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EFFECT OF SECONDARY AIR INJECTION ON THE COMBUSTION EFFICIENCY OF SAWDUST IN A FLUIDIZED BED COMBUSTOR

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Abstract - Agricultural wastes like bagasse, paddy husks, sawdust and groundnut shells can be effectively used as fuels for fluidized bed combustion; otherwise these biomass fuels are difficult to handle due to high moisture and fines content. In the present work the possibility of using sawdust in the fluidized bed combustor, related combustion efficiencies and problems encountered in the combustion process are discussed. The temperature profiles for sawdust with an increase in fluidizing velocity along the vertical height above the distributor plate indicate that considerable burning of fuel particles is taking place in the freeboard zone rather than complete burning within the bed. Therefore, an enlarged disengagement section is provided to improve the combustion of fines. The temperature profiles along the bed height are observed at different feed rates. The feed rate of sawdust corresponding to the maximum possible temperature was observed to be 10.2 kg/h. It is observed that 50-60% excess air is optimal for reducing carbon loss during the burning of sawdust. The maximum possible combustion efficiency with sawdust is 99.2% and is observed with 65% excess air.
Keywords: Sawdust fuel; Fluidized bed; Waste utilization; Low CO emission; High combustion efficiency; Freeboard burning.

INTRODUCTION

Fluidized bed energy technology offers several unique characteristics for using biomass in small-scale energy conversion operations. A fluidized bed consists of a chamber in which solid particles are kept in a state of suspension by high velocity air forced upward through the particles. The turbulent mass of solid particles stores heat and transfers this heat rapidly to any fuel that is introduced into the bed. This results in an efficient energy conversion method, which can use a wide range of low-grade fuels, such as biomass.

The potential of agricultural residues for energy production has been investigated by many researchers (Jenkins and Bhatnagar, 1991; Kjellstrom, 1993; Babu et al., 1995). Peel and Santos (1980) and Peel (1989) investigated the combustion of sawdust, bagasse, rice husk, wood chips and corn cobs in a 200 mm diameter fluidized bed combustion test rig. It has been suggested that satisfactory combustion of the bagasse, sawdust and the rice husks could be achieved with under-bed feeding only. Bhattacharya et al. (1983) reported the combustion efficiency of rice husk in the range of 81

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to 98%. It has also been suggested that higher combustion efficiencies could be achieved by providing an enlarged freeboard, which effectively reduces the flue gas velocity. Hellwig (1985) analyzed the heat distribution during the combustion of wood chips and straw and showed that over 67% of their calorific values were released through the combustion of the volatiles. Preto et al. (1987) conducted combustion studies with rice husk as fuel in a pilot-scale fluidized bed combustor and achieved a combustion efficiency of about 95% with low emissions of NO_x and SO_x. The fluidizing velocities were varied from 0.4 to 2.2 m/s with excess air levels of 30 to 95%. The bed temperatures were maintained between 650 and $900^o C$. These results confirm that at all fluidizing velocities a significant amount of combustion is taking place in the freeboard. Bhattacharya and Weizhang (1990) reported that the loss of unburnt carbon in the form of carbon monoxide (CO) is in the order of 3 to 10%. The higher CO emissions were observed at higher fluidization velocities and this could be because of shorter residence time. Salour et al. (1993) found that the combustion efficiency in bubbling beds is dependent on the particle size of the fuel, excess air level, bed temperature and gas velocity. It is recommended that the height of freeboard must be increased to increase the combustion efficiencies. Kaferstein et al. (1997) studied the process of biomass combustion during batch experiments in a bubbling fluidized bed and observed that during the combustion of the biomass a rapid consumption of oxygen was taking place in two distinct phases: a short phase for volatile combustion and a long char combustion phase. Kouprianov and Permchart (2003) studied the effects of operating conditions (load and excess air) as well as fuel quality and bed height on the major gaseous emissions (CO_2, CO and NO_x) in a conical fluidized bed combustor while burning mixed sawdust generated from different woods in Thailand. It was found that bed height had very little influence on the emission profiles. The CO_2 emission profiles along combustor height were found to be almost independent of the combustor load and fuel quality. The maximum CO levels of 1700 to 2300 ppm were observed for low moisture sawdust and 3000 to 6000 ppm were observed for sawdust with a high moisture content. The critical excess air was in the order of 40 to 45% when firing sawdust with a low moisture content and 65 to 70% when sawdust with a high moisture content was fired. The NO_x emissions from the conical FBC were found to be in the range of 110 to 120 ppm for sawdust with low moisture and of 80 to 90 ppm for high moisture sawdust. Permchart and Kouprianov (2004) conducted experimental studies in a conical fluidized bed combustor with different biomass fuels: rice husk, sawdust and bagasse. It has been revealed that for the maximum combustor load and excess air of 50 to 100%, a combustion efficiency of over 99% could be achieved when firing sawdust and bagasse. Srinivasa Rao and Venkat Reddy (2005) conducted studies on combustion of rice husk in an atmospheric fluidized bed in an environment of high excess air (air fuel ratios 8.3 to 12.6) and found that when the flow rate of air was increased to 22.4 L/s (94.39% excess air) a maximum temperature of $800^o C$ was attained in the bed. The temperature near the distributor plate was limited to $200^o C$ to $300^o C$ and temperature gradient along the bed height was found.

