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Fuel gas generation in a lab-scale dual fluidized bed rice husk gasifier

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A B S T R A C T

Dual fluidized bed gasification (DFBG) technology is a promising option to produce medium calorific value N_2-free fuel gas. The present study deals with the development of a self sustaining lab scale are shown below DFBG consisting of a bubbling fluidized bed gasifier and a fast bed combustor, coupled through L-valves. Stable operation of this set up can be achieved by supplying liquefied petroleum gas to the combustor to generate N_2 free syngas in the gasifier. Rice husk was used as the feedstock, steam as the gasifying agent. Experimental investigation includes a parametric study of process parameters such as gasifier temperature and steam flow rate. H_2 concentration increases with higher gasification temperature and higher steam flow. CO concentration increases with increasing gasification temperature but decrease with higher steam flow. Higher heating value of fuel gas, char conversion and cold gas efficiency has been analyzed as performance indicators of the system.

1. Introduction

The primary thermal conversion technologies for biomass to electricity are combustion, gasification and pyrolysis. An overview of existing technologies is described in Bain et al. [1998]. Gasification combined with a gas engine/turbine has the advantage of having higher electric efficiency and lower cost of electricity generation than direct combustion [Bridgewater et al.1995]. Biomass gasification is carried out in a fixed bed, fluidized bed, moving bed or special design reactors [Bolhàr-Nordenkampf et al. 2004]. The gasification reactions are endothermic, energy is supplied by partial combustion of the feedstock in the gasifier with a sub-stoichiometric amount of air as gasification agent. Such air gasification produces a poor quality gas with a higher heating value (HHV) around 4-7 MJ/m^3. Oxygen and steam blown processes produce syngas with HHV in the range of 10-18 MJ/m^3 [Schuster et al.2001, Doherty et al. 2008, Doherty et al. 2009]. Fuel gas of medium HHV can be produced by using the dual fluidized bed gasifier (DFBG) system [Schuster et al. 2001]. This system has two fluidized bed reactors coupled together for circulation of bed materials. The combination of a bubbling/turbulent fluidized bed gasifier and a fast bed combustor was reported to facilitate gasification reactions and suppress the tar production [Xu et al.2006].

Hydrodynamic study of the present DFBG system has been carried out by [Karmakar et al. 2010]. This system is similar to the system as described in Loffler et al. [2003] and Kaiser et al. [2003] except for the mechanism of solids transfer between the vessels. L-valves have been installed in the transfer pipes instead of loop seals. Like loop seals, L-valves are also non-mechanical devices and require aeration to generate a solids flow. L-valves have the additional advantage of being inexpensive, easy to fabricate and have no moving mechanical parts.

Rice husk is an agricultural residue that offers reasonably high energy content (12-18 MJ/kg). However, for gasification using rice husk as a feedstock, it becomes more difficult for its physical and chemical properties. Grate furnaces and downdraft gasifiers are inefficient for rice husk conversion to energy for high ash content, low bulk density, poor flow characteristics and low ash melting point. Fluidized bed reactors seem to be the better choice for extracting energy from rice husk through gasification process.

The present study aims to develop a lab scale dual fluidized bed rice husk gasifier for production of hydrogen rich fuel gas. Steam was used as the gasification agent to produce hydrogen rich fuel gas. Using rice husk as feedstock, the effects of operating parameters like the gasifier temperature and the steam feed rate on the composition of fuel gas have been studied.

2. Materials and methods

2.1 Rice husk as the feedstock:

Rice husk has a high (19.52 %) ash content and a major part of the ash is silica. Therefore, a circulating fluidized bed gasifier is a promising choice for rice husk, as the bed temperature can be kept below the ash melting temperature. The proximate and ultimate analyses of rice husk are presented in Table 1.

