FLUIDIZATION BEHAVIOR OF SOLID PARTICLES IN FAST PYROLYSIS REACTOR FOR THERMOCHEMICAL CONVERSION OF BIOMASS SOLID WASTES

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ABSTRACT

Among various methods of thermochemical conversion of organic solid wastes into bio-crude, fluidized bed pyrolysis system has attained much attention. In this study one cylindrical cold model fluidized bed reactor has been constructed from fiberglass pipe. Different tuyere-type gas distributors of 6, 9 and 12 standpipes, with 4 holes on each pipe have been fabricated. The river sand of different particle sizes has been used as bed material. The fluidizing gas was dry air, supplied by a compressor. The airflow rate and pressure drop across the beds of different height were measured. From this study optimum values of fluidization parameters in a fast pyrolysis reactor: bed particle size, bed height, minimum fluidizing velocity, bed particle entrainment and elutriation, behavior of solid mixing and suitable distributor plate have been found out. From the experimental results a mathematical model has been developed. The results obtained by the cylindrical model are compared with the relevant work and those from the mathematical model. These values of the parameters are proposed to design and fabricate a laboratory scale fluidized bed pyrolysis reactor to convert biomass solid wastes into pyrolysis liquid oil, which may be used as an alternative fuel.

INTRODUCTION

Since the advent of the energy crisis associated with environmental concern due to waste disposal, increasing emphasis has been placed on exploring the possibilities of recovering energy from biomass solid waste. Biomass solid waste is attractive from the point of view of ease of availability, high carbon content, low moisture and ash content, low or even no cost, no conflict arising from alternate usage, solving solid waste disposal problems and keeping the environment clean [Islam and Ani, 2000]. In some cases, it may have some existing usage; however, there may be better utilization and application from the points of view of energy recovery and environment that need to be emphasized. There are various biomass solid wastes available in Bangladesh. A few examples are: rice-straw, sugarcane bagasse, jute-stick, wheat-straw, sawdust, rice-husk, empty fruit bunches, livestock, scrap tyre, refused plastic, wastepaper etc. These carbonaceous solid wastes are renewable energy sources and therefore, the potential of converting them into useful energy such as liquid fuel, should be seriously considered. In this way, the wastes would be more readily usable and environmentally more acceptable. It is found from the characterization of these biomass solid wastes that these solid wastes contain higher percentage of volatile matter [Islam et al 2001; Islam et al 2001]. These high volatile-content biomass have high potential for pyrolysis liquid production. Fluidized bed pyrolysis is more attractive among various thermochemical conversion process because of its higher conversion capability of biomass and its solid wastes into liquid product [Bridgewater and Bridge, 1991; Zailani, 1995]. Fluidized bed reactor utilizes inert solid particles in an inert gaseous atmosphere as heat carrier. Fluidization allows a complete mixture of inert bed particles and feedstock and hence, producing high heat transfer. Prior to design a fluidized bed reactor, some basic hydrodynamic studies on its cold model are usually required to find out the suitable operating conditions of the reactor. Thus, there is a great possibility of producing large amount of bio-crude from these biomass solid wastes by suitable design of a fluidized bed pyrolysis system. The most important process parameters for fluidized bed system are fluidizing gas velocity,

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bed height, bed particle size, distributor plate, bed particle entrainment and elutriation, and behavior of solid mixing [Buekens and Schoeters, 1984]. These parameters have been designated by the cold model study of the fluidized bed reactor.

