Statistical Evaluation of the Process Parameters on the Biomass Gasification

M. Senthil Kumar* and S. Vivekanandan**

Abstract : The main objective of this paper is to study the effect of design and operating parameters, mainly as bed temperature (T), pressure (P), equivalence ration (R), feed rate (F) and particle size (S) on the performance of the gasification process of saw dust as biomass in a continuous fixed bed updraft reactor. In the present investigation, an empirical relationship was developed to predict the process of generating fuel gas with better quality through gasification of biomass in a fluidized bed reactor using response surface methodology (RSM). Six major components of the producer gas such as O_2, H_2, CO, CO_2, CH_4, and N_2 were analyzed in the laboratory along with the evaluation of tar yield and cold gas efficiency. It was observed that the concentrations of hydrogen, oxygen nitrogen and carbon monoxide were increased with rise in gasification temperature, pressure and equivalent ratio (0.2-0.35). On the other hand, higher equivalence ratios (0.4-0.5) caused to decrease the concentrations of hydrogen, oxygen, nitrogen and carbon monoxide. Higher equivalence ratio also resulted in more gas yields and cold gas efficiency due to increase in the exothermic reactions. Furthermore, it was observed that the CH_4 and CO_2 decreased with the increase of temperature and pressure. The developed model was made a good prediction for the experimental data as observed for the gas species concentrations.

Keyword: Saw dust; Fluidized bed gasifier; Producer gas; Response surface methodology and Equivalent ratio.

Highlights :
- An empirical relationship developed to predict the quality of the producer gas
- Response surface methodology (RSM) was used.
- Saw dust used as the feed stock
- The developed model was made a good prediction for the experimental data.

1. INTRODCTION

Biomass is the only source of carbon-based renewable fuels and the sustainable exploitation of this resource is essential to secure the energy security. An agricultural residue that could be utilized for the recovery of energy is saw dust because of its reasonably high energy content (12–18 MJ/kg) [1].

The fluidized bed gasifier allows efficient conversion of solid biomass into high quality product gas via steam gasification. The process is currently performed on wood chips from natural sources. In addition to biomass from forestry, woody residues, such as bark, waste wood, and sawdust are interesting feedstock because of their availability and price. Depending on the origin of the residues, the particle size of the materials covers a wide range from fine particles of sawdust The particle size of the biomass has an important influence on its thermal conversion and this has been studied by several research groups. It is commonly agreed on that the particle size determines the mass and heat transfer into and from the particle. Four different regimes of heat transfer are distinguished related to particle size: In the

*   Assistant Professor, Department Of Mechanical Engineering, Annamalai University, Annamalai Nagar, Chidambaram-608002,Tamil Nadu, India.

**   Associate Professor, Department Of Mechanical Engineering, Annamalai University, Annamalai Nagar, Chidambaram-608002, Tamil Nadu, India.

kinetic regime, particles are considered to be isothermal and to heat up instantly. In the thermally thin regime, external heat transfer occurs between the particle and the surroundings, but the intra-particle temperature gradient is found to be negligible. In the thermally thick regime, the reactions of the particles are controlled by external and internal heat transfer [2]. The gasification technologies are broadly of two types – fixed bed and fluidized bed. According to Ergudenler [3], the fluidized bed technology offers great advantages when used for gasifying low density biomass residues to generate low calorific value fuel gas. The conventional types of gasifiers are not suitable for conversion of saw dust because of its high ash content, low bulk density, poor flow characteristics and low ash melting point. The fluidized bed gasifier seems to be the suitable option for such conversion with better control of bed temperature in isothermal conditions. Although many countries are rapidly deploying biofuels worldwide, this is almost exclusively first generation biofuels technology. The countries are rapidly moving forward with developing and deploying biochemical technology, using the sugarcane and the corn. Sugar cane ethanol is generally regarded as having little to no impact on primary food supplies and prices, since Brazil has increased its production of sugar cane to more than offset the amount of sugar diverted to ethanol production. However, food supply and price concerns have been raised about corn ethanol production [4] because corn grain is an important food and animal feed commodity. The EU, the largest biodiesel producer, uses rapeseed oil as its main feedstock and again concerns about fats and oils supplies and prices have been raised over the diversion of rapeseed oil to biodiesel production. Because of these concerns and the overall limitations of first generation biofuels technology primarily due to competition with feed and food production, advanced or second generation biofuels technologies will be required to meet aggressive volume goals for biofuels deployment [5]. A number of different conversion technologies exist [6, 7] for the conversion of cellulosic biomass to biofuels. The predominant differentiation between the conversion options is the primary catalysis system [8].. Pyrolysis on the other hand, is the milder depolymerization of biomass producing a liquid intermediate (pyrolysis oil or ''bio-oil'') in the absence of oxygen at lower temperatures, typically in the range of 400–650°C. Good reviews of pyrolysis techniques and the current technical status of these techniques are provided [9]. but so far no such plants have been built in India, possibly because of low electricity prices and a lack of sufficient incentives to adopt renewable energy [10].Several studies have investigated equilibrium modelling of gasification and most of them used the relatively simple Gibbs free energy minimization method [11–13]. Ptasinski etal. [14] and Prins et al. [15] studied the effect of varying feedstock compositions on gasification efficiency. Mahishi and Goswami [160] used equilibrium modelling to study the effects of operating conditions on hydrogen yields using both steam and oxygen as gasifying agents. They found that wood should be gasified at ambient pressure, 1000 K, an equivalence ratio of 0.1 and a steam to biomass ratio of to obtain the maximum hydrogen yield; however, the effect of moisture was not included. A comparison of their equilibrium calculations with experimental data showed that the data correlated best at longer residence times (>1.4s) and temperature above 800°C. Although experimental data is available in literature for bagasse gasification, the gasifiers were not necessarily optimized for the specific downstream application of FT synthesis. In addition, none of the previous process modelling studies evaluated the effects of changing operating parameters on gasification efficiency. From the literature review, it is understood that there are a large number of fluidized bed biomass gasifiers developed worldwide, unfortunately most of these projects are struggling to reach commercialization. Very few investigations have been done related to the prediction of the quality of the producer gas, incorporating the process parameters like temperature, equivalent ratio and steam to biomass ratio alone. Hence, the present work was aimed to develop a fluidized bed biomass gasifier using air as the gasifying agent and to investigate the effect of process parameters on the gasifier performance. A pilot scale fluidized bed saw dust gasifier had been developed for this purpose. The effect of equivalence ratio, gasifying temperature and pressure, along the particle size and feed rate on the fuel gas composition had been studied. An empirical relationship was developed to predict the product gas composition with the assumptions that the principal reactions were at thermodynamic equilibrium condition. The experimental data and the predicted vales have been analyzed, compared and discussed in the present work.

2. EXPERIMENTAL WORK

2.1. Feedstock and inert bed materials

The feed stock selected to study the fluidized bed gasification was saw dust with different granular sizes. These biomaterials were collected from rural industries of Cuddalore district, India. The proximate and ultimate analyses of saw dust used as feed stock are presented in Table 1.