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Table 1: Ultimate and Proximate Analysis of rice husk

Ultimate Analysis		Proximate Analysis	
Components	Percent	Components	Percent
Carbon	38.43	Volatile matter	55.54
Hydrogen	2.97	Fixed Carbon	14.99
Sulphur	0.07	Moisture	9.95
Nitrogen	0.49	Ash	19.52
Oxygen	36.36		
Ash	21.68		
HHV = 15.68 MJ/kg			

From the ultimate analysis and on the basis of dry and ash free condition, rice husk was represented by the molecular formula $CH_{0.93}O_{0.71}$ and its molecular weight was calculated to be 24.29 kg/kmol. Rice husk can be difficult to fluidize because of its cylindrical shape and non-granular and flaky nature. The fluidization behaviour was improved by adding small particles like silica sand.

2.2 Bed materials

Silica sand was used as bed material due to its availability and high heat retention capacity. It also helps to maintain a uniform bed temperature by improving the fluidization behaviour of the rice husk in the gasifier. The sand was sieved to obtain mean diameter of 0.211 mm. The cumulative distribution of the particle diameter is shown in Fig 1. The density of the samples was experimentally found to be 2645 kg/m^3.

2.3 Experimental setup

A schematic view of the DFBG test set-up is shown in Fig 2. The system consisted of a bubbling fluidized bed gasifier, an inclined transfer pipe fitted with the L-valve, a fast bed combustor, a cyclone separator and a down comer equipped with another L-valve. The dual bed system was placed on a steel structure to allow expansion in the upward direction while operating at high temperatures.

The gasifier was cylindrical in shape with an inside diameter of 0.1 m and a height of 1.8 m. The main body was made of mild steel and the inner surface was lined with refractory material of 0.1 m thickness to prevent heat losses. Steam from the boiler was passed through a single perforated distributor plate to fluidize the gasifier bed. The temperatures were recorded using Ni-Cr-Ni thermocouples. The biomass feeding system consisted of a storage hopper and two screw feeders. The hopper was placed on top of the upper screw feeder and filled manually. A lock hopper was connected to prevent the bypass of product gas from the gasifier. The upper screw feeder was used to supply the required feed to the lower feeder by means of a variable speed drive. The lower screw feeder operated at higher speed to push the rice husk directly into the gasifier bed. It was also water-cooled to avoid pyrolysis outside the gasifier. A 25 mm diameter transfer pipe was used to transport the bed materials and the char from the BFB gasifier to the CFB combustor. The pipe was made of stainless steel and insulated externally to prevent heat loss. Steam was injected through a 6 mm port to the L-valve to move the solid char into the fast bed combustor.