In the literature it was found that a wide variety of fuels could be burnt in a fluidized bed. Till now most of the research work has concentrated on the combustion of rice husk only. There is ample scope to investigate combustion behaviour of other agro-waste products such as sawdust, groundnut shells, coconut shells, etc. In most of the cases freeboard burning has been very high and measures have been taken to reduce freeboard burning of the fuel. From the results of experiments reviewed, it could be concluded that due to sudden devolatilization of biomass fuels, combustion operations should be conducted in an environment of high excess air. Splitting the total air into primary air and secondary air and supplying them at the appropriate levels could improve combustion efficiency. It is also necessary to reduce harmful emissions from the combustion of biomass fuels. High levels of CO emissions in the flue gas are still reported.

EXPERIMENTAL SETUP

A circular cross-section fluidized bed combustor was fabricated with an inner diameter of 150 mm and a main combustion chamber with a height of 1000 mm using stainless steel material with a thickness of 3 mm. A pressure reduction vessel with a height of 500 mm and a diameter of 350 mm is attached to the main chamber, which is in turn connected to the cyclone separator. The inner side of the vessel is coated with a castable refractory material to minimize heat losses from the vessel. A copper tube with a diameter of 12 mm is wound around the vessel to control the bed temperature by circulating cool water through the coil. The outer side of the vessel is covered with rock wool as an insulation material to reduce the heat transfer from the vessel. Thermocouples and pressure taps are provided at various points to measure temperature and pressure respectively. A provision is also made to supply secondary air. A blower, which is connected to 10 H.P. D.C motor, is used to supply the primary and secondary air. A screw feeder is used to feed the required fuel into the combustion chamber. After the combustion, flue gases and solid ash particles are separated in the cyclone separator. A schematic diagram of the fluidized bed combustor is presented in Fig. 1.