FLUIDIZATION MECHANISM IN THE SAND BED

Generally, if a gas is passed upward through a bed of fine particles at a low flow rate, the gas percolates through the void spaces between stationary particles. With an increase in the superficial gas velocity, defined as the volume flow rate of the gas per unit cross sectional area of the empty container holding the bed, particles move apart and a few vibrate and move in restricted regions, allowing the bed to expand and causing the pressure across the column of the bed ΔP, to increase. At further higher gas velocity, a point is reached where all particles are just suspended by the force exerted by the upward flowing gas. At this point, the frictional force between the particles and gas just counter balances the weight of the particles. The vertical component of the compressive force between adjacent particles disappears; resulting in the pressure drop through any section of the bed equals the weight of gas and particles in the section. The superficial gas velocity through the column of the bed, responsible for this state is called the minimum fluidization velocity, U_{mf} and its value depends on the physical properties of the gas and the sand particles.

At gas velocity just above U_{mf}, bubbles are formed at the orifice where the fluidizing gas enters the bed. These small gas bubbles grow and coalesce as they rise through the bed surface where they burst, splashing the particles into the freeboard zone, a region above the bed. At higher fluidizing gas velocity, U_f all particles in the bed are in motion, and increasing the gas velocity leads to particles moving in a more rapid independent motion. Increasing gas velocity further will only cause incremental increase in ΔP. More vigorous bubbling can be observed, with bigger and bigger bubbles appearing at the surface. Eventually at very high gas velocity, the terminal velocity, U_t the particles are released, at which the particles will be entrained by upward force of the flowing gas. With a further increase in gas velocity, particles are carried out of the bed with the gas.

Fluidizing a bed of solid particles with a gas provides a means of bringing the two into intimate contact and this can be very useful in gas-solid phase heat transfer where isothermal condition throughout the bed is desired. The rapid movement of the fluidized bed particles also allows fast mixing with foreign solid particles introduced in the bed and hence, it would allow a rapid solid to solid phase heat transfer between the two types of solid particle. These conditions are very ideal for fast heating rate of feedstock, which is a prerequisite criterion of fast pyrolysis process.

EXPERIMENTAL SET-UP

The cold model consisted of two chambers: reactor chamber and primary chamber. The inside diameter of the both the chamber was 6.2cm. The height of the reactor and primary chamber were 50cm and 18cm respectively. Both of them were fabricated from 5mm thick fiberglass pipe. Tuyere-type gas distributor separated these two chambers. The distributor plate supported the sand bed in the reactor chamber and provided uniform air distribution. Three distributor plates of diameter 11cm with 6, 9 and 12 tuyere-nozzles were fabricated from 5mm thick acrylic plate. The distributor plate with tuyere nozzles work quite well, in promoting particle circulation in the bed and in preventing back flows of particles to the air plenum [Kunni and Levenspiel, 1991]. The nozzles were made from 10mm diameter brass rod. The design of the nozzle was such that they may prevent back flow of fluidized bed. In this study each nozzle has 4 holes of 1mm diameter. The reactor chamber was filled with river sand as fluidized bed of height 4 to 8cm supported by the distributor plate. The average particle sizes of sand were 75, 181, 256 and 450μm. the fluidizing gas was dry air, supplied from a compressor, euro 30 of model E3-G100. The capacity of the compressor was 400 litters with a maximum pressure of 20 bars. The air from the compressor reached to the sand bed through the affected by particle size. For larger sand particles the value of U_{mf} increases and higher amount of fluidizing gas is required. Figure 5 shows that the minimum fluidization velocity, U_{mf} is independent of the distributor plate i.e. number of nozzles. However, for a certain fluidization velocity, fluidizing gas pressure increases with the number of nozzles. From the visual observations it is found that a uniform particle movement in the sand bed is achieved using distributor plate with more number of nozzles. From the visual observation it was also found that the good fluidization velocity was in between 9.5cm/sec to 17.32cm/sec that is about in the range $2U_{mf} \leq U_f \leq 3.5U_{mf}$ for the following conditions: sand bed height 6cm; sand particle size 150-212μm and 212-300μm; number of nozzles in the distributor plate 9.