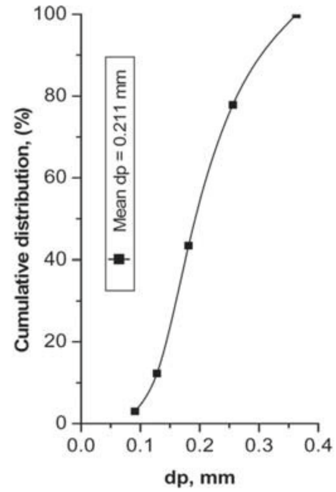

Fig 1 Cumulative particle size distribution of the silica sand used

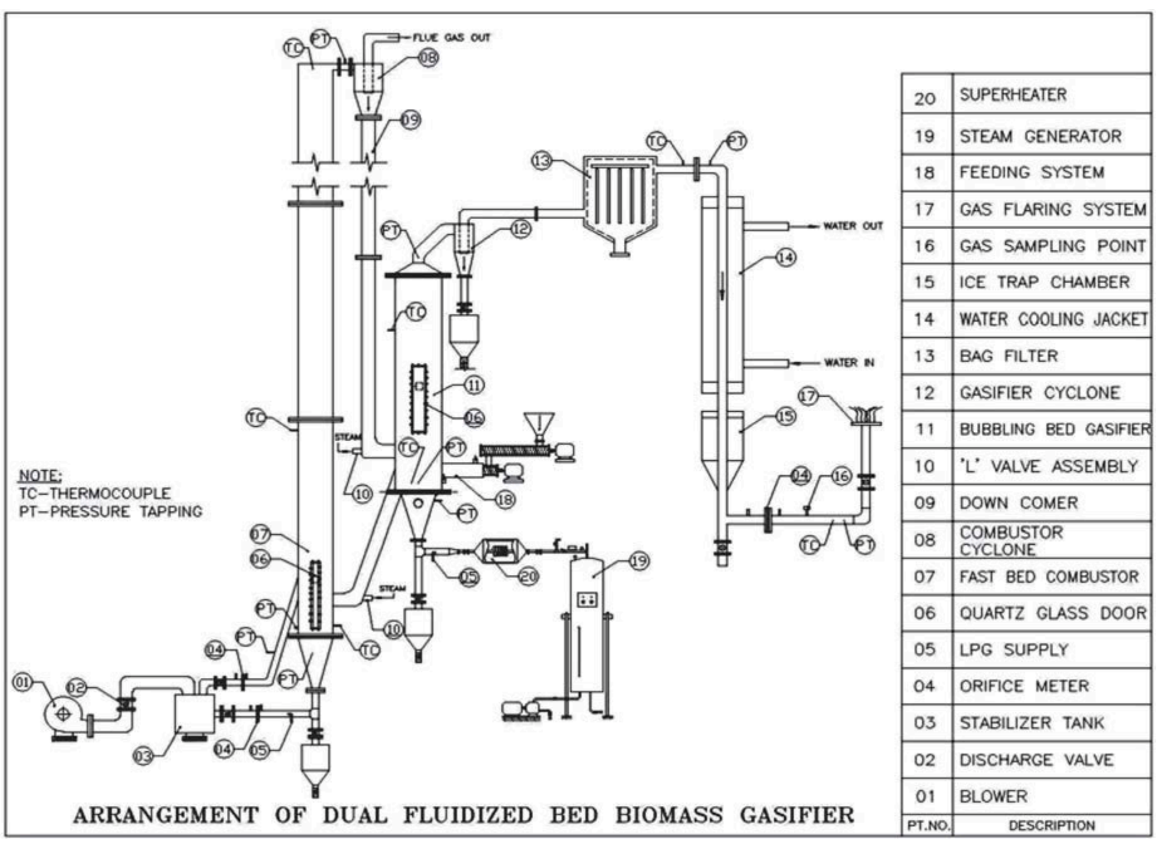

Fig 2 Flow diagram of DFBG developed at CSIR-CMERI Durgapur

The fast bed combustor was a 5.95 m tall vertical pipe, 50 mm in diameter. The combustor was internally lined with refractory material to prevent heat losses. A single perforated 3 mm thick SS distributor plate was used at the bottom of the riser. Air was supplied at two locations of the combustor – the primary air (PA) at the bottom and the secondary air (SA) at a height of 0.3 m above the distributor plate. The function of the PA was to maintain bubbling conditions in the dense bed at the bottom. The SA generated pneumatic transport of hot bed particles to the exit of the combustor. Ni-Cr-Ni thermocouples were used for measuring the temperature. Two view ports fitted with quartz glass allowed viewing the inside of the combustor during operation. A cyclone separates the hot bed material from the flue gas-solid mixture to move into the down comer, 25 mm in diameter and 4 m in length. The inside of the stainless steel down comer was lined with refractory to prevent heat losses from the hot sand materials. Steam flow through an L-valve placed in the down comer allowed to control the transfer of bed materials to the gasifier.

An LPG connection was provided in both plenum chambers of the combustor and the gasifier and ignition torches were used to ignite the air-LPG mixture for preheating the system during start up. Steam required for the chemical reactions in the gasifier was obtained from a electrically heated boiler. The steam was further superheated to 200°C in a heat exchanger to maintain the gasifier bed in bubbling conditions.