Table 1
Ultimate and proximate analysis of Saw dust

Ultimate Analysis		Proximate Analysis	
Components	Percent	Component	Percent
Carbon	31.54	Volatile matter	54.91
Hydrogen	4.07	Fixed carbon	11.51
Sulphur	0.96	Moisture	11.43
Nitrogen	3.99	Ash	22.14
Oxygen	25.87		
HHV = 15.52 MJ/kg			

Considering the major elements and on the basis of dry and ash free condition, the saw dust was represented as $CH_{1.49}O_{0.64}$ on molar basis. The inert bed material used was sand and its particle size distribution was selected as 0.400 mm using sieve analysis. The properties of these materials and the procedures followed in finding out physical and chemical properties are mentioned in detail. Absolute specific gravity of the selected materials was measured using specific gravity bottle method. To minimize the complexities, resulting from the non-uniform particle size distribution in the bed, the average particle diameter was used to represent the particle size. Sieve analysis is commonly used to predict the particle size distribution of the feed stock having size of 70-500 μm. The test materials were dried and then sieved in a set of standard sieves and particle size distribution was observed [17]. Using oven method (110°C till reaching standard borne dry weight), moisture content of feed stock was measured (ASTM, E – 871). Proximate composition such as volatile matter (ASTM, E – 872) and ash (ASTM, E- 830) and fixed carbon (by weight difference) was found out by ASTM procedures. The elemental composition of the feed stock was found out using Elemental Analyzer (Carlo Erba EA 1108) coupled with auto sampler AS-200 and data processor DP 200-PRC. The minimum fluidization velocity was measured using pressure drop method. U tube manometers are used to measure the pressure drop below and above the distributor plate and at different heights of fluidized bed reactor. The air velocity corresponding to the peak pressure drop gives the experimental value of minimum fluidization velocity [17].

2.2. Experimental Set up

A pilot scale fluidized bed saw dust gasifier (capacity: 20 kg/h) had been developed and installed in the laboratory to carry out the experimental investigation. The schematic diagram of the setup is shown in Fig. 1. The cylindrical gasifier with 108 mm inside diameter up to a height of 1400 mm made of carbon steel material having inside refractory lining of thickness 0.1 m. The gasifier is fitted with a multiple hole distributor plate of 105 mm diameter was used for air distribution The ash discharge systems were provided for periodical disposal through the lock hopper arrangements. Silica sands were initially put into the gasifier through the screw feeder and air was introduced at the bottom of gasifier to maintain the bed in fluidized state. The air flow, after the discharge of blower, was controlled by a regulating valve and the flow was then estimated by an orifice meter placed in the supply pipe on the basis of pressure drops recorded across it. The orifice had been calibrated prior to the experiment with two reference instruments; namely a digital micromanometer (make: Furnace Control, England) and a thermal anemometer (make: Dantec, Denmark). The pressure drops across the orifice were recorded in the manometer and the corresponding flow rates were measured by the anemometer; the calibration curve was thus generated by plotting the flow rates along abscissa and the corresponding pressure drops along the ordinate.
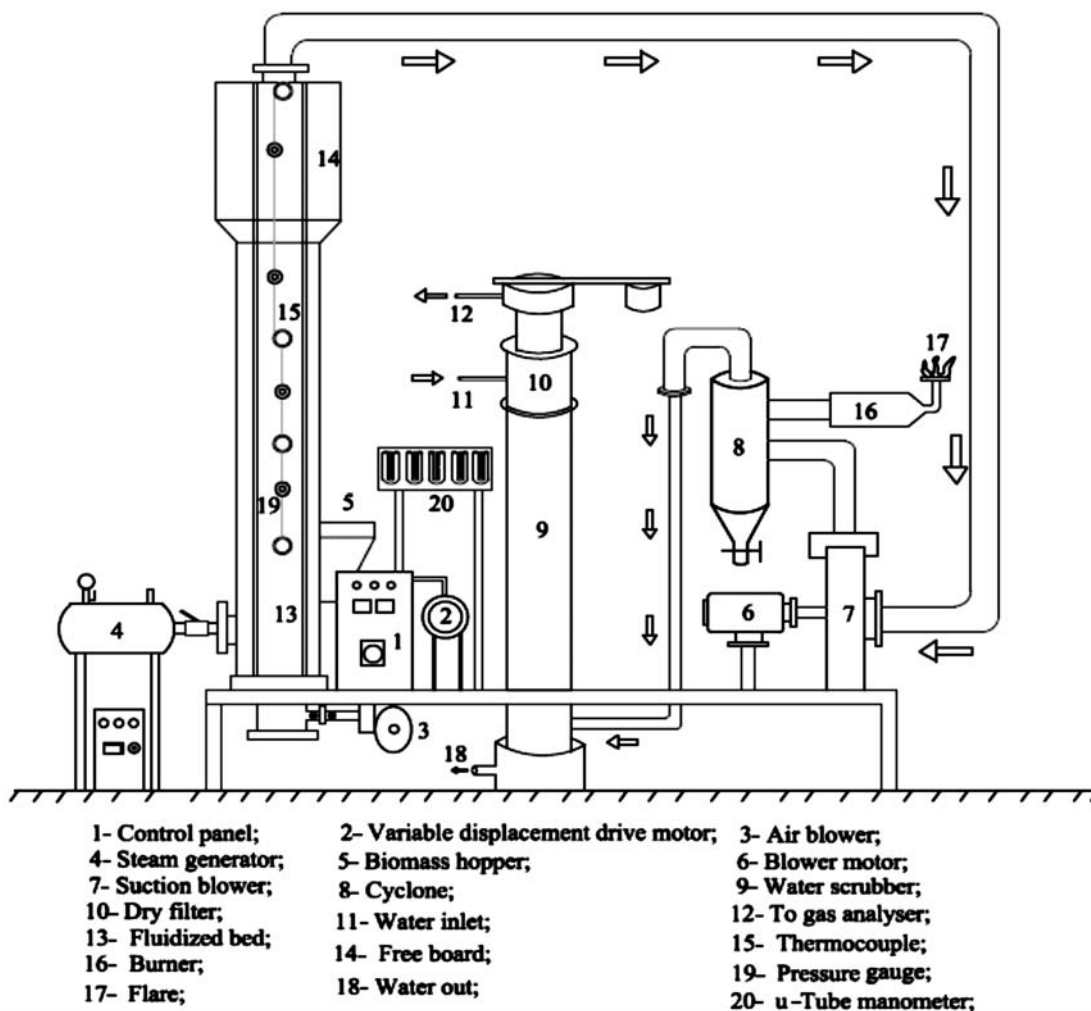
Figure 1: Experimental Set Up

1- Control panel;
4- Steam generator;
7- Suction blower;
10- Dry filter;
13- Fluidized bed;
16- Burner;
17- Flare;