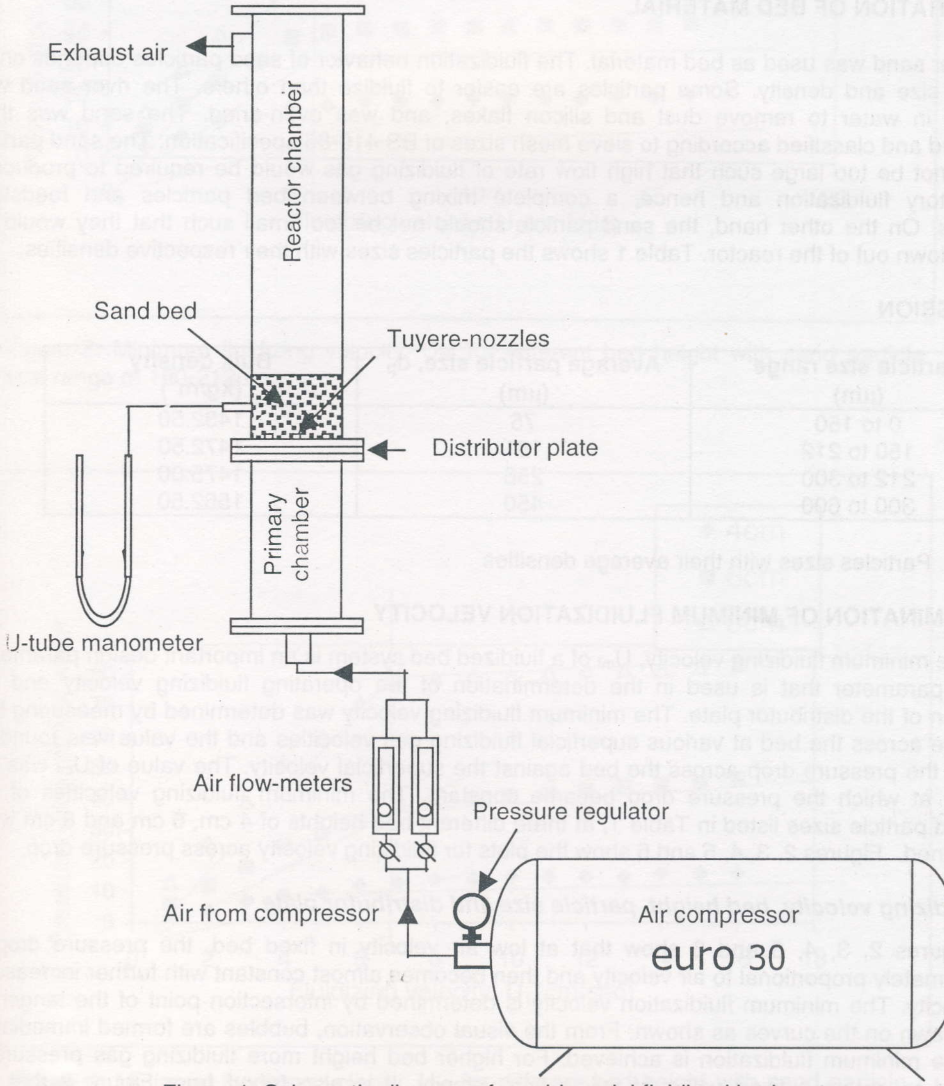

Figure 1. Schematic diagram of a cold model fluidized bed system.

primary chamber and distributor plate with a multi-stage pressure regulator of model MUREX N-10 set at 1 bar. The air flow rate was measured by two identical variable controlled air flow-meters of model MUREX 1040 of capacity 0-40 l/min arranged in parallel. During fluidization the pressure in the sand bed was higher than the atmospheric pressure. An opening of 2.5cm diameter on one side of the reactor at 45cm height from the distributor plate was made to exhaust fluidizing gas. A pipe of inside diameter 4mm was fitted across the sand bed to connect a glass U-tube manometer for measuring pressure drop. The glass U-tube manometer measured the pressure drop across the beds of 4, 6 and 8cm height for each particle size. The schematic diagram of the cylindrical cold model fluidized bed apparatus is shown in Figure 1.