The product gas from the cyclone was made dust-free and cleaned by passing it through a bag filter. After the bag filter, a water cooler and ice trap system were used to capture tar. An orifice meter was positioned to measure the gas yields before sampling. The pressure drop across the orifice was measured using a micro manometer. The orifice was calibrated prior to the experiments with two reference instruments, a digital micro manometer (Furnace Control, England) and a thermal anemometer (Dantec, Denmark). During the experiments, the flow rate measurements were corrected by a temperature factor to obtain the flow rate in NTP. The sampling system consisted of gas sampling probes fitted with a septum located downstream of the orifice meter. The stainless steel probes were 12.5 mm in diameter and 50 mm in length. A 10 ml syringe was used to collect the dry and clean product gas sample for analysis in a Gas Chromatograph (Chemito, model – GC 1000). A Thermal Conductivity Detector (TCD) and a Flame Ionization Detector (FID) were used as detectors. Standard gas mixtures were used for calibration. Helium was used as carrier gas.

2.4 Major gasification reactions

The biomass primary pyrolysis is described by the following global reaction [Di et al. 2000, Sheth et al. 2006].

$$Biomass \rightarrow w_C Char + w_G Gas \ + w_T Tar \tag{1}$$

The weight fractions of char, gases, and tar produced, as well as the composition of the gases produced depend on the type of biomass used (composition and structure), but also on the heating rate and the residence times [Nicolas et al. 2011].

Tar can undergo so-called secondary pyrolysis described by the following reaction and the composition of the gases produced is found experimentally. The secondary pyrolysis is slightly exothermic [Nicolas et al. 2011].

$$Tar \rightarrow w_{CH_4} \ CH_4 + w_{CO} \ CO + w_{CO_2} \ CO_2 \tag{2}$$

Different gas phase reactions account for the combustion of tar, CH_4, CO, and H_2 [Bryden et al. 1996] are shown below.

$$\begin{aligned} &Tar + O_2 \rightarrow CO + H_2O \\ &CH_4 + 1.5O_2 \rightarrow CO + 2H_2O \\ &2CO + O_2 \rightarrow 2CO_2 \\ &2H_2 + O_2 \rightarrow 2H_2O \end{aligned} \tag{3}$$

The combustion reactions are highly exothermic and can provide the heat required for the primary pyrolysis. The water gas shift reaction is accounted for and is described by [Yoon et al. 1978, Biba et al. 1978].

$$CO + H_2O \Leftrightarrow CO_2 + H_2 \tag{4}$$

Finally, the combustion of char and the gasification of char by CO2, H2, and H2O are described in [Groeneveld et al. 1980, Kashiwagi and Nambu. 1992].

$$\begin{aligned} &C + O_2 \rightarrow CO_2 \\ &C + CO_2 \rightarrow 2CO \\ &C + 2H_2 \rightarrow CH_4 \\ &C + H_2O \rightarrow CO + H_2 \end{aligned} \tag{5}$$

The char combustion is exothermic, whereas the char gasification reactions are endothermic.

2.5 Operating methodology

The fundamental idea of a dual bed gasification system is to physically separate the gasification reaction and the combustion reaction in order to get a largely nitrogen-free product gas.

During start up, adequate quantity silica sand was fed through the screw feeder arrangement. The primary air was introduced at the bottom of both reactors to operate the beds in bubbling/turbulent conditions and subsequently the secondary air supply in the fast bed combustor for pneumatic transport of the bed materials. The air supply through the L-valves was then opened to ensure transport of sand between the reactors. After reaching steady circulation, the LPG-valve was opened and the air-LPG mixture was ignited. The air flows were adjusted to control the circulation of hot sand in the system. Both the gasifier and the combustor were allowed to run to preheat the refractory linings until the bed temperature was around 850 °C. The feeding of rice husk was then started and the LPG supply to the gasifier was slowly turned off. The fluidizing air to the bubbling bed gasifier was gradually replaced by superheated steam, similar for the flow in the L-valves. The LPG supply to the combustion chamber was not shut off immediately to maintain the combustor temperature. LPG was often required in the combustor to compensate for the heat losses through its large surface area. The heat required for the gasification of the rice husk was supplied by the circulating sand. The non-gasified char particles were burnt in the fast bed combustor after being transported from the gasifier. The return of the bed materials to the gasifier was controlled by the steam injection through the downcomer L-valve. The steam flow in the bubbling gasifier and the air flow in the fast bed combustor could be adjusted within a given range. The plant was operated steady state for half an hour before collecting the gas samples.