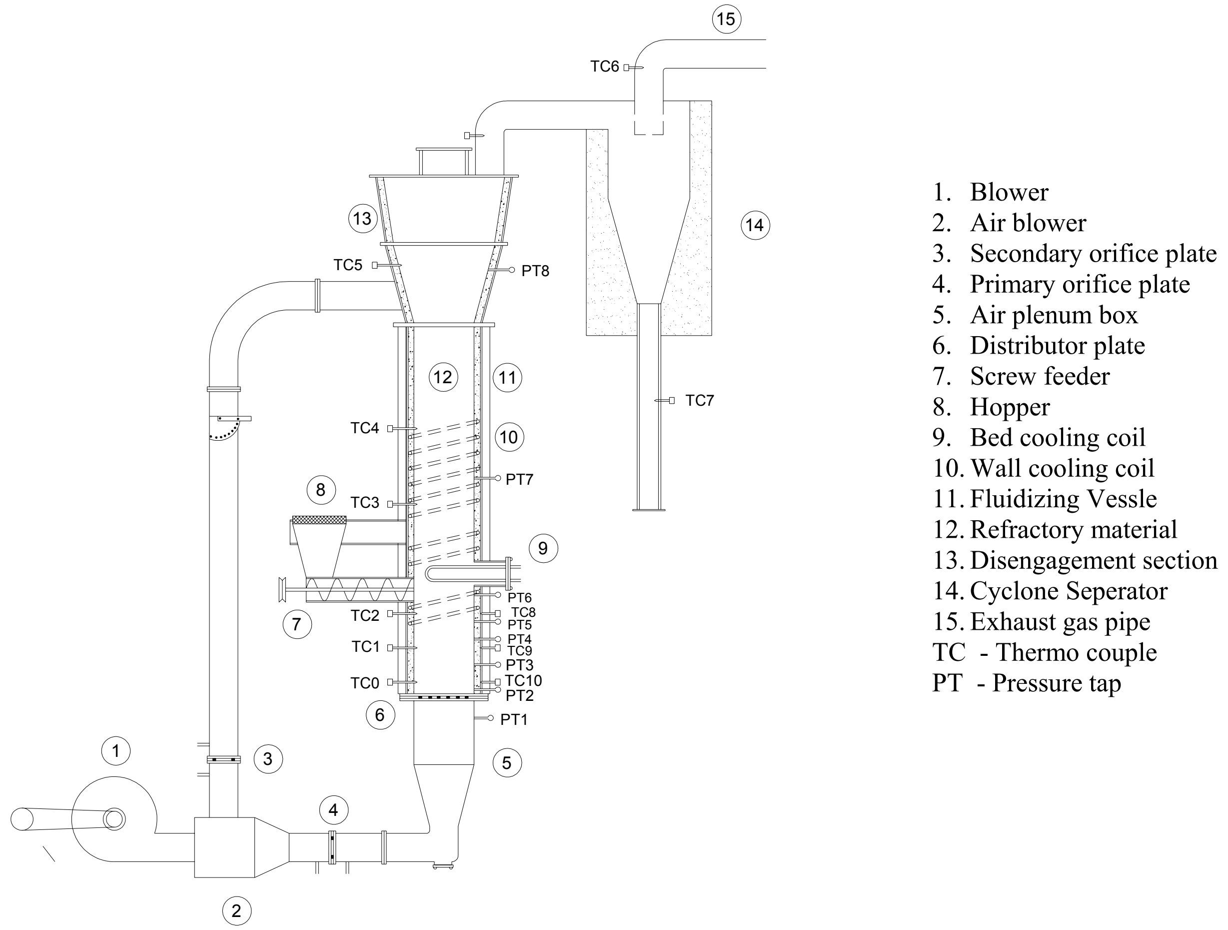

Figure 1: Experimental setup for combustion studies

EXPERIMENTATION

Calculation of Mean Particle Size of Sand

In the present investigation six sieves are selected with aperture sizes of 1000, 850, 710, 600, 425 and 355 μm as per the specifications of IS: 460 (part-1) – 1985 to get the suitable size of sand. The sieves are arranged in descending order of their dimensions with a pan at the end. Then the set of sieves is installed in the sieve shaker machine and sand is poured on top of the first sieve. The machine is first operated for 10 minutes and sand passing through 1000 μm and retained at 850 μm is collected. The average aperture size 925 μm is considered as the mean diameter of the sand particles. Similarly the mean particle diameter is 655 μm for the sand passing through 710 μm and retained at 600 μm and the mean particle diameter is 400 μm for the sand passing through 425 μm and retained at 355 μm. From the hydrodynamic analysis, 400 μm is found to be suitable for better fuel particles mixing characteristics.

Preparation of True Sample of Fuel

Selecting a sample from a single pile of fuel material at the top of a bag an improper sampling. Therefore, a coning and quartering method [IS 436 (Part II)-1965] is adopted for sample selection and the same is used for density calculation, particle size measurement and proximate and ultimate analysis.

Measurement of Fuel Particle Size

In the experimental study, eight sieves with aperture sizes of 1000, 850, 710, 600, 425, 355, 250 and 180 μm are selected as per standards given in IS: 460 (part-1) -1965. The set is then kept in the sieve shaker machine and 250 gm of sawdust is poured into the first sieve and the machine is operated for 15 minutes. The mass of sawdust is collected in each pan and is given in Table 1. The corresponding mass fractions of the particles of given sizes are presented in Table 2. The mean diameter of the sawdust is found to be 578 μm.

Table 1: Sieve analysis of a sample of particles of saw dust

Mass of particles retained in sieve, g	Sieve aperture size, μm
0	1000
10.6	850
41.2	710
72.5	600
100.8	425
15.3	355
9.6	250
0	180

Table 2: Ratio of mass fraction to range of diameter

Range of Diameters (μ)	d_{pi} (μ)	Mass fraction in interval x_i	$(x/d_p)_i$
1000-850	925	10.6/250 = 0.0424	$0.0424/925 = 4.58 \times 10^{-5}$
850-710	780	41.2/250 = 0.1648	$= 2.10 \times 10^{-4}$
710-600	655	= 0.29	$= 4.43 \times 10^{-4}$
600-425	512.5	= 0.4032	$= 7.86 \times 10^{-4}$
425-355	390	= 0.0612	$= 1.57 \times 10^{-4}$
355-250	302.5	= 0.0384	$=1.27 \times 10^{-4}$
			$\Sigma(x/d_p)_i = 1.73 \times 10^{-3}$

Design of Gas Distributor

The sphericity, ϕ, and voidage at minimum fluidization velocity, ε_{mf}, corresponding to sand particle size of 0.4 mm taken as 0.67 and 0.49 (Kunii and Levenspiel, 1969). In addition the following data is used for design of the gas distributor plate.

$L_s = 0.1$ m

Area of vessel = 0.01767 m^2

Pressure drop across the bed at minimum fluidization is given as

$$\Delta p = L_{mf} \times (1- \varepsilon_{mf}) \times (\rho_p - \rho_g) \times g \tag{1}$$

$$L_{mf} = \{(1-\varepsilon)/(1-\varepsilon_{mf})\} \times L_s \tag{2}$$

$$\Delta p_d = 0.1 \times \Delta p \tag{3}$$

$$U_o = C_d \, (2.\Delta p_d/\rho_g)^{0.5} \tag{4}$$

$$U_{mf} = \pi/4 \times d_{or}^{\,2} \times U_o \times N_o \tag{5}$$

Minimum fluidization velocity is calculated from the relation

$$Re_{mf} = d_p \, U_{mf} \, \rho_g/\mu = Ar / \{(150(1-\varepsilon_{mf})/\varepsilon_{mf}^3 + \sqrt{1.75 \, Ar/\varepsilon_{mf}^3}\} \tag{6}$$

where Ar = Archimedes number = $d_p^3 \, \rho_g \, (\rho_p - \rho_g) \, g/\mu_g^2$

From the equations 1 to 6, the number of orifices per square meter area is

$N_o = 62247.266$ holes/m^2

No. of holes per plate = 62247.266 ×0.01767 = 1100 holes.

The distributor plate is of the straight multiorifice type with an opening of 7.6%. The distributor plate is designed in such a way that the primary air passing through the bed in the minimum fluidization state is nearly equal to the theoretical air required for the combustion of fuel. As the velocity of air is increased beyond the minimum fluidization state, the excess air is in the form of bubbles and the bed becomes agitated due to high solid-gas contact.

Measurement of Pressure Drop

To measure pressure drop across the bed and distributor, eight pressure taps are provided along the axis of the vessel. The first pressure tap is provided at 50 mm below the distributor plate and the other pressure taps are located at heights of 30, 80, 150, 220, 290, 650 and 1150 mm above the distributor. All the pressure taps are connected to a water-filled manometer bank.