PREPARATION OF BED MATERIAL

The river sand was used as bed material. The fluidization behavior of sand particles depends on its particle size and density. Some particles are easier to fluidize than others. The river sand was washed in water to remove dust and silicon flakes, and was oven-dried. The sand was then screened and classified according to sieve mesh sizes of BS 410-86 specification. The sand particle should not be too large such that high flow rate of fluidizing gas would be required to produce a satisfactory fluidization and hence, a complete mixing between bed particles and feedstock particles. On the other hand, the sand particle should not be too small such that they would be easily blown out of the reactor. Table 1 shows the particles sizes with their respective densities.

DISCUSSION

Particle size range (μm)	Average particle size, d_p (μm)	Bulk density (kg/m^3)
0 to 150	75	1432.50
150 to 212	181	1472.50
212 to 300	256	1475.00
300 to 600	450	1562.50

Table 1. Particles sizes with their average densities

DETERMINATION OF MINIMUM FLUIDIZATION VELOCITY

The minimum fluidizing velocity, U_{mf} of a fluidized bed system is an important design parameter. It is a parameter that is used in the determination of the operating fluidizing velocity and the selection of the distributor plate. The minimum fluidizing velocity was determined by measuring bed pressure across the bed at various superficial fluidizing gas velocities and the value was found by plotting the pressure drop across the bed against the superficial velocity. The value of U_{mf} was the velocity at which the pressure drop became constant. The minimum fluidizing velocities of the selected particle sizes listed in Table 1, at three different bed heights of 4 cm, 6 cm and 8 cm were determined. Figures 2, 3, 4, 5 and 6 show the plots for fluidizing velocity across pressure drop.

A. Fluidizing velocity, bed height, particle size and distributor plate

Figures 2, 3, 4, 5 and 6 show that at low air velocity in fixed bed, the pressure drop is approximately proportional to air velocity and then becomes almost constant with further increase of air velocity. The minimum fluidization velocity is determined by intersection point of the tangential lines drawn on the curves as shown. From the visual observation, bubbles are formed immediately after the minimum fluidization is achieved. For higher bed height more fluidizing gas pressure is required to obtain a fixed value of fluidization velocity. It is also found from Figure 2 that the minimum fluidization velocity is 5cm/sec for particle size of 150-212μm. Figure 4 shows that the minimum fluidization velocity, amount of fluidizing gas and fluidizing pressure are significantly