2- Variable displacement drive motor;
5- Biomass hopper;
8- Cyclone;
11- Water inlet;
14- Free board;
18- Water out;

3- Air blower;
6- Blower motor;
9- Water scrubber;
12- To gas analyser;
15- Thermocouple;
19- Pressure gauge;
20- u–Tube manometer;

During experiment, the pressure drops were noted to get the corresponding air flow rates from the curve at different equivalence ratios. External electric heating was used for preheating the bed materials as well as the refractory lining during start up. The electric heating was switched onto and the gasifier was allowed to run until the bed temperature was 450°C. The raw saw dust was then fed through the under-bed feeding system having a screw feeder. The feed rate was controlled by the screw feeder fitted to a variable speed drive and it push the solid fuel immediately into the gasifier preventing pyrolysis outside the chamber. Supply of air was then regulated to maintain the desired equivalence ratio.The cyclone at the outlet of gasifier was used to separate the solid particles from the fuel gas mixture. The bag filter placed after the cyclone further cleaned up the gas by capturing dust and other smaller particles. The water cooler and an ice trap system were used in series to cool the fuel gas to separate the tar through condensation. A second orifice meter (50 mm diameter) was positioned in the fuel gas pipe (108 mm diameter) to estimate the gas yields. The calibration of the orifice was done prior to the experimental work by following the similar procedure as it was done in case of orifice meter in airline to generate a separate calibration curve. While the gasifier was running, the pressure drops across the orifice were noted in manometer to get the corresponding gas flow rates from the curve. The flow rates thus obtained corresponding to gas temperatures were then corrected by the temperature factor to get the actual flows at NTP. Equivalence ratio is very important in gasification process as it determines the fraction of the fuel that is burnt and thereby it controls the bed temperature. It also affects the fluidization of the bed. The lower limit of equivalence ratio is decided by the minimum quantity of air required to burn a portion of the fuel to release enough heat to support the endothermic reactions, to meet the sensible heat losses in gas, char and ash, and to maintain the required bed temperature of the reactor. As saw dust has high ash content, it requires larger fraction

of the fuel to be burnt – this ultimately demands a higher equivalence ratio [18]. In Hartiniati et al. [19], it is reported that the equivalence ratio was maintained between 0.30 and 0.48 during experimentation in a pilot scale fluidized bed gasifier fueled by saw dust. Later on, Mansaray et al. [20] also investigated the saw dust gasifier performance in a fluidized bed system by varying the equivalence ratio at 0.25, 0.30 and 0.35. In view of these observations, the gasifier was operated with equivalence ratios of 0.20-0.50 in the present investigation to get the experimental results.

2.3. Identifying the important factors and the feasible working limit

From the studies [21-30], the predominant factors that have a greater influence on the quality of the producer gas and the cold gas efficiency have been identified. They are: (*i*) bed temperature (T), (*ii*) pressure (P), (*iii*) the feed rate of the feed stock (F), (*iv*) the equivalent ratio (E) and (*v*) particle size (S). A number of trial experiments were carried out in the laboratory and the detailed review from the literature [21-30] and the following conclusions have been arrived at;

  (*i*) If the bed temperature was less than 650°C, the catalyst was required for higher production of hydrogen and nitrogen. The temperature at the bottom part of the gasifier is stable for all the fuels at around 650°C [21].

 (*ii*) If the bed temperature was greater than 950°C, expected that the gas composition will change with temperature inside the gasifier, but no clear trend was observed for the individual gas components during gasification [22].

(*iii*) If the pressure is less than 1 bar, the high purity of the produced gas is not required [23].

(*iv*) If the pressure is greater than 5 bar, the process control for chemical cycles due to the production of hydrogen in high pressure is to some extent difficult [24].

 (*v*) If the feed rate was less than 5 kg/hr, but as time passes, feed stock and ned materials gather on the bottom, forming a solid bed. [25]

(*vi*) If the feed rate was less than 20 kg/hr, may decrease the residence time of the material inside it and thus decrease its exposure to melting inside it. Hence, the gasifier used for the present work id designed with the maximum feed rate of 20 kg/hr [26].

(*vii*) If the equivalent ratio is less than 0.2, the change is temperature is very insensitive.

(*viii*) If the equivalent ratio is greater than 0.5, It is noted from the data that, at increased values of ER, the higher heating value of synthetic gas was reduced which is in good agreement with the results of the study conducted [22, 27].

(*ix*) If the particle size is less than 70 µm, implied higher conversions, and with lower solid temperatures into the bed and lower concentration of some gases, this means lower combustion richness [28].

 (*x*) If the particle greater than 500 µm, reduces the pre-treatment costs, but the devolatilization time increases, and thus for a defined throughput the gasifier size increases [29, 30].

2.4. Experimental Design Matrix

Owing to a wide range of factors, the use of five factors and central composite rotatable design matrix was chosen to minimize number of experiments. The assay conditions for the reaction parameters were taken at zero level (center point) and one level (-1) and (+1). The design was extended up to a ± α (axial point) of 2.378. The center values for variables were carried out at least 10 times for the estimation of error and single runs for each of the other combinations; thirty two runs were done in a totally random order. The design would consist of the 10 corner points of the 2^5 cube, the 16 star points, and 6 center points. The star points would have a = 32^ (1/4) = 2.378

For the convenience of recording and processing experimental data, the upper and lower levels of the factors were coded here as +2.378 and -2.378 respectively. The coded values of any intermediate value could be calculated using following relationship

$$X i \ = \ (2.378 \text{x} \{2X\text{-}[X\text{max - Xmin}]\})/(X\text{max - Xmin}) \ (1)$$

Where

Xi is the required coded value of a variable X; X is any value of the variable from Xmin to Xmax ;

Xmin is the lower level of the variable;

Xmax is the upper level of the variable.

Design matrix consisting of 32 sets of coded conditions (comprising a full replication five factorial of 16 points, 10 corner points and six centre points) was chosen in this investigation. Table 2 represents the ranges of factors considered, and Table 3 shows the 32 sets of coded and actual values with experimental results.