Measurement of Temperature

For the present study an alloy of chromel-alumel (ANSI symbol K) of gauge 8 (3.2 mm) is used. As the operating range for this alloy is -200^{o} C to 1260^{o} C it is also recommended for use in oxidizing atmospheres. To measure the temperature at different points, 14 thermocouples are arranged throughout the bed. Six thermocouples are inserted into the vessel at heights of 40, 150, 250, 600, 825 and 1150 mm and the other six thermocouples are used to measure the surface temperatures of the vessel at heights of 150, 250 and 600 mm above the distributor plate. One thermocouple is located in the flue gas passage pipeline and another thermocouple is used to measure the temperature of the ash collected at the bottom of the cyclone separator.

Measurement of Minimum Fluidization Velocity

Initially sand with a mean particle size of 0.4 mm is fluidized to a bed height of 100 mm. The purpose of the inert bed of solids is to avoid overheating the distributor plate and also to retain the heat generated during the combustion process. By introducing the cooling coil into the bed, the heat can be transferred to the water circulating through the coil. In the present study, a bed height of 100 mm is selected for the experimentation. Fuel particles are introduced to a height of 200 mm above the bed of solids. Characteristics of the bed material are presented in Table 3. The properties of the fuel particles are given in Table 4. The proximate and ultimate analysis of sawdust are given in Table 5 and Table 6. The flow rate of air through the bed is increased gradually until the pressure drop is constant, which indicates that the bed has reached fluidization state. Superficial velocity is plotted on the X-axis and pressure drop on the Y-axis. The minimum fluidization velocity is obtained from the plot of superficial velocity versus pressure drop.

Table 3: Characteristics of bed material used for experimentation

S. No	Parameter	Sand
1	Mean particle size, d_p in mm	0.404
2	Max. size of particle, mm	0.542
3	Min. size of particle, mm	0.27
4	Particle density, ρ_p in Kg/m^3	2519
5	Bulk density, ρ_b in Kg/m^3	1600
6	Terminal velocity of the particle, U_t in m/s	3.18
7	Static voidage , ε_o	0.36
8	Spherocity, ϕ_s	0.67

Table 4: Properties of sawdust

S. No.	Property	Saw dust
1	Mean particle size, mm	0.578
2	Bulk density, kg/ m^3	286.4
3	Particle density, kg/ m^3	716.2
4	Calorific value, kcal/kg	4464

Table 5: Proximate analysis of fuels

S. No.	Property	Saw dust
1	Moisture, %	8.15
2	Volatile matter, %	81.17
3	Ash, %	1.994
4	Fixed Carbon, %	8.686

Table 6: Ultimate analysis of fuels

S. No.	Property	Saw dust
1	Carbon, %	48.496
2	Hydrogen, %	3.96
3	Oxygen, %	27.15
4	Nitrogen, %	0.24
5	Sulpher, %	0.01

Initial Start-Up of the Bed

Before the fuel is burnt in the fluidized bed, it is necessary to preheat the inert bed of solids to about $500^{o}\,C$ using an auxiliary heating system. An over-bed LPG burner with a small amount of oxygen is directed at the surface of the bed for the purpose of preheating. After the bed reaches the required temperature, fuel particles are gradually supplied through the screw feeder to the bed. The initial startup system is removed from the bed once the temperature of the bed reaches $500^{o}\,C$. The feed rate is maintained until the bed reaches a stable temperature. Normally about 2 to 3 hours are required for preheating the bed. A considerable amount of visible smoke and emissions are observed after initiating the fuel feed until the bed reaches a temperature of 600 to $600^{o}\,C$.

Fuel Feed Control

The fuel is fed in through the screw feeder at a desired rate, which is varied from 3.2 to 10.4 kg/hr. The combustion intensity is defined as the rate of feeding fuel into the fluidization vessel per unit area available for burning. Combustion intensities ranging from 208 to 675 kg/ m^2-hr are attained. The fuel feed rate is gradually increased till ash is accumulated over the bed. As this occurs, the bed is defluidized. This condition is indicated by a sudden reduction in pressure across the bed. For sawdust the maximum fuel feed rate observed is 10.2 kg/hr.

Operating Procedure

After initial startup of the bed, the flow rate of primary air is increased till the bed reaches the minimum fluidization state. The operating conditions maintained in the bed are presented in Table 7. The theoretical air required for the combustion of fuel particles is calculated from the ultimate analysis and the excess air supplied at each flow rate is determined. Then for every fuel feed rate temperatures are recorded throughout the vessel with the use of thermocouples. To determine the carbon loss during the combustion process, ash is collected from the cyclone separator and proximate analysis is carried out to determine the carbon present in the residual ash. The carbon monoxide in the flue gas is measured using a flue gas analyzer, which gives percentage of volume of CO per unit volume of flue gas released through the exhaust end. From the ultimate analysis the carbon content of the fuel can be assessed. The ash agglomeration over the bed results in defluidization of the bed and can be observed in the form of pressure fluctuations, which can be monitored by U- tube manometers.