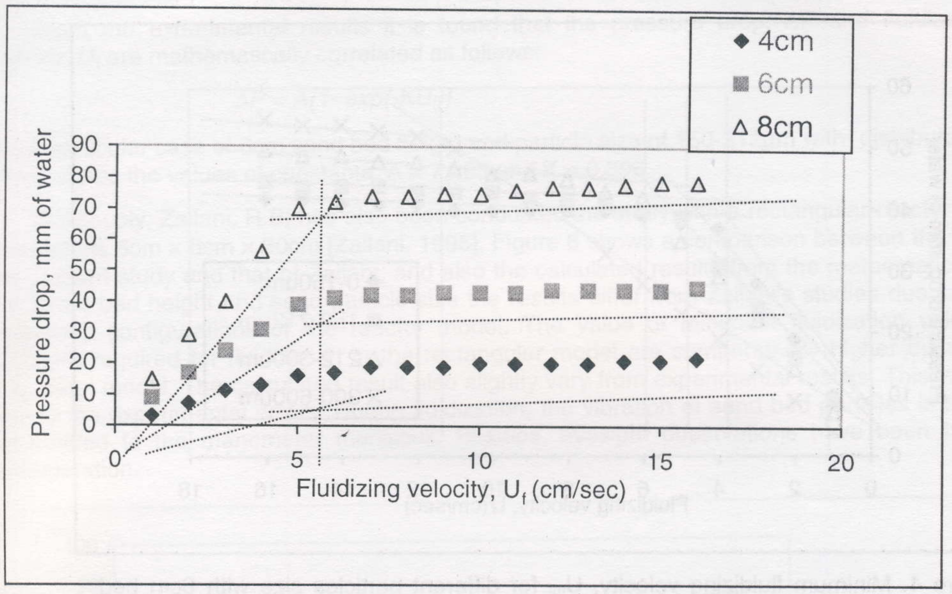

Figure 2. Minimum fluidizing velocity, U_{mf} for different bed height with sand particle size range of 150-212μm.

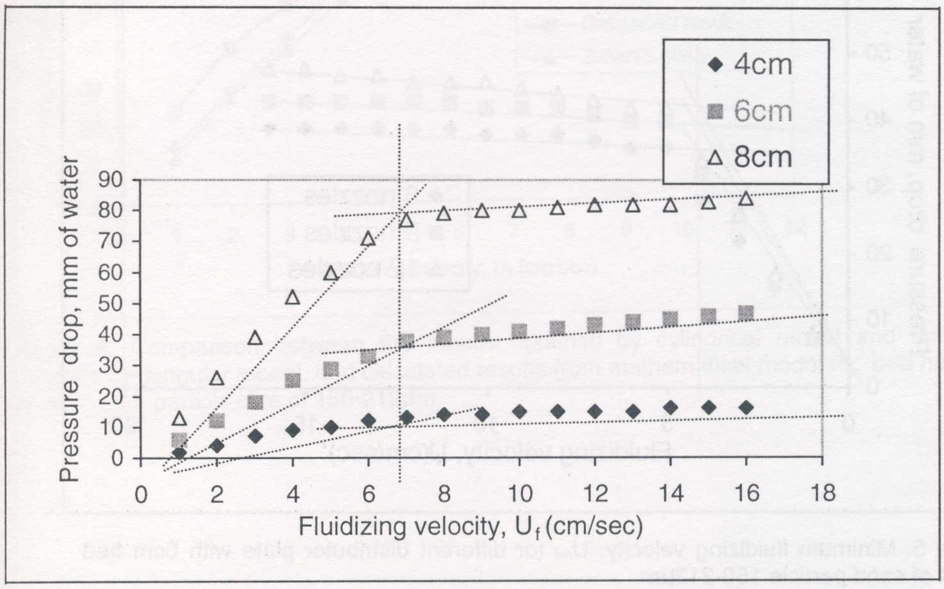

Figure 3. Minimum fluidizing velocity, U_{mf} for different bed height with sand particle size range of 212-300μm.