Table 2
Important factors and their levels

Factors	*Units*	*Factors levels*				
		-2.378	-1	0	+1	+2.378
Bed Temperature (T)	Celsius	650	725	800	875	950
Pressure (P)	MPa	1	2	3	4	5
Feed rate (F)	Kg/h	5	8.75	12.5	16.25	20
Particle size (S)	μm	70	142.5	215	357.5	500
Equivalence ratio (E)		0.2	0.275	0.35	0.425	0.5

2.5. Experimental Testing

During experimentation, special care was taken to maintain the desired bed temperatures as the selected feedstock was saw dust which had 17.09% ash, higher than any woody biomass and its ash had more than 95% silica. One of the important features of saw dust 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 during the conversion process [31]. The upper temperature is fixed by slagging phenomena which primarily depends upon the ash composition and the reaction atmosphere (like oxidation or reduction). Above this temperature, silica and potassium oxide in ash fuses on the surface of saw dust char particles forming a glass-like barrier that prevents the further reaction of the remaining carbon [32]. Some studies [33, 34] also indicate that oxidation of saw dust at a temperature higher than 900°C results in a physical structural transformation of silica from its original amorphous state to a crystalline state thereby encapsulating residual carbons. Once the structural changes of silica occurs, the combined carbon becomes unavailable for further oxidation reactions even at higher temperatures. In view of this, the gasifier was operated in the range of 600–950°C when the experiments were carried out with equivalence ratio 0.2 and 0.5. The gasification temperature was raised up to 700°C only in case of equivalence ratio of 0.25. The gasifier temperatures were recorded using Ni–Cr–Ni thermocouples with a digital display system. The gas sampling system was composed of probes fitted with septum. The sampling point was located at the outlet pipe of gasifier. The gas sampling probe made of glass was 50 mm in diameter and 500 mm in length. A syringe of volume capacity of 10 ml was used to collect the gas sample. The sample was analyzed in the Gas Chromatograph (Make – Chemito, model – GC1000) to get the raw experimental data and those were compared with the predicted values of the developed model. The energy content of the gas is assessed through the variable CGE (cold gas efficiency). This variable represents the ratio between the energy content of the permanent gas (HHV_{gas}) and the energy content of the initial biomass feedstock ($HHV_{saw\ dust}$) without taking into account the heat input in the reactor:

$$CGE = HHV_{gas}/HHV_{saw\ dust} \tag{1}$$

At the end of the experiment the residual tar were weighed and stored in a sealed recipient for further characterization. The tar yield is expressed as the ratio of the residual tar to the initial mass of saw dust

$$Y_{Tar}\% = [(M_{Tar}) / (M_{saw\ dust})] \times 100 \tag{2}$$

3. DEVELOPING THE EXPERIMENTAL DESIGN MATRIX

In the present investigation, to correlate the process parameters and the quality of the prodcer gas, a second order quadratic model was developed. In this study, the RSM provides a quantitative form of relationship between the desired response (Quality of the Producer gas) and the independent input variables, bed temperature (T), pressure (P), the feed rate of the feed stock (F), the equivalence ratio (E) and particle size (S), and can be expressed as a function, as in Equation (3)

$$\text{Producer gas (G)} = f(T, P, F, E, S) \tag{3}$$

Table 3
Experimental Results

Ex. No	Input parameters					Gas Composition							
	Bed Temperature(°C)	Pressure (bar)	Feed Rate (kg/hr)	Equivalent ratio	Particle size (μm)	Oxygen	Hydrogen	Carbon monoxide	Carbon-dioxide	Methane	Nitrogen	Tar yield	Cold gas efficiency
1.	725	2	8.75	0.275	392.5	0.21	5.89	12.83	10.89	1.25	54.6	4.6	64.11
2.	875	2	8.75	0.275	177.5	0.35	7.15	11.89	12.15	1.4	54.5	4.5	64.5
3.	725	4	8.75	0.275	177.5	0.25	6.16	12.24	11.25	1.31	54.65	4.65	64.65
4.	875	4	8.75	0.275	392.5	0.36	7.99	11.11	12.89	1.41	54.02	4.02	65.12
5.	725	2	16.25	0.275	177.5	0.28	6.38	12.908	11.52	1.32	53.99	3.99	63.99
6.	875	2	16.25	0.275	392.5	0.16	4.81	11.89	10.15	1.21	54.91	4.91	64.91
7.	725	4	16.25	0.275	392.5	0.24	5.11	12.282	12.94	1.28	54.86	4.86	64.86
8.	875	4	16.25	0.275	177.5	0.22	5.03	12.1	11.18	1.26	53.95	3.95	65.95
9.	725	2	8.75	0.425	177.5	0.33	8.06	12.27	11.97	1.37	54.44	4.44	64.44
10.	875	2	8.75	0.425	392.5	0.18	4.99	11.79	10.62	1.22	53.81	3.81	63.81
11.	725	4	8.75	0.425	392.5	0.32	7.68	12.14	13.88	1.36	54.15	4.15	64.15
12.	875	4	8.75	0.425	177.5	0.11	4.75	12.132	9.99	1.15	54.11	4.11	64.11
13.	725	2	16.25	0.425	392.5	0.15	5.77	12.216	10.35	1.19	54.53	4.53	64.53
14.	875	2	16.25	0.425	177.5	0.185	5.34	12.16	10.67	1.23	54.25	4.25	64.25
15.	725	4	16.25	0.425	177.5	0.25	6.6	12.2	11.65	1.29	54.45	4.45	64.45
16.	875	4	16.25	0.425	392.5	0.24	5.45	12.266	11.16	1.28	54.78	4.78	64.78
17.	650	3	12.5	0.35	285	0.25	5.61	12.42	11.45	1.29	54.51	4.51	62.51
18.	950	3	12.5	0.35	285	0.32	7.21	11.05	13.36	1.21	53.23	3.23	65.13
19.	800	1	12.5	0.35	285	0.188	3.77	12.148	10.69	1.28	54.19	4.19	64.19
20.	800	5	12.5	0.35	285	0.36	6.13	11.922	12.24	1.3	53.06	3.06	64.26
21.	800	3	5	0.35	285	0.24	6.45	12.018	11.16	1.28	53.54	3.54	65.54
22.	800	3	20	0.35	285	0.18	4.11	12.178	10.62	1.24	54.34	4.34	64.04
23.	800	3	12.5	0.2	285	0.16	4.6	12.346	10.44	1.25	54.68	4.68	64.68
24.	800	3	12.5	0.5	285	0.2	6.44	12.216	10.8	1.22	54.53	4.53	64.93
25.	800	3	12.5	0.35	70	0.32	6.64	12.046	12.1	1.36	53.68	3.68	63.68
26.	800	3	12.5	0.35	500	0.18	4.91	11.896	10.65	1.22	52.93	2.93	63.93
27.	800	3	12.5	0.35	285	0.22	5.16	11.988	10.5	1.26	53.39	3.39	63.39
28.	800	3	12.5	0.35	285	0.13	3.33	11.788	9.25	1.15	52.39	2.69	62.69
29.	800	3	12.5	0.35	285	0.13	3.33	11.788	9.25	1.26	52.39	2.39	62.39
30.	800	3	12.5	0.35	285	0.13	3.33	11.788	9.25	1.19	52.39	2.39	62.39
31.	800	3	12.5	0.35	285	0.13	3.33	11.788	9.89	1.19	52.39	2.39	62.39
32.	800	3	12.5	0.35	285	0.13	3.33	11.788	9.25	1.19	52.39	2.39	62.39

