the developed FBG to produce good amount of syngas at a range of 18-20vol % for carbon monoxide and 13-16vol % for hydrogen as major components of syngas produced with a stable gas flame, hence demonstrates in an improve of fuel flexibility. The results also promote further utilization of coal wastes and biomass that could provide a cheap, reliable, and low carbon means for producing syngas for fuel flexibility in Tanzania.

Keywords: Fluidized bed Gasifier, gasification and syngas
1.0 Introduction

Worldwide, 80% of all energy used by human comes from fossil fuels (Chopra & Jain, 2007). Therefore, researchers over the world are trying to tap the sources of energy that are inexhaustible, cheap, absolutely pollution free and specially suited to deserts and isolated places (Rezaiyan & Cheremisinoff, 2005). Syngas generation technology from solid energy carriers with very small amount of CO$_2$ as by-product is an implementation of the Kyoto Protocal, for the arrival of commercially available fuel cell technology in combination with rising prices for oil and natural gas. Souza-Santos (2004) in his study on energy sources, demonstrated that one of the attractive technologies for fuel from coal waste and biomass is called gasification and the equipment used in the gasification process is commonly referred to as Gasifier. Gasification is a process of converting carbonaceous materials through incomplete combustion to combustible gases consisting of carbon monoxide, hydrogen, small amounts of carbon dioxide and methane etc (Pinto et al., 2003 & Hernández et al., 2010). This gas mixture is commonly known as a producer gas or syngas.

1.2 Objective

Due to abundant availability of coal and biomass in Tanzania (and elsewhere in the world) pull towards new technologies and economically viable methods have to be developed to process these feedstock. Development of gasifier for production of syngas from coal waste and biomass could improve Tanzania energy security in agricultural sector. Therefore, development and fabrication of gasifier that can accommodate a wide range of feed stocks, including coal waste and biomass is the main focus of the research paper, because together with the possible high syngas conversion efficiencies, most coal gasification projects have suffered from good control of temperature in the reactor, usually low volumetric energy density, widely distributed occurrence, heterogeneous solid fuels composition, high ash and un-favourable H$_2$:CO ratio after gasification.

2.0 Method and Materials

Reaction kinetics of a FBG involves solid particles that are exposed to gas of highly fluctuating chemical composition due to strong exchange of gas between the emulsion phase and the bubble phase by means of gas flow through the bubble and bubble coalescence and breakup (Silaen & Wang, 2010). Hydrodynamic studies provide informations on the basic flow pattern, mixing, particle attrition behaviour, and mass and heat transfer for design of the FBG. The flow diagram of the gasifier model development is as depicted in Figure 1.

A study of hydrodynamics and mixing characteristics of selected coal waste and biomass materials in a laboratory fluidized bed reactor were conducted to design a fluidized bed gasifier. Main focuses being to model coal waste or biomass interactions with bed materials and setting operating parameters for the gasifier.
Feedstock of 46.4% carbon content, 6.32% hydrogen content, 34-38% volatile matter, moisture content between 23 and 27% and more than 1% sulphur was used in the experiment. Since the operation temperature is below 900°C, thus high temperature resistant steel may not required leading to minimized energetic losses and a simple reactor design for cost reduction compared to conventional gasification reactors was used.

The gasifier is 300mm Internal Diameter (I.D) and 1200mm high from the level of air distributor (Fig.2). The distributor configuration is designed to promote the re-circulation of gas and solid in the bed zone as well as to prevent the bed aggregation. K-type thermocouple probes were installed to continuously monitor the temperature variation along the height of the gasifier, in which two thermocouples were located below and within the bed zone. One was located along the freeboard zone and one at the exit before the cyclone. These thermocouples are used for the control of temperatures in the gas distributor, the fluidized bed and the freeboard. The air is sent to the air accumulator of the gas distributor, where it is fed through 72 holes (2.5mm ID), that are spaced in 25.4mm along the surface of the gas distributor. The air is generated using an air blower.
Figure 2
FBG Test rig components

1-Fluidized bed reactor  2-Air blower with flow control valve 3- Air accumulator (plenum)  
4-VS-Motor  5-Hopper 6-Cyclone 7-Thermocouples 8-Screw conveyor  9-Gas sampling port (GSP) 10-Manometers

About 25kg of silica sand was put into the gasifier to serve as the bed material. Ambient air was fed into the reactor at the flow rate corresponding to the superficial velocity at 1.8m/s, which was above the predetermined minimum fluidization velocity, i.e 1.2m/s at ambient temperature. The superficial velocity was equivalent to an air flow rate of 74.3kg/h. The bed material was then heated by Liquefied Petroleum Gas burner until the bed temperature reached approximately 600°C. With the fixed flow rate of air, feedstock started to be fed into the reactor at controlled feed rates in the range of 32-54kg/h. The rates were varied to achieve a wide range of air to fuel rations. During the experiment, the pressure in the reactor was controlled to be slightly lower than ambient in order to prevent any gas leakage. The temperature profile along the height of the reactor was continuously monitored and recorded for analyses.