In the dual-bed set-up studied, the rice husk feed rate was 5.0 kg/hr and the steam flow rate to the gasifier was varied between 1.5 to 2.8 kg/hr. The fresh rice husk was fed at the bottom of the bubbling bed gasifier to prevent segregation of solids of varying density in the gasifier. In the fast bed combustor, on the other hand, the rice husk char inlet was placed 200 mm above the distributor plate, but below the secondary air injection.

The total air flow rate in the combustor was in the range of 7.9 to 8.8 kg/hr, which was necessary to achieve complete combustion of the char and maintain the superficial velocity higher than the terminal velocity of silica sand (1.54 m/s). The primary air in the bottom zone of the combustor ensured proper fluidization of char-sand mixture, whereas the secondary air served to carry the hot sand particles. The combustor temperature was maintained in the range of 850-950 ^{0}C by combustion of the solid char coming from the gasifier.

Controlling the system temperature is crucial to avoid problems resulting from agglomeration of the ash, sand and char mixture and subsequent blockage of flow through the system, especially because rice husk was used which contains 21.68 % ash, higher than any woody biomass, and the ash contains more than 95% silica. One of the important features of rice husk gasification is that the bed temperature can be kept as low as 600-650 °C, thereby preventing sintering and agglomeration of this ash which would otherwise cause serious operational problems [Mansaray and Ghaly. 1999]. The upper temperature is fixed by slagging phenomena which primarily depend on the ash composition and the reaction atmosphere (like oxidation or reduction). Above this temperature, silica and potassium oxides in the ash fuse on the surface of the rice husk char particles forming a glass-like barrier that prevents further reaction of the remaining carbon [Natarajan et al. 1998]. Some studies [Kaupp 1984, Schiefelbein 1989] also indicate that oxidation of rice husk at temperature higher than 700 °C results in a physical structural transformation of silica from its original amorphous state to a crystalline state thereby encapsulating residual carbons. Once structural changes of silica occur, the remaining carbon becomes unavailable for further oxidation reactions, even at higher temperatures. Therefore, the gasifier was operated in the range of 640-750 °C.

3.0 Results and discussion

Table 2 shows the fuel gas composition, heating value, gasifier char conversion and cold gas efficiency obtained with the dual fluidized bed gasifier at various operating conditions. The fuel gas compositions were corrected for the presence of 4-7% of nitrogen due to leakage. The main components were H_2, CH_4, CO and CO_2. Steam being the major source of hydrogen during gasification, the hydrogen content was higher than typical in all test runs.

3.1 Effect of the gasifier temperature

The gasifier temperature affects the syngas composition to a large extent. The gasifier temperature was varied by controlling the air flow and LPG supply in the combustion chamber. The maximum temperature in the gasifier of 750 °C was reached when the combustor was operated at 952 °C. Figure 3 shows the syngas composition obtained at gasifier temperatures in the range of 640-750 °C with a steam flow rate of 2.2 kg/hr. It is seen that the H_2 and CO concentrations increase with increasing gasifier temperature, while the CO_2 and CH_4 concentrations decrease. According to Le Chatelier's principle, higher temperatures favour the reactants in exothermic reversible reactions and the products in endothermic reversible reactions. Considering the endothermic steam gasification of char and the steam reforming reactions, higher H_2 and CO concentrations are indeed expected at higher temperatures. Through the exothermic water gas shift reaction and the endothermic Boudouard reaction, on the other hand, higher CO and H_2 concentrations are expected. A similar trend was reported by Fercher et al. [1998] and Herguido et al. [1992].