The combustion efficiency is thus calculated with the following equation:

Combustion efficiency = [Calorific value of rice husk – (Heating value of refuse at cyclone separator + Heating value in the flue gasses)]/ Calorific value of rice husk

$$\eta_c = \frac{CV - (HVR + HVG)}{CV} \qquad (A.11)$$

$$\eta_c = \frac{CV - [(W_c \times CVC) + (W_{co} \times CVCO)]}{CV} \quad (A.12)$$

where

CV = Calorific value of fuel, kcal/kg
CVC = Calorific value of carbon, kcal/kg
CVCO = Calorific value of carbon monoxide, kcal/kg
W_c = Weight of carbon in refuse, kg
W_{co} = Weight of carbon monoxide in flue gases, kg

Uncertainty analysis

The uncertainties associated with the experimental data are estimated in this section. Let us assume that z is a given function of the independent variables (x_1, x_2, x_3, x_4,………, x_n). Let ω_z be the uncertainty in z and ω_1, ω_2, ω_3, ω_4, ……………..,ω_n be the uncertainty in the dependent variables. If the uncertainties in the independent variables are given at the same odds, then the uncertainty of z having these odds is given by Holman (1984) as

$$\omega_h = \left[\left(\frac{\partial z}{\partial x_1}\omega_1\right)^2 + \left(\frac{\partial z}{\partial x_2}\omega_2\right)^2 + \dots\dots\dots\dots + \left(\frac{\partial z}{\partial x_n}\omega_n\right)^2\right]^{\frac{1}{2}} \quad (7)$$

$$\frac{\omega_h}{h} = \left[\left(\frac{\omega_1}{x_1}\right)^2 + \left(\frac{\omega_2}{x_2}\right)^2 + \dots\dots\dots\dots + \left(\frac{\omega_n}{x_n}\right)^2\right]^{\frac{1}{2}} \quad (8)$$

From the above equations it is found that the uncertainties in superficial velocity are in the range of 8.09 to 11.116% and the uncertainties encountered in experimental investigations of combustion efficiency for different fuels and under different operating conditions are in the range of 0.341% to 0.6578%.

Table 7: Operating conditions of fluidized bed

S. No	Parameter	Sand
1	Air flow rate, L/s	3.12-18.47
2	Superficial velocities in the vessel, U_g in m/s	0.18-1.05
3	Superficial velocities in Disengagement section, m/s	0.04-0.26
4	Voidage at minimum fluidization state, ε_{mf}	0.49
5	Pressure drop at minimum fluidization, $(\Delta p)_{mf}$ in N/m^2	1532.63
6	Static bed height of inert particles, L_s in mm	100
7	Minimum fluidization velocity, U_{mf} in m/s	0.66
8	Bed height at minimum fluidization, L_{mf} in mm	124

RESULTS AND DISCUSSION

Effect of Fluidizing Velocity on the Temperature Profile in the Combustion Chamber

As the fuel is fed into the fluidized bed, due to high heat transfer from the bed, the fuel particles get heated up rapidly and release a large amount of volatile matter. A high percentage of volatile matter burns in the freeboard, while the high-carbon-content char left after emission of volatile matter is burnt within the bed. The temperature profile for the sawdust with an increase in fluidizing velocity along the vertical height above the distributor plate is shown in Fig. 2. A constant feed rate of 10.2 kg/hr is maintained for the sawdust. It is evident that with the increase in fluidizing velocity the temperatures at all the locations increase. The experiments showed that a temperature gradient exists inside the bed at low fluidizing velocities. One possible reason is that low velocities result in inadequate mixing. As the fluidizing velocity is increased further, a strong combustion intensity zone moves to the top of the freeboard and losses in unburnt combustibles are found to increase. Temperatures are observed near the distributor zone of around 630 °C for sawdust. For all fluidizing velocities the maximum temperature is obtained at a height of 250 mm from the bed and then a gradual drop in temperature up to a height of 825 mm from the bed is observed. Thereafter a rise in temperature is observed due to disengagement of some of the fuel particles from the flue gas and partial combustion of fuel in this section. Similar tendencies were also observed by Preto et al. (1987) during the combustion of rice husk and wood chips. There is increased freeboard combustion during over-bed feeding, which results in an increased freeboard temperature. The active combustion of fuel particles took place at heights between 150 and 600 mm above the distributor plate. Armesto et al. (2002) carried out combustion of rice husk in a fluidized bed combustor and observed a maximum temperature at a height of 400 mm from the distribution plate. In the present work over a height of 600 mm above the distributor plate a gradual reduction in temperature is observed till the height of 1000 mm. In the disengagement section, fuel particle velocity is reduced due to enlarged area and that causes further combustion of fuel particles. Therefore, again a slight rise in temperature is found in the disengagement section.

Effect of Feed Rate on the Temperature Profile in the Combustion Chamber

After initial startup of the bed, the primary airflow rate as well as the feed rate of fuel are gradually increased to sustain the combustion process. The feed rate of sawdust corresponding to the minimum fluidization state is 6.8 kg/h. In this minimum fluidization state the airflow rate is kept constant and the feed rate of fuel is increased till the maximum possible temperatures are attained at all the locations of the combustion chamber along the axial height. This temperature profile is depicted in Fig. 3. At lower feed rates the temperatures are almost uniform at a height between 150 and 600mm. At heights above 600mm from the distributor plate, there is a slight reduction in temperature. Another notable feature is that at low feed rates no temperature rise in the enlarged freeboard is observed, which indicates that no combustion is taking place in this section and the burning of particles is completed before it reaches the enlarged section. As the feed rate is increased further the temperature profile changes slightly and the peak temperatures are obtained at a height of 250mm for sawdust. There is also a slight increase in freeboard temperatures as the feed rate is increased.