The main stream of the product gas went through the gas cleaning section (cyclone) and burnt at the flare. Some of the product gas was continuously sampled to analyze its compositions. To finish the experiment, the raw material feed rate was decreased gradually so that all the remaining raw material/char were burnt out. The reactor was left to cool down naturally.

Flow of air is controlled so that only about 25% of the incoming raw materials are burned in the bed to rise and maintain the temperature at 850°C, thus high temperature resistant steel may not required leading to minimized energetic losses and a simple reactor design for cost reduction compared to conventional gasification reactors was used. The remaining material is decomposed into gas. Once the bed is preheated, no more fuels are needed because the feedstock reactions supply all the required heat. The heat produced during exothermic processes is stored (accumulate) in the bed material and due to intense mixing of the bed inventory (fluidization) it is transferred to the processes that require heat input. In this way large temperature peaks in the oxidation zone are avoided and a nearly uniform temperature distribution was observed in the reactor. The outgoing product gas is then sent to a Gas Sampling Port (GSP), while the analysis of the permanent gas is performed by an analyzer.
equipped with a detector for CO, CO$_2$ and CH$_4$ detection and a thermo conductibility detector for H$_2$. The apparatus is equipped with two manometers to measure the flow through the orifice and the flow through the vessel. The orifice manometer is used to calculate the gas velocity through the reactor, while the vessel manometer is used to calculate the pressure drop over the bed.

3.0 Design and Development of the FBG

The prime reason behind the attempt to develop gasifier was its capability to produce producer gas suitable for heat applications as the feedstock is found nearby proximity. The first step was to find a feasible design, which could be taken as the basis for the gasifier construction. It was also decided that, at this stage, syngas would be burnt in a simple flare. The main body of the gasifier was divided into four pieces for easy fabrication of the internal parts. These four sections are hopper (fuel chamber), middle cylinder (reaction chamber), cyclone (entrained particles separator) and plenum (air distributor). For reliability, heat resistance and experimentation, it was planned to build with stainless steel. Drawings for different parts and the full gasifier were developed with solid works drawing package. After developing solid works model of gasifier, using Computational Fluid Dynamic (CFD) technique, the gasifier was simulated for optimum operating parameters before fabrication. Non-metal items included sealing material/gaskets, thermal resistant cement (refractory castable 160 LC cement, service temperature 1600°C), insulating tape etc were collected. The electrical equipment included an air supply blower with duct and control system, thermocouples and gas velocity measuring speed-gun were also made available.

3.1 Operation and Testing of Gasifier

The gasifier was first fired using coal wastes as feedstock. Next five days it was tested with different biomass as feedstock. During the runs, it was possible to obtain stable flame for more than 60 minutes. Values measured during the testing were recorded as shown in figures 4 to 6.

Before starting, all the parts of the gasifier were properly tightened and it was placed on open space. The blower for air supply was kept off. The air supply was full at the starting of gasifier and then maintained around 40% of the stoichiometric condition to ensure the partial combustion of the coal waste mixture with the help of a blower-controller, gate valve and air flow meter. After 5 minutes, the producer gas in the form of thick white smoke came out through the flare. The producer gas was ignited with a firing-torch at the flare.

Testing of the air-blown gasification tests were performed according to the ISO 9096 standards. Measuring principle was based on discontinuous sampling of gas stream in the laboratory atmospheric gasifier situation, whereby the gas was sampled in specific period of 10min through the sampling port in which the flow rate was maintained with the aid of process pressure. The sampling line consists of sampling probe, three ports and shut-off-valves in non-isokinetic sampling principle for temperatures more than 350°C for resistant to process temperature.

Static pressure difference and dynamic pressure (pressure velocity) were measured by Pitot-tube, while estimation of the product gas generation rate was done by applying mass balance. To obtain pressure velocity, static pressure was subtracted from total pressure readings in the Pitot tubes. The sampling flow was controlled by monitoring adjusting air flow meter. Then the gas velocity at sampling point was obtained by:

\[
\quad u_a = \frac{2\Delta p_{pt}}{\sqrt{\rho_a} P_a} = u_{ma} = \frac{q_{wa}}{3600A}
\]  

(1)
Where:

\( u_a \) = gas velocity

\( P_{pt} \) = Pitot Pressure

\( q_{ua} \) = gas flow rate

As a control, the product gas flow rate for the gasifier was calculated from carbon based mass balance and nitrogen based method. The nitrogen content in the product gas was taken as difference of 100\% of all other domain gas components namely, CO, H\(_2\), CO\(_2\) and CH\(_4\) in which,

\[
q_{u,g} = q_{u,a} \times \frac{N_{2,a}}{N_{2,g}} \tag{2}
\]

Where, \( q_{u,g} \) is the producer gas flow rate (m\(^3\)/hr),

\( q_{u,a} \) = The gasification agent flow rate (primary air flow rate) (m\(^3\)/hr)

\( N_{2,a} \) = The nitrogen content of gasification agent (vol. \%)

\( N_{2,g} \) = The nitrogen content of producer gas (vol. \%)

For the product gas yield, carbon based method was applied, which is based on a total carbon balance, provided the main gas compositions, fuel feeding rate, fuel moisture content, carbon content together with bottom ash and fly ash are known as follows:

\[
q_{mf,cc_f} + \sum q_{u,a,cc_a} = q_{ug} \times \left(CC_g + C_tcc_t + C_pcc_p\right) + q_{mash,ccash} \tag{3}
\]

Where: \( q_{m,f} \) is the fuel feeding rate (kg/hr)

\( cc_f \) is the carbon content of fuel (kg C/kg dry coal) = 0.48 for coal wastes

\( q_{u,a} \) is the gasification agent feeding rate (m\(^3\)/hr)

\( cc_a \) is the carbon content of gasification agent (kg/m\(^3\))

\( q_{u,g} \) is the producer gas generation rate (m\(^3\)/hr)

\( cc_g \) is thecarbon content of non-condensable gases (CO, CO\(_2\) and CH\(_4\)) in producer gas (kg C/m\(^3\))

\( c_t \) is the tar concentration in producer gas (kg/m\(^3\))

\( cc_t \) is the carbon content of tar (kg C/kg tar)

\( c_p \) is the particle concentration in producer gas (kg/m\(^3\))

\( cc_p \) is the particle content of particles (kg C/kg dry coal)

\( q_{m,ash} \) is the bottom ash rate (kg dry ash/hr)

\( cc_{ash} \) is the carbon content of bottom ash (kg C/kg dry ash)

For the coal wastes \( CC_f = 0.46 \) and \( CC_a = 0 \), the product gas yield is:

\[
Y_g = \frac{q_{uy}}{q_{m,f}} = \frac{CC_f}{CC_g + C_tcc_t + C_pcc_p} \times c_{ash,f} \times cc_{ash} \tag{4}
\]

Where \( Y_g \) is the producer gas yield (m\(^3\)/kg dry coal)

\( c_{ash,f} \) is the ash content of the fuel (kg C/kg dry coal)

The product gas flow rate was calculated by using carbon based method on dry basis. Multiplication of the gas yield by the fuel feeding rate provides the actual product gas flow rate:

\[
q_{u,g} = Y_g \times q_{m,f} \tag{5}
\]

Where:

\( q_{u,g} \) = product gas flow rate
According to Pan et al. (2000) and Kumabe et al. (2007), the relevant formula used for the cold gas efficiency of the FBG (experimental) is:

\[
\eta = \frac{H_g \times \Delta T}{H_s \times M_s} \times 10t0\% \tag{6}
\]

Where:
- \(H_g\) = Lower heating value (LHV) of syngas
- \(H_s\) = LHV of solid fuel
- \(Q_g\) = Volume flow rate of gas
- \(M_s\) = Solid fuel consumption

Using the obtained values, efficiencies and thermal power were found as follows:

\(\eta_g\) = Overall cold gas efficiency = 66.66%

\(\eta_{th}\) = Thermal efficiency of the gasifier = 90.1%

\(P_g\) = Thermal power of the syngas, \(Q_g \times H_g = 0.0256 \times 4400 = 111.87kW\),

\(P_{th}\) = Thermal power consumption (full load) and \(P_g/\eta_{th} = 111.87/0.90 = 124.36kW\).

4. Results and Discussions

1kg of feedstock in 25kg of silica sand as the bed material were fed into the reactor at the flow rate corresponding to the superficial velocity of 1.8m/s, and air flow rate of 74.3kg/h produced syngas with 20 (%v/v) of carbon monoxide, 14 (%v/v) of hydrogen and small amounts of methane and carbon dioxide. It was found that high temperature favoured hydrogen production in the process due to the endothermic nature of most of the gasification reactions. Similar effects of temperature were found by (Kumar et al., 2009). Pan et al. (2000) observed a reduction in carbon monoxide composition with increase of coal in the feedstock when mixtures of pine chips by using Black coal and Sabero coal as feedstock for gasification in a fluidized bed gasifier, with temperature of the system at the range of 840-910°C.