Table 4

ANOVA Test Results

Source	Oxygen F value	Oxygen p-value	Hydrogen F value	Hydrogen p-value Prob>F	Carbon-monoxide F value	Carbon-monoxide p-value Prob>F	Carbon-dioxide F value	Carbon-dioxide p-value Prob>F	Methane F value	Methane p-value Prob>F	Nitrogen F value	Nitrogen p-value Prob>F	Tar Yield F value	Tar Yield p-value Prob>F	Cold gas effciency F value	Cold gas effciency p-value Prob>F
Model	2.738	0.045	2.983	0.033	13.687	0.000	2.706	0.046	4.148	0.010	4.836	0.005	4.641	0.006	2.795	0.042
T	0.111	0.745	0.419	0.531	116.745	0.000	0.204	0.660	3.826	0.076	3.258	0.098	3.266	0.098	5.528	0.038
P	3.684	0.081	1.262	0.285	10.395	0.008	5.827	0.034	1.009	0.337	1.153	0.306	1.156	0.305	1.327	0.274
F	3.929	0.073	8.022	0.016	10.438	0.008	1.604	0.231	6.711	0.025	1.980	0.187	1.984	0.187	0.003	0.958
E	0.780	0.396	0.700	0.420	0.313	0.587	0.237	0.636	4.698	0.053	0.340	0.572	0.341	0.571	0.929	0.356
S	2.404	0.149	1.332	0.273	7.790	0.018	0.010	0.923	4.698	0.053	0.007	0.935	0.007	0.935	0.018	0.895
TP	0.028	0.869	0.159	0.697	6.397	0.028	1.044	0.329	0.205	0.659	0.432	0.524	0.433	0.524	0.311	0.588
TF	0.001	0.981	0.007	0.933	7.785	0.018	0.085	0.776	0.004	0.950	0.685	0.425	0.687	0.425	0.517	0.487
TE	4.576	0.056	5.920	0.033	32.429	0.000	2.482	0.143	8.490	0.014	0.003	0.956	0.003	0.956	1.800	0.207
TS	1.623	0.229	1.007	0.337	4.817	0.051	0.065	0.803	3.526	0.087	0.003	0.956	0.003	0.956	0.014	0.907
PF	0.971	0.346	0.026	0.874	2.880	0.118	0.320	0.583	1.212	0.295	0.195	0.667	0.196	0.667	0.209	0.656
PE	0.001	0.981	0.005	0.945	18.137	0.001	0.021	0.887	0.004	0.950	0.309	0.590	0.309	0.589	1.007	0.337
PS	13.881	0.003	6.105	0.031	0.576	0.464	11.399	0.006	24.857	0.000	0.000	0.991	0.000	0.991	0.027	0.872
FE	0.555	0.472	0.912	0.360	1.498	0.247	0.142	0.713	1.513	0.244	0.782	0.395	0.784	0.395	0.004	0.949
FS	0.708	0.418	0.507	0.491	0.011	0.917	1.032	0.331	0.943	0.352	4.072	0.069	4.082	0.068	0.133	0.722
ES	0.486	0.500	0.000	0.987	1.901	0.195	0.085	0.776	0.943	0.352	0.526	0.483	0.528	0.483	0.002	0.967
T^2	10.203	0.009	16.415	0.002	0.113	0.743	18.205	0.001	2.592	0.136	19.619	0.001	18.565	0.001	5.139	0.045
P^2	8.455	0.014	3.681	0.081	8.866	0.013	7.388	0.020	9.076	0.012	13.521	0.004	12.643	0.005	9.673	0.010
F^2	1.541	0.240	5.763	0.035	13.495	0.004	3.136	0.104	3.844	0.076	21.569	0.001	20.465	0.001	18.376	0.001
E^2	0.212	0.654	7.569	0.019	32.434	0.000	1.759	0.212	1.175	0.302	44.701	0.000	43.136	0.000	18.644	0.001
S^2	5.212	0.043	9.758	0.010	5.156	0.044	6.604	0.026	9.076	0.012	7.261	0.021	6.616	0.026	4.998	0.047
Lack of Fit	0.136		0.234		0.106		0.026		0.743		0.394		0.374		0.076	

The empirical relationship must include the main and interaction effects of all factors and hence the selected polynomial is expressed as follows:

$$Y = bo + \sum bi\ xi + \sum bii\ xi^2 + \sum bij\ xi\ xj \qquad (4)$$

For five factors, the selected polynomial could be expressed as;

$$\text{Quality of the Producer Gas (G)} = \{b_0 + b_1 (T) + b_2 (P) + b_3 (F) + b_4 (E) + b_5(S) + b_{11}(T^2)$$
$$+ b_{22}(P^2) + b_{33}(F^2) + b_{44}(E^2) + b_{55}(S^2) + b_{12} (TP) + b_{13} (TF)$$
$$+ b_{14}(TE)\ b_{15}(TS) + b_{23}(PF) + b_{24}(PE) + b_{25}(PS) + b_{34}(FE)$$
$$+ b_{35}(FS) + b_{45}(ES)\} \quad (5)$$

where b_0 is the average of response and $b_1, b_2, b_3 \ldots b_{11}, b_{12}, b_{13} \ldots b_{22}, b_{23}, b_{33}$, are the coefficients that depend on their respective main and interaction factors, which are calculated using the expression given below,

$$Bi = (\sum (Xi, Yi))/n \qquad (6)$$

Where 'i' varies from 1 to n, in which Xi the corresponding coded value of a factor and Yi is is the corresponding response output value (Producer Gas) obtained from the experiment and 'n' is the

total number of combination considered. All the coefficients were obtained applying central composite rotatable design matrix including the Design Expert statistical software package. After determining the significant coefficients (at 95% confidence level), the final relationship was developed including only these coefficients.

The Analysis of Variance (ANOVA) technique was used to find the significant main and interaction factors. The results of second order response surface model fitting as Analysis of Variance (ANOVA) are given in the Table 4. The determination coefficient (r^2) indicated the goodness of fit for the model. The Model F-value of (Oxygen = 3.84, Hydrogen = 5.85, Nitrogen = 5.23, Carbon-monoxide = 4.41, Carbon-di-oxide = 5.33, Methane = 4.15, Cold Gas Efficiency = 4.97 and Tar Yield = 8.11) implies the model is significant. There is only a 0.01% chance that a "Model F-Value" this large could occur due to noise.