From the variation of volumetric CO concentration with temperature for the Boudouard reaction, it can be seen that high temperature favours CO formation.

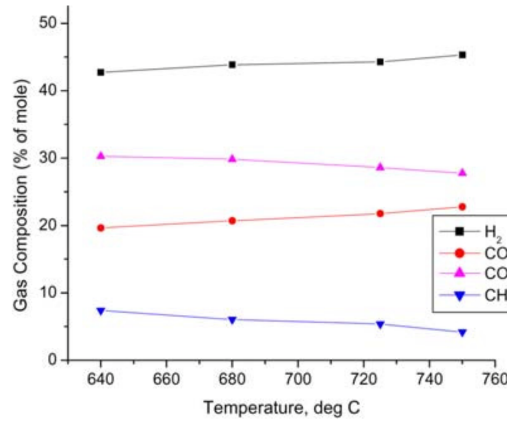

Fig 3 Gas composition vs. gasifier temperature at a steam flow of 2.2 kg/hr

Table 2 shows the operating parameters of the dual bed system at different operating conditions. The dry gas yield increases from 0.90 Nm^3/kg at 640 °C to 1.13 Nm^3/kg at 750 °C at constant steam flow rate of 2.2 kg/hr. Table 2 also shows that the higher heating value (HHV) of the syngas, computed from the measured gas analysis, decreases with increasing gasifier temperature from 10.88 MJ/Nm^3 at 640 °C to 10.33 MJ/Nm^3 at 750 °C.

Table 2: Experimental data at different operating conditions

Expt. run no,	1	2	3	4		5	6	7	8
Biomass Feed (kg/hr)	5	5	5	5		5	5	5	5
Steam rate to gasifier (kg/hr)	2.2					1.5	1.9	2.4	2.8
Gasifier temp. (°C)	640	680	725	750		700			
Combustor air supply (kg/hr)	7.9	8.5	8.5	8.6		8.4	8.4	8.5	8.8
LPG supply (lpm)	0.90	1.00	1.04	1.20		1.12	1.16	1.18	1.22
Combustor temp. (°C)	848	882	935	952		895	900	912	918
Gas Composition (mole %)									
H_2	42.72	43.84	44.28	45.32		43.24	44.11	45.04	45.40
CO	19.64	20.70	21.75	22.74		24.32	22.53	20.66	19.78
CO_2	30.26	29.42	28.59	27.76		26.04	27.24	29.05	29.88
CH_4	7.38	6.04	5.38	4.18		6.4	6.12	5.25	4.94
HHV of prod. gas (MJ/Nm^3)	10.88	10.62	10.55	10.33		11.15	10.92	10.46	10.27
Dry gas yield (Nm^3/ kg raw biomass)	0.90	0.97	1.05	1.13		0.93	0.96	1.04	1.10
Gasifier char conversion (%)	24.64	27.99	31.84	35.82		31.52	30.15	29.95	30.33
Cold gas efficiency (%)	58.55	61.24	65.67	68.53		61.17	61.71	63.93	66.24

3.2 Effect of the steam flow rate

The steam flow rate was varied between 1.5 – 2.8 kg/hr with a biomass feed rate of 5.0 kg/hr. Figure 4 presents the syngas composition for a gasifier temperature of 700 °C. It shows that the H_2 and CO_2 concentrations increase with increasing steam flow rate, resulting from the gasification of char by steam and the conversion of hydrocarbon species and carbon monoxide respectively through steam reforming and water gas shift reactions. The CO and CH_4 concentrations indeed decrease with increasing steam flow rate.