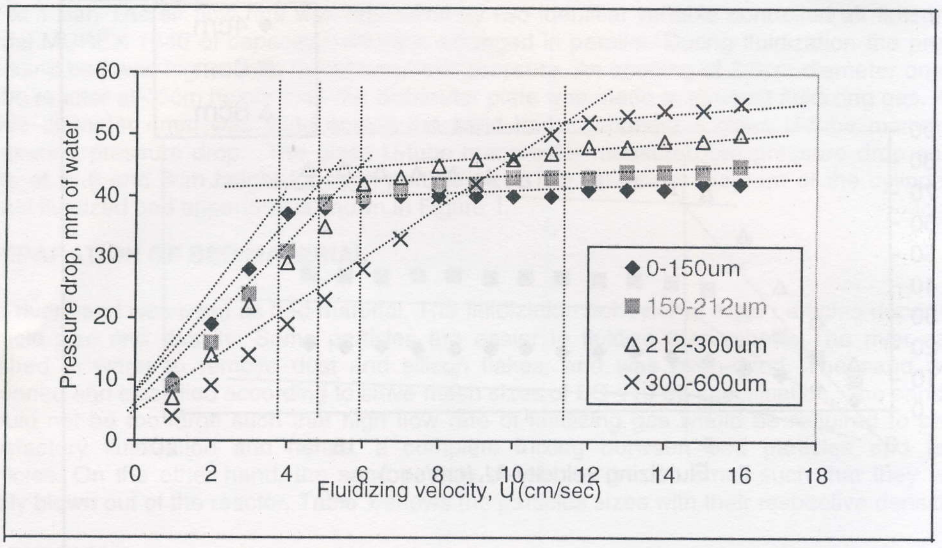

Figure 4. Minimum fluidizing velocity, U_{mf} for different particles size with 6cm bed height.

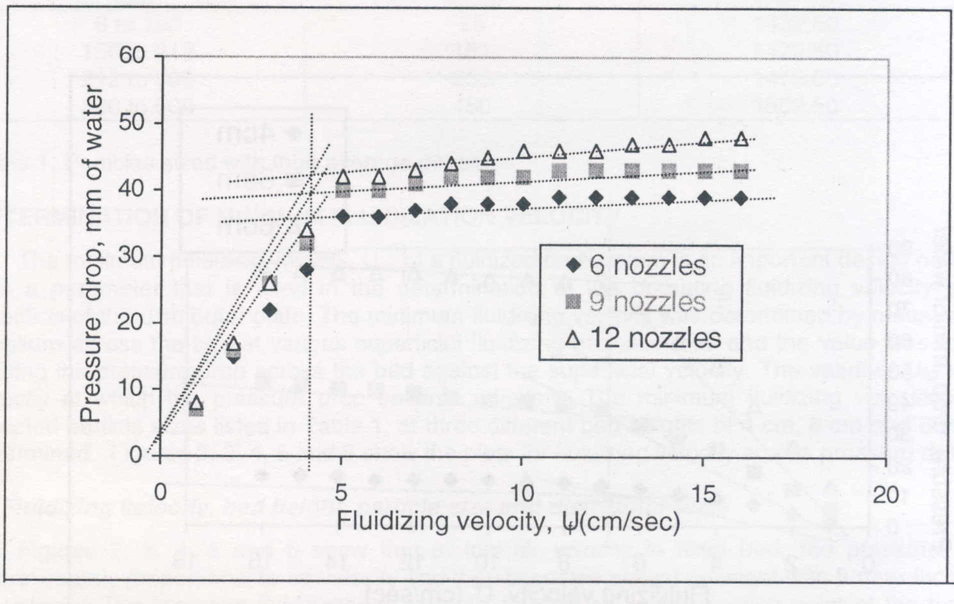

Figure 5. Minimum fluidizing velocity, U_{mf} for different distributor plate with 6cm bed height of sand particle 150-212μm.

B. Mathematical correlation and comparison of results with relevant work

From the experimental results it is found that the pressure drop, ΔP and fluidization gas velocity, U_f are mathematically correlated as follows:

$$\Delta P = A\{1 - \exp(-KU_f)\}$$

For a particular case of 8cm sand bed height and particle size of 150-212μm with distributor plate of nine nozzles the values of constants, A = 7.283 and K = 0.396.

Previously, Zailani, R.B. has also been conducted the study with a rectangular reactor model of dimensions 6cm × 6cm × 30cm [Zailani, 1995]. Figure 6 shows a comparison between the results of the present study and that of Zailani, and also the calculated results from the mathematical model. For same bed height and sand particle size the results differ from Zailani's studies due to different geometric configurations of the reactor model. The value of minimum fluidization velocity and pressure required for fluidization in the rectangular model are comparatively higher than those of cylindrical model. The calculated result also slightly vary from experimental results. This variation is due to the experimental errors. During fluidization, the vibration of sand bed particles is somewhat transmitted to the manometer meniscus. Besides, eyesight observations have been taken into consideration.