4. RESULTS AND DISCUSSION

4.1. Effect of Temperature

O_2, N_2, H_2 and CO concentrations were found to increase with increase in bed temperature, and decreased the concentrations of CH_4 and CO_2. This may be explained with Le Chatelier's principle which states that higher temperature favors the reactants in exothermic reactions and the products in endothermic reaction. Higher temperatures favoured the formation of O_2, N_2, H_2 and CO coupled with increased reforming of methane. However, if the temperature is further increased, H_2 is converted to CO and H_2O by the reverse water gas shift reaction, which is favoured at high temperatures. In this case, the maximum H_2/CO ratio occurred at between 750-950°C, but this varied according to the other operating variables. According to previous studies [35, 36], the carbon boundary temperature for grass, which has a similar lingo cellulosic composition and heating value as bagasse, occurs at around 850°C, which would correspond to the optimum gasifier temperature. This correlates with the results presented here, since at 850°C all the carbon was converted, and the highest system efficiency was observed for all cases, due to the increase in external gasifier heat requirements at elevated temperatures. However, this is the theoretical case and is only applicable when the residence time is long enough for equilibrium to be reached. Based on the bagasse gasifier tested by DeFilippis et al. [17], this minimum temperature was assumed 650°C for the purposes of this study. Therefore, the endothermic reaction was strengthened with increasing temperature, which resulted in more H_2 and less CH_4 concentrations. It was also found that the experimentally measured CH_4 concentrations were more than the calculate values. At 950°C and 0.35 ER, the measured concentration of CH_4 was experimentally found to be 1.34 % whereas the calculated value was 1.46 % – the possible explanation could be that the equilibrium state might not have reached in the bed. Nevertheless the gasification process, involving reactions of homogeneous and heterogeneous in nature along with initial drying and pyrolysis, is very complicated to determine the actual kinetics of the chemical reactions. Some of these facts might have caused to vary the predicted values from the experimental data. At 950°C, at equivalence ratio of 0.35, the cold gas efficiency is higher.

4.2. Effect of Pressure

An increase in gasifier pressure leads to reduced partial pressures of CO and CH_4 coupled with an increase in CO_2, H_2 and O as reflected in Table 4. This trend is consistent with LeChatelier 's principle, and has been reported in the literature for other feedstock and gasifier types [37, 38]. In practice, high pressure gasification may have economic advantages in upstream processing due to smaller equipment sizes. Higher overall efficiencies could also be achieved if hot gas cleaning is used, but this is still in development. Currently wet gas cleaning is the only available option, in which case the energy losses associated with compression and decompression are high if coupled with a high pressure gasifier. It is estimated that higher O_2, N_2, H_2 and CO concentrations are increase with the increasing pressure of air flow. However, CH_4 and CO_2 decreased with the increase in pressure. The cold gas efficiency was estimated during gasification period, to 64 %. Generally, increasing the pressure will increase the cold gas efficiency since the heating

value of the produced gas will increase with pressure [23] as a result of CH_4 production through the steam reforming reaction. On the other hand, increasing temperature will decrease the heating value of produced gas and hence lower the cold gas efficiency [24]. A produced gas with a high CH_4 content could be used for example power production. However, if the produced gas is intended to be used for catalytic conversion to methanol, DME, Fischer Tropsch products etc, then the yield of H_2 and CO and the O_2 ratio should instead be considered [17] Improvements in determining the syngas flow through e.g. flow measurement or by trace experiments will be evaluated in future work as well as optimization of the operation. Similarly, cold gas efficiencies varied from 20 to 87% in an entrained flow biomass gasifier and depended on addition of steam and air preheating [39]. The composition of the syngas at equilibrium was determined at 5 bar, 0.36 % O_2 and 53% N_2. The calculations are performed by determining the minimum of Gibbs free energy of a specific system based on a database containing thermodynamic data for various chemical species and phases. Compared to thermodynamic equilibrium the syngas contains less CO but more CO2. The syngas also contains 1.3% CH_4 which is not predicted at all at equilibrium.

A possible explanation could be that the heterogeneous reactions involved in char gasification are too slow to be completed within the residence time of the reactor at the current gasification conditions. This will thereby result in less CO_2 and more CO, which could explain the difference between measured CO_2 and CO concentrations compared to equilibrium values. The syngas can also have become shifted in the quench, which could also explain differences between the measured syngas composition after the quench compared to the syngas composition at equilibrium.

4.3. Effect of Feed rate

Major constituents of synthetic gas such as carbon monoxide, carbon dioxide, methane, hydrogen, nitrogen and oxygen were observed for each gasification trial and are presented in Table 4. The data revealed a reduction in carbon monoxide content with increase of feed rate, whereas CO_2 was decreased with increase of feed rate. The value of carbon monoxide was in the range of 12%. Data on carbon dioxide content indicated that, with the increase of feed rate from 5 kg/hr to 20kg/hr, the CO_2 content was decreasing. The maximum 12.49% and minimum (12.14%) value of CO_2 was observed at 0.35 equivalent ratio. The results are in good agreement with the findings of previous research [30, 39]. The study on fluidized bed gasification of saw dust reported a decreasing trend of CO_2 and increasing trend of CO with increasing feed rate. The availability of higher quantity of oxygen led to reduction in CO level [30]. With the increased supply of stoichiometric air supply, the quantity of hydrogen was reduced in synthetic gas during gasification of all the three biomaterials. The content of hydrogen in the product gas was in the range of 4–8% (saw dust). The results of earlier researches [40, 41] showed the same trend of decrease of CO and reduction of H_2 with increase of feed rate. The reduced level of methane content in the product gas of saw dust was noted during increased feed rate. The content of methane was increasing during the progress of gasification and hence maximum value of methane content was observed at the later stages of gasification. The overall range of methane content was 1–1.3% in all the trials of fluidized bed gasification. With the increase of rate of feed stock, the percentage of oxygen and hydrogen was decreased. The same pattern of change of synthetic gas constituents was observed in the earlier study conducted [42].

4.4. Effect of Equivalence ratio

Equivalence ratio (ER) is defined as the ratio of the actual air– fuel ratio to the stoichiometric air–fuel ratio. In autothermal gasification process, a part of fuel is burnt to release energy to sustain the endothermic gasification reactions. During the experimentation, ER was varied at 0.2-0.5 changing the air flow rates with the varying parameters. Gasification with oxygen is exothermic; therefore the desired increase in temperature above the carbon boundary temperature can be achieved by feeding more oxygen to the gasifier. Increasing the equivalence ratio leads to over-oxidization and partial combustion of the syngas to produce H_2O and CO_2. As this will decrease the gasification efficiency, the equivalence ratio should be kept