The feed rate of sawdust corresponding to the maximum possible temperature is observed at 10.2 kg/h. As the feed rate is increased further, ash agglomeration is found to take place over the bed and this results in defluidization of the bed. Defluidization can be identified by a sudden reduction in pressure across the bed. Combustion intensity is defined as feed rate of fuel per unit of cross-sectional area for complete combustion within the bed. The maximum combustion intensity that can be achieved with sawdust in the fluidized bed is found to be 577.2 kg/h-m^2. Singh et al. (1980) reported that the maximum combustion intensity of rice husk in a grate-type furnace is about 70 kg/h-m^2. In a fluidized bed reactor Bhattacharya et al. (1983) achieved a combustion intensity of 530 kg/h-m^2 with rice husk fuel.

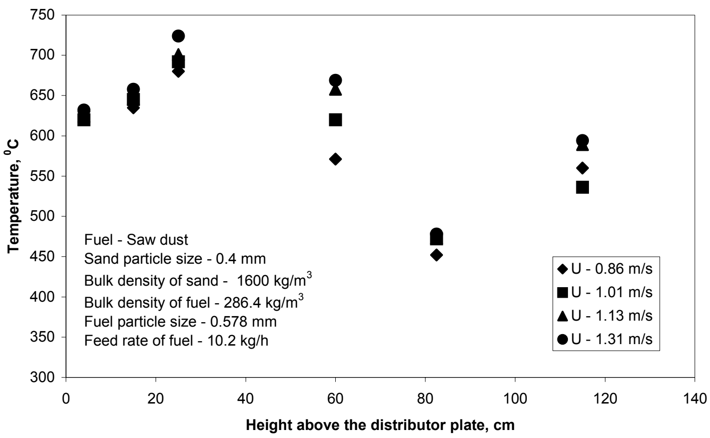

Figure 2: Temperature profiles along axial height above the distributor plate at different fluidizing velocities

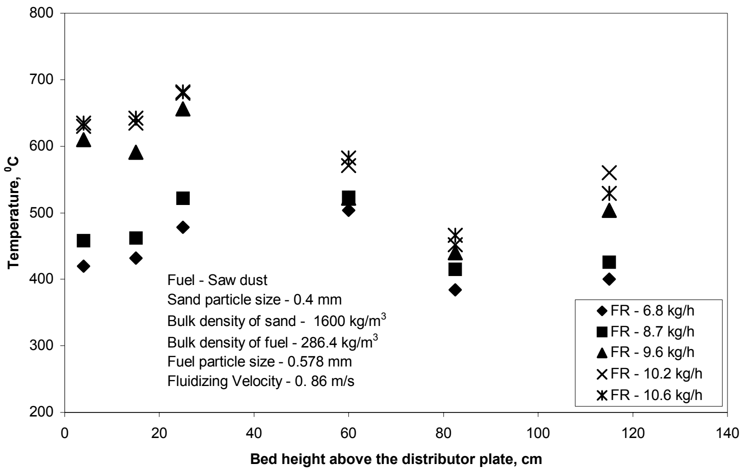

Figure 3: Temperature profiles along axial height above the distributor plate at different feed rates and at constant fluidizing velocity of 0.64 m/s

Effect of Secondary Air on the Temperature Profile in the Combustion Chamber

To achieve higher combustion efficiency, the primary air is passed through the distributor plate at a rate slightly higher than that required for char combustion and the remaining excess air is supplied as secondary air at the entrance of the enlarged section of the freeboard. A substantial increase in the temperatures throughout the combustor is observed as shown in Fig. 4. A considerable rise in temperature in the enlarged freeboard is also observed. This indicates that air staging can improve combustion. When the secondary air is injected, the temperatures at different heights are observed to be more uniform than without secondary air.

Effect of Excess Air on Carbon Carry-Over Loss in the Combustion Chamber

The distributor plate is designed in such a way that the primary airflow rate in the minimum fluidization state is approximately equal to the theoretical air required for complete burning of fuel particles. As the primary airflow rate is increased beyond the minimum fluidization state, the quantity of excess air also increases. During the combustion process a large percentage of volatile matter released from the fuel gets burnt along the bed height, while a small amount of high-carbon-content char is left within the bed. The fine ash particles carried along with the flue gas are separated from the gas in the cyclone separator. A part of the energy released during the combustion process is lost either through the presence of CO (carbon monoxide) in the flue gas or in the form of unburnt combustibles along with the ash. In Fig. 5 heat loss owing to the incomplete combustion of fuel is plotted against fluidizing velocity when the fuels are fired at the maximum combustor loading. With an increase in air velocity the percentage of CO is found to decrease for the sawdust fuel. Little change in CO levels is observed beyond the fluidizing velocity of 1.1 m/s. The minimum CO emission observed for sawdust is 0.845%. The higher CO levels for sawdust are attributed to the fact that incomplete combustion of volatile matter of these fuels means that fluidizing velocity influences these emissions. Figure 5 also shows the effect of heat loss owing to carbon carryover along with ash against fluidizing velocity. As the velocity of air is increased the carbon char loss is also found to increase.