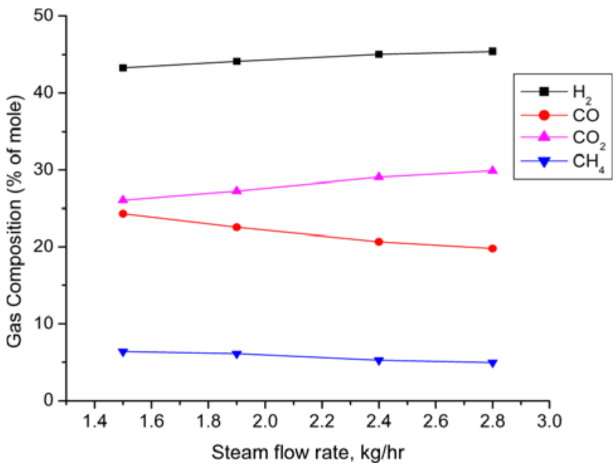

Fig 4 Gas composition vs steam flow rate at a gasifier temperature 700 °C

Table 2 indicates that an increase in the steam flow rate results in an increase in the dry gas yield from 0.93 Nm^3/kg at a steam feed rate of 1.5 kg/hr to 1.10 Nm^3 /kg at a steam feed rate of 2.8 kg/hr. The higher heating value (HHV) of the syngas decreases with increasing steam flow rate from 11.15 MJ/Nm^3 at 1.5 kg/hr to 10.27 MJ/Nm^3 at 2.8 kg/hr.

3.3 Char conversion in the dual bed gasifier

When a solid fuel is injected into a fluidized bed gasifier, the reactivity of the fuel char plays an important role. For a dual-bed system, a certain amount of char leaves the gasification zone without taking part in the chemical reaction. The gasifier char conversion in the dual-bed system determines its efficiency. When assuming that the biomass char is mainly converted by steam gasification and produces CO only, the char conversion can be estimated from:

$$\text{Char conversion in gasifier} = \frac{\text{Weight of carbon in CO present in the fuel gas}}{\text{Weight of carbon in raw biomass}}$$

Table 2 shows that char conversions in the range of 24.64 – 35.82 % were achieved.

3.4 Cold gas efficiency

The cold gas efficiency (CGE) can be used to evaluate the gasification performance. It is defined as the percentage of the heating value of the biomass fed that is recovered in the syngas. The auxiliary fuel LPG that is supplied to the dual fluidized bed system to maintain the combustor temperature has to be accounted for. The gasification efficiency can then be expressed as follows:

$$\text{CGE} = \frac{\text{Gas yield}\left(\text{Nm}^3\ \text{kg}^{-1}\ \text{fuel}\right)\ \text{x}\ \text{Syngas HHV}\left(\text{MJ Nm}^{-3}\right)}{\text{HHV of fuel}\left(\text{MJ kg}^{-1}\right) + \text{Heat addition through LPG}\left(\text{MJ kg}^{-1}\right)}$$

The gasification efficiency of the dual fluidized bed system for the different experimental runs is given in Table 2. Cold gas efficiencies in the range of 58.55 - 68.53% could be achieved. The cold gas efficiency is seen to increase with both reactor temperature and steam flow rate.

4.0 Conclusions

Lab scale DFBG was developed at CSIR-CMERI Durgapur and experimental investigation was carried out using rice husk as feedstock for generation of hydrogen rich fuel gas using steam as gasifying medium.

HHV of fuel gas was found to be around 11 MJ/Nm^3. HHV of fuel gas was found to decrease with increase of gasifier temperature and steam flow rate. Concentrations of H_2 and CO were seen to increase with higher gasifier temperature, while those of CO_2 and CH_4 were decreased. Higher steam supply resulted in higher concentrations of H_2 and CO_2 and lower concentrations of CO and CH_4. Cold gas efficiencies were estimated in the range of 58.55 to 68.53 % and the gasifier char conversion was around 30 %. It was observed that the recirculated char alone could not maintain the desired temperatures in the combustor and small quantity of LPG was required to compensate the heat loss.

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