to a minimum within the practical constraints. It is found the theoretical optimum conditions for maximum efficiency and hydrogen production from atmospheric gasification of dry biomass to be 825°C and an equivalence ratio of 0.35.However, they did not account for practical considerations such as tar formation. It has been reported in literature that a 20% secondary air injection above the gasifier freeboard can reduce tar formation [43]. The results from the study [38] showed a good correlation between experimental and predicted results for bagasse gasification with no tar formation at equivalence ratios of 0.35. It was seen that higher ER values decreased the concentrations of hydrogen and carbon-di-oxide and degraded the gas quality with more N_2 dilution and higher CO_2 concentration due to oxidization of larger fraction of carbon in feedstock; as a result, the heating values of fuel gas decreased. High degree of combustion occurs at high equivalent ratio which supplies more air into the gasifier and improves char burning to produce CO_2 instead of combustible gases such as CO, H_2 and CH_4. In biomass gasification, the ER varies from 0.10 to 0.50 [44]. It was observed that increasing the ER reduced the amount of CO and CH4. Similar trends were obtained from the research [45, 46]. From the present research, the effect of ER variation (0.2-0.4) is one of the most important operation parameters on the quality of the producer gas. H_2 production peaked at ER of 0.35. Lower heating value of the producer gas was obtained at high ER which was due to the promotion of the oxidation reaction and dilution of the producer gas with N_2. ER not only represents the oxygen quantity introduced into the reactor but also affects the gasification temperature under the condition of auto thermal operation. Higher ER will cause gas quality to degrade because of more oxidization reactions. On the other side, higher ER means higher gasification temperature, which can accelerate the gasification and improve the product quality to a certain extent. Therefore the gas composition is affected by the two contradictory factors of ER. Through the analysis on both the experimental data and model results of varying ER, it can be understood that it is unfeasible to apply too small or too large ER in biomass air-steam gasification. Too small ER will lower reaction temperature, which is not favorable for biomass steam gasification. Too large ER will consume more H_2 and other combustible gases through oxidization reaction. So there exists an optimal value for ER, which is different according to different operating parameters.

4.5. Effect of Particle size

It has been accepted that small particle size biomass significantly increases the overall energy efficiency of the gasification process, but it also increases the gasification plant cost. On the other hand, an increase in biomass particle size reduces the pre-treatment costs, but the devolatilization time increases, and thus for a defined throughput the gasifier size increases. Therefore, a balance should be considered while investigating the effect of biomass particle size on the gasification efficiency [47]. The non-uniformity of the biomass particles will influence gasification reaction rate. However, due to intense mixing caused by the fluidized sand, temperature longitudinally does not vary much and are almost similar, indicating that the irregular shapes and size of wood chips do not affect the temperature [48]. On the other hand, the research made earlier [49] observed that the producer gas yield, LHV and carbon conversion were improved as the biomass particle size decreased. It was explained that small biomass particles contribute to large surface area and high heating rate which in turn produce more light gases and less char and condensate. Therefore, the yield and composition of the producer gas improved while using the small particle biomass. A possible explanation is that for small particle sizes the pyrolysis process is mainly controlled by reaction kinetics; as the particle size increases, the product gas resultant inside the particle is more difficult to diffuse out and the process is mainly controlled by gas diffusion.

5. CONCLUSIONS

The present study was focused on the gasification of saw dust in a pilot scale fluidized bed reactor installed in the laboratory. The gasifier was operated at bed temperatures ranging from 650 °C to 950 °C with varying equivalence ratios of 0.2 – 0.5 , pressure 1 to 5 bar, feed rate 5 -20 kg/hr and particle size 70 -500 μm to investigate the fuel gas compositions.

The empirical relation was developed in order to quantify the composition of fuel gas. This model gave results with high accuracy showing similar trends in predicting the variation of gas species concentrations in line with experimental data.

It was noticed that the amount of CH_4 produced during the gasification process was more in comparison to the predicted values. The possible reason could be that the equilibrium state might not have reached for not having enough bed temperature in gasifier.

It was seen that hydrogen, oxygen, nitrogen and carbon monoxide contents in fuel gas were increased with rise in bed temperatures, equivalent ratios.

The cold gas efficiency was found to increase at higher temperature, equivalence ratio and pressure due to presence of more CO_2 and O_2 in the fuel gas, even though the rate of carbon conversion was more with the rise in bed temperature.

6. REFERENCES

1. A.V.Bridgwater, The technical and economic feasibility of biomass gasification for power generation. Fuel. 74(3) (1995) 631–53.

2. C. Di Blasi, Kinetic and heat transfer control in the slow and flash pyrolysis of solids, Industrial and Engineering Chemistry Research 35 (1) (1996) 37–46.

3. A.Ergudenler, Gasification of wheat straw in a dual distributor type fluidized bed reactor. Unpublished PhD. thesis, Technical University of Nova Scotia, Halifax, Nova Scotia, Canada;(1993).

4. Mitchell D (2008) A note on rising food prices. Policy Research Working Paper 4682, The World Bank Development

5. Farrell AE, Plevin RJ, Turner BT, Jones AD, O'Hare M, Kanman DM (2006) Ethanol can contribute to energy and environmental goals. Science 311:506–509.

6. Rammamorth R, Kastury S, Smith WH (2000) Bioenergy: vision for the new millennium. Science Publishers, Enfield

7. Huber GW, Iborra S, Corma A (2006) Synthesis of transportation fuels from biomass: chemistry, catalysts, and engineering. Chem Rev 106:4044–4098.

8. Foust TD, Wallace R, Wooley R, Sheehan J, Ibsen K, Dayton D, Himmel M, Ashworth J, McCormick R, Hess JR, Wright C, Radtke C, Perlack R, Mielenz J, Wang M, Synder S, Werpy T (2007) A national laboratory market and technology assessment of the scenario. Technical Report, NREL/TP-510-4094.

9. Czernik S, Bridgwater AV (2004) Overview of applications of biomass fast pyrolysis oil. Energy Fuels 18(2):590–598

10. M.M.Bergqvist , K.S. Wardh, A.Das, E.O.Ahlgren, A techno-economic assessment of rice husk-based power generation in the Mekong River Delta of Vietnam. Int J Energy Res. 32(12) (2008)1136e50.

11. C.R. Altafini, P.R. Wander, R.M.Barreto. Prediction of the working parameters of a wood waste gasifier through an equilibrium model. Energy Convers Manage. 44(2003)2763–77.

12. M.Baratieri, P.Baggio, L.Fiori, M.Grigiante. Biomass as an energy source: thermodynamic constraints on the performance of the conversion process. Bioresour Technol. 99(2008)7063–73.

13. Z.Zainal , R. Ali , C.Lean , K.Seetharamu,. Prediction of performance of a downdraft gasifier using equilibrium modeling for different biomass materials. Energy Convers Manage. 42(2008)1499–515.

14. K.J.Ptasinski, M.J.Prins, A.Pierik, Exergetic evaluation of biomass gasification. Energy. 32(2007)568–74.

15. M.J.Prins, K.J.Ptasinski , F,Janssen. From coal to biomass gasification: comparison of thermodynamic efficiency. Energy.32 (2007)1248–59.

16. M.R.Mahishi, D.Goswami, Thermodynamic optimization of biomass gasifier for hydrogen production. Int J Hydrogen Energy. 32(2007)3831–40.

17. P.De Filippis, C.Borgianni ,M.Paolucci , F.Pochetti , Gasification process of Cuban bagasse in a two-stage reactor. Biomass Bioenergy.27(2004)247–52.

18. Z.A. Bin ZainalAlauddin, L. Pooya, M. Mohammadi, A.R. Mohamed, Gasification of lignocellulosic biomass in fluidized beds for renewable energy development. ar view, Renewable and Sustainable Energy Reviews. 14(2010) 2852–2862.