Secondary air may be injected into the freeboard above the fuel bed to intensify mixing of combustion air with the volatiles, thus enhancing gas phase combustion reactions. To minimize the further formation of CO, the excess air is supplied in the form of secondary air by maintaining the primary airflow rate just above the minimum fluidization state. The corresponding heat loss against excess air for the same fuel loading is shown in Fig. 6. The CO levels are observed to decrease drastically, even at low excess air. This reduction in CO levels may be ascribed to the high turbulence created by the supply of secondary air, which results in better combustion of volatile matter. The loss of carbon char also found to decrease significantly with the supply of excess air as secondary air near the enlarged freeboard region, as depicted in Fig. 6. For sawdust the carbon loss is in the range of 0.4 to 0.1% with the increase in excess air.

From the above discussion, it is evident that the increase in excess air through the provision of secondary air gives better results than the increase in primary flow rate through the distributor plate. The enlarged freeboard has considerable influence in reducing the formation of CO and better combustion of char particles. Fifty to 60% excess air is found to be optimal for reducing carbon loss in the burning the sawdust.

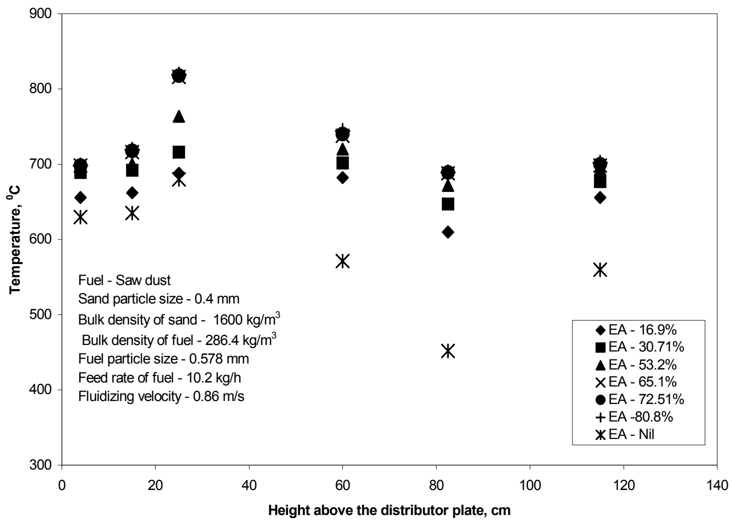

Figure 4: Temperature profiles along axial height above the distributor plate at different excess air flow rates

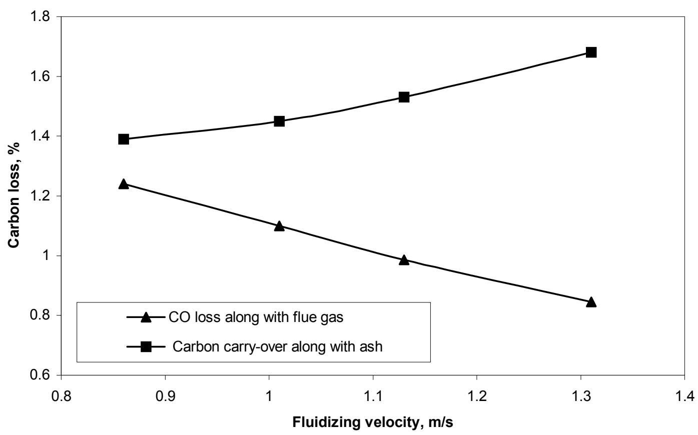

Figure 5: Effect of fluidizing velocity on carbon loss during the combustion of sawdust

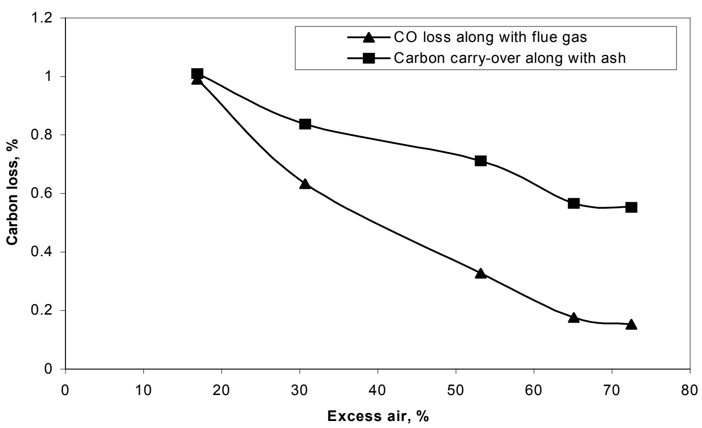

Figure 6: Effect of excess air on carbon loss during the combustion of sawdust

Effect of Operating Conditions on Combustion Efficiency

For the estimation of combustion efficiency, the heat losses owing to incomplete combustion (accounting for the CO emission) and unburnt carbon in fly ash are determined. Fig. 7 shows the combustion efficiency for the sawdust fuel at maximum combustor loads for different fluidizing velocities. The maximum possible combustion efficiency for sawdust is 97.4% at a fluidizing velocity of 1.13 m/s. In Fig. 8 it can be observed that only a slight increase in combustion efficiencies is observed for sawdust because of the low ash-fuel ratio, even though there is a considerable reduction of carbon losses. Permchart and Kouprianov (2004) reported that for the maximum combustor load and with excess air of 50 to 100%, a combustion efficiency of over 99% could be achieved when sawdust and bagasse were fired. In the present study the maximum combustion efficiency is found to be 99.2% for 65% at excess air, which results in higher combustion efficiency at a relatively small quantity of excess air with a minimum heat loss by the exhaust gases. The reason for higher combustion efficiency in the present study is due to modifications of the fluidization vessel and the quantity of excess air supplied in the disengagement section.