The change in the temperature may be explained by the influence of the feed-to-oxidizer ratio and the geometric length of the gasifier that contains the pyrolysis, combustion and reduction zones. It is a fact that these three zones (pyrolysis, combustion and reduction zones) exist virtually inside the gasifier, which is not heated externally. The length of these zones also varies depending upon the thermal degradation behaviour and duration (residence time) of the feedstock. Therefore, the gasifier’s height and approximate length of the three zones are decided according to the thermal degradation behaviour of the feedstock and its approximate residence time in each zone. In the present study the temperature of the reduction zone became similar to the combustion zone and caused an overall reduction of the temperature. Most of the published studies on co-gasification used external heaters to maintain gasifier temperature (Kumabe et al., 2007).
In this study carbon-dioxide is observed to be very small compared to other literatures due to the fact that the non-agglomerating coal waste and biomass were used in the gasification process, therefore no agglomerated layer of material was formed, then it increased the flow of fuel hence the oxygen/fuel ratio decreased causing a limited combustion in the reduction zone and hence low carbon dioxide in the syngas. Melting of inorganic ash in the study was controlled by lowering the bed temperature and making it less than the melting points of the ash constituents in the feedstock. This was done through reduction of the air intake into the system.

The effect of gasifying temperature on product gas composition is shown in Figure 3. It was observed that an increase in temperature brings about an increase in the concentration of CO and H₂ from feedstock. The hydrogen concentration in the syngas was lower due to lower air/fuel ratio for the air gasification conditions and the limited of stem resulting in a reduction of hydrogen formation via the water-gas shift reaction. The bed temperature was significantly high enough to favour the relatively higher concentration of carbon monoxide.

The results obtained by sampling directly from the bed clearly showed inter-bed variations in the composition. With increasing height above the distributor, there is a steady increase in both CO and H₂ (figure 3), reaching a maximum near to the top of the bed. However the measured concentration of carbon dioxide (CO₂) was observed to initially drop, reach a minimum at some location within the bed, and then to rise again to levels similar to those measured immediately above the distributor. Between 5cm and 10cm, Boundourd reaction $s_c + co_2 \leftrightarrow 2co + increases production of co$, consequently, the CO₂ concentration diminishes. The drop of carbon monoxide at 50 cm in the graph is due to the gas phase equilibrium $co + h_2o \leftrightarrow co_2 + h_2$ which shifts to the right until the gases are ejected into the freeboard and hence increasing hydrogen formation in the reactor.

Figure 3

*Variation of syngas gas (%) with height above grate (cm)*

![Graph showing variation of syngas gas with height above grate.](image)

In all cases, more than half of the synthesis gas volume was N₂, i.e. 52-58%, and CO was more than CO₂. To eliminate the dilution effect of N₂, which may lead to a misinterpretation of the gas product formation, the concentrations of each gas were converted into the N₂-free basis.
Flame temperature and duration of produced gas from the feed stocks have been presented in figure 4. Symgas flame temperature and flame shapes show maximum flame temperature of 970.60°C after 30 minutes then the temperature gradually drops to 324.60°C after 1hr when the fuel feeding stops at different mixing ratio of kaolinite. The average flame temperature of about 770°C is sufficient to meet the requirements of many firing applications.

Figure 4
(a) Variation of syngas flame temperatures (b) Duration on flame temperature for 60min

The effect of feeding rate on product gas flame is shown in figure 5, while the variation of fluidization velocity with the flame shape has been depicted in figure 6 (at Ø = 0.4).

Figure 5
Variation of Product Gas Flame Shape with Fuel Feeding Rates at (a) 12.3kg\textsuperscript{\text{-1}} (b) 13.6 kg\textsuperscript{\text{-1}} (c) 14.4 kg\textsuperscript{\text{-1}} (d) 15 kg\textsuperscript{\text{-1}}
4. Conclusions and Recommendations

4.1 Conclusions

The development of the gasifier was done in which 1kg of feedstock (at 15% moisture content) produces 2.0-2.5 Nm$^3$ of product gas, with approximately 1.027kg/Nm$^3$ cool and clean gas density. The produced gas has an energy content of approximately 5-6 MJ/m$^3$. Results illustrate the potential of the developed FBG to produce good amount of syngas at a range of 18-20vol % for carbon monoxide and 13-16 vol% for hydrogen as major components of syngas produced with a stable gas flame, hence demonstrates in an improve of fuel flexibility. The results also promote further utilization of coal wastes and biomass that could provide a cheap, reliable, and low carbon means for producing syngas for fuel flexibility in Tanzania.

4.2 Recommendations

Further work to initiate the combustion in the oxidation zone is needed for the developed gasifier.

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