19. K.G.Mansaray, A.E.Ghaly, A.M.Al-Taweel, F. Hamdullahpur, Ugursal, Air gasification of rice husk in a dual distributor type fluidised bed gasifier. Biomass Bioenergy. 17(1999)315–332. (Temperare)

20. Simin Shabani, Mojtaba Aghajani Delavar, Mohammadreza Azmi. Investigation of biomass gasification hydrogen and electricity co-production with carbon dioxide capture and storage, international jornal of hydrogen energy 38, pp. 3630-3639, 2013 (Pressre)

21. Koc R, Kazantzis NK, Hua Ma Y. A process dynamic modeling and control framework for performance assessment of Pd/ alloy-based membrane reactors used in hydrogenproduction. Int J Hydrogen Energy 36:4934e51.(2011)

22. P.Mathieu, R.Dubuisson, Performance analysis of a biomass gasifier. Energy Convers. Manage..Eqivalent ratio & Feed rate. 43(2002)1291–1299

23. C.L.Lin, M.L.Wey, S.D.You, The effect of particle size distribution onminimum fluidization velocity at high temperature. Powder Technol. 126 (2002) 297–301. (Particle Size & Feed rate)

24. A.V.Drift, J.Doorn, J.W.Vermeulen, Ten residual biomass fuels forcirculating fluidized-bed gasification. Biomass Bioenergy. 20(2009) 45–56.

25. S.Luoa, B.Xiao, Z. Hua, S. Liua,Y. Guana, L. Caia, Influence of particle size on pyrolysis and gasification performance of municipal solid waste in a fixedbed reactor. Bioresour Technol. 101(16)(2010) 6517–6520. (Particle Size)

26. Susana Martı´nez-Lera, Jose´ TorricoJavier Pallare´s Antonia GilDesign and first experimental results of a bubbling fluidized bedfor air gasification of plastic waste, J Mater Cycles Waste Manag.15 (2013)370–380.

27. M. Risberg, O.G.W. Ohrman , B.R. Gebart , P.T. Nilsson, A. Gudmundsson , M. Sanati, Influence from fuel type on the performance of an air-blown cyclonegasifier, Fuel. 116 (2014) 751–759.

28. K.G.Mansaray, A.E.Ghaly, Al-Taweel AM, Hamdullahpur F, Ugursal VI. Airgasification of rice husk in a dual distributor type fluidized bed gasifier.Biomass Bioenergy.17(1999)315–32.

29. Kaupp Albrecht. Gasification of rice hulls: theory and practices. Deutsches Zentrum Fuer Entwicklungs Technologien (GATE): Eschborn; (1984).

30. G.F. Schiefelbein , Biomass thermal gasification research, recent results – UnitedStates DOE's research program. Biomass.19(1989)145–59.

31. Z.A. Bin Zainal Alauddin, L. Pooya, M. Mohammadi, A.R. Mohamed, Gasification of lignocellulosic biomass in fluidized beds for renewable energy development: a review, Renewable and Sustainable Energy Reviews. 14 (2010) 2852–2862.

32. J.Hanb, H.Kimb, H.Minamib, W. Shimizuc, T.G Wang, The effect of the particle size of alumina sand on the combustion and emission behavior of cedar pellets in a fluidized bed combustor. Bioresour. Technol. 99(9)(2008) 3782–3786.

33. Y. Zhao, S. Sun, H. Zhou, R. Sun, H. Tian, J. Luan, J. Qian, Experimental study onsawdust air gasification in an entrained-flow reactor, Fuel Processing Technology.91 (2010) 910–914.

34. P.K. Senapati, S. Behera, Experimental investigation on an entrained flow typebiomass gasification system using coconut coir dust as powdery biomass feedstock,Bioresource Technology. 117(2012) 99–106.

35. L.Wang, C.L.Weller, D.D.Jones, M.A.Hanna. Contemporary issues inthermal gasification of biomass and its application to electricity and fuel production.Biomass Bioenergy. 32(2008)573–81.

36. P.Mathieu, R.Dubuisson, . Performance analysis of a biomass gasifier. EnergyConvers. Manage. 43(2002) 1291–1299.

37. A.K.Sharma. Equilibrium modeling of global reduction reactions for a downdraft (biomass) gasifier. Energy Convers Manage. 49(2008)832–42.

38. M.Vaezi ,M.Passandideh-Fard, M.Moghiman, M.Charmchi. Gasification of heavy fuel oils: a thermochemical equilibrium approach. Fuel. 90(2011) 878–85.

39. S.Rapagna, N.Jana, A.Kiennemann, P.U.Foscolo, Steam-gasification of biomass in a fluidized-bed of olivine particles. Biomass Bioenergy. 19(2000)187–197.

40. C.Franco, F.Pinto, I.Gulyurtlu, I.Cabrita, . The study of reactions influencing the biomass steam gasification process. Fuel. 82(2003) 835–842.

41. M.J.C.Van der Stelt, H.Gerhauser, J.H.A.Kiel, K.J.Ptasinski, Biomass upgrading by torrefaction for the production of biofuels: a review. Biomass Bioenergy. 35(2011)3748–3762.

42. J.I.Hayashi, S.Hosokai, N.Sonoyama, Gasification of low-rank solid fuels with thermochemical energy recuperation for hydrogen production and power generation. Process Saf. Environ. 84(2006) 409–419.

43. S. Hu, J. Xiang, L. Sun, M. Xu, J. Qiu, P. Fu, Characterization of char from rapid pyrolysis of rice husk. Fuel Process Technol. 89(2008)1096–1105.

44. D. Lv, M. Xu, X. Liu, Z. Zhan, Z. Li, H. Yao, Effect of cellulose, lignin, alkali and alkaline earth metallic species on biomass pyrolysis and gasification, Fuel Processing Technology. 91 (2010) 903–909.

45. A. Sluiter, B. Hames, R. Ruiz, C. Scarlata, J. Sluiter, D. Templeton, D. Crocker, Determination of structural carbohydrates and lignin in biomass, Laboratory Analytical Procedure (LAP), National Renewable Energy Laboratory Technical Report, NREL/TP-510-42618 NREL Golden, NREL, CO, USA, 2008.

46. B. Ramajo-Escalera, A. Espina, J.R. García, J.H. Sosa-Arnao, S.A. Nebra, Model-free kinetics applied to sugarcane bagasse combustion, Thermochimica Acta. 448 (2006) 111–116.

47. M.V. Gil, D. Casal, C. Pevida, J.J. Pis, F. Ribiera, Thermal behaviour and kinetics of coal/biomass blends during co-combustion, Bioresource Technology. 101 (2010) 5601–5608.

48. A. Demirbas, Combustion characteristics of different biomass fuels, Progress in Energy and Combustion Science. 30 (2004) 219–230.

49. H. Haykiri-Açma, Combustion characteristics of different biomass materials, Energy Conversion and Management. 44 (2003) 155–162.