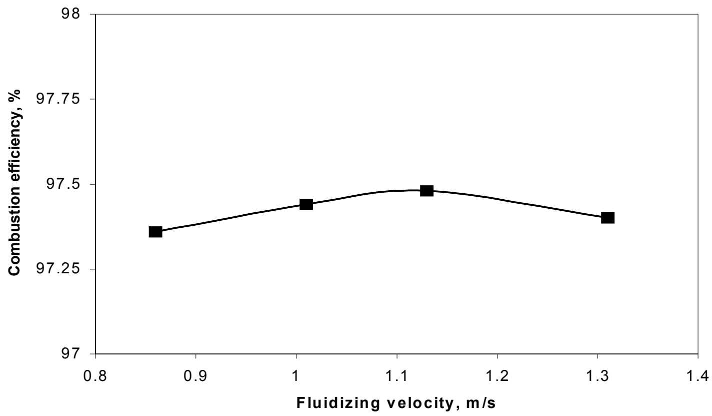

Figure 7: Effect of fluidizing velocity on combustion efficiency

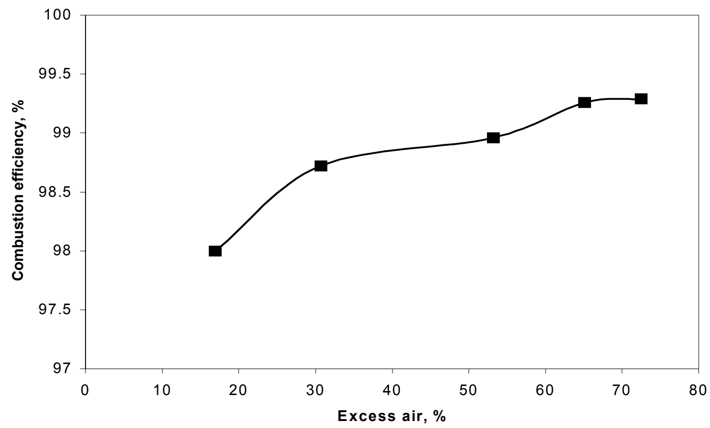

Figure 8: Effect of excess air on combustion efficiency

CONCLUSIONS

1. The axial temperature profiles along the sand and fuel mixture bed (i.e., up to 15 cm) are fairly uniform under all operating conditions. Above the bed, i.e., in the freeboard zone of the combustion chamber of the fluidized bed, variations in temperature are observed.

2. The maximum temperature at all the velocities is obtained at a height of 250 mm from the bed and then a gradual reduction in temperature up to a height of 825 mm occurs. Again a rise in temperature at a height of 1150 mm above the distributor plate is observed, which indicates considerable burning of fuel in the freeboard and in the enlarged disengagement section.

3. With the supply of secondary air before the enlarged section of the freeboard, a substantial increase in temperature throughout the combustor is observed. The effect of secondary air on the temperature profile is predominant due to effective burning of fine sawdust particles in the freeboard.

4. With an increase in air velocity, the percentage of CO is found to decrease. The injection of secondary air into the freeboard above the fuel bed creates an intense mixing of combustion air with the volatiles, thus reducing the CO levels in the flue gas. The enlarged freeboard has a considerable influence in reducing the formation of CO, in addition to improving the combustion of char particles. Between 50 and 60% excess air is found to be optimal in reducing carbon loss during the burning of sawdust.

5. A maximum possible combustion efficiency of 99.25 % could be achieved with the enlarged freeboard and the supply of secondary air in the freeboard zone.

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NOMENCLATURE

A	bed cross-sectional area	m^2
Ar	Archimedes number	$\frac{g\rho_g(\rho_p-\rho_g)D_p^3}{\mu^2}$
d_p	diameter of the particle	m
d_{or}	orifice diameter	m
g	acceleration due to gravity	m/s^2
L_s	length of static bed	m
L_{mf}	length of bed at minimum fluidization	m
N_o	number of holes per unit surface area of distributor	(-)
Re_{mf}	Reynolds number at minimum fluidization	(-)
U	superficial velocity of air	m/sec
U_{mf}	minimum fluidization velocity	m/sec
U_o	orifice gas velocity	m/s

Greek Letters

ε	bed voidage	(-)
ε_{mf}	bed voidage at minimum fluidization	(-)
μ	gas viscosity	$N\ s/\ m^2$
ϕ	sphericity	
ρ_g	fluid or gas density	kg/m^3
ρ_p	density of the particle	kg/m^3
Δp	pressure drop across the bed	N/m^2
Δp_d	pressure drop across the distributor	N/m^2

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