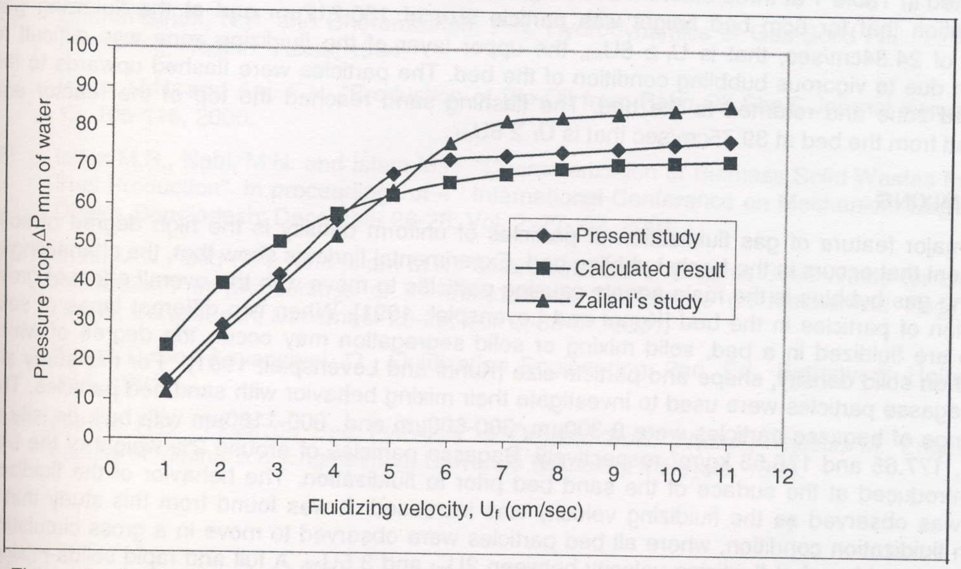

Figure 6. Comparison between the results obtained by cylindrical model and that of Zailani's rectangular model, and calculated results from mathematical model for bed height of 8cm and particle size of 150-212μm.

BED PARTICLE ENTRAINMENT AND ELUTRIATION

Generally, at the surface of a bubbling fluidized bed, where the fluidizing gas velocity is several times the minimum fluidizing velocity, large bubbles burst, throwing particles to the freeboard zone. These particles become entrained by the upward flowing gas stream. The freeboard zone should be of sufficient height to allow the entrained particles to disengage and return to the bed surface. At a sufficiently high gas flow rate, the terminal velocity of the solid is exceeded, the upper surface of the bed disappears, and entrainment becomes appreciable. Instead of bubbles, a turbulent motion of particles clusters and voids of gas are observed and this condition is termed as turbulent fluidized bed. With a further increase in gas velocity, particles are elutriated, i.e. carried out of the bed with the gas. The hydrodynamics of entrainment of particles in fluidized bed are very complex and are influenced by many parameters. Experimental observations have shown that the rate of entrainment of particles depend on the superficial gas velocity, particle density, particle size, gas density and the bed geometry such as cross-sectional area and height [Cheremisinoff and Cheremisinoff, 1984; and Kunni and Levenspiel, 1991]. Due to this complex flow phenomenon, extensive study on the entrainment of sand particles from the bed was not intended. The investigation was limited to the determination of the superficial gas velocity at which the splashing particles would reach the gas exit point from the reactor and be elutriated. The investigations were conducted on the selected particle sizes listed in Table 1 at three different bed heights of 4 cm, 6 cm and 8 cm. It was found from this investigation that for 6cm bed height with particle size of 150-212μm and at the fluidizing gas velocity of 24.84cm/sec, that is $U_f \geq 5U_{mf}$, the upper layer of the fluidizing zone was difficult to maintain due to vigorous bubbling condition of the bed. The particles were flashed upwards to the freeboard zone and returned to the bed. The flashing sand reached the top of the reactor and elutriated from the bed at 39.75cm/sec that is $U_f \geq 8U_{mf}$.

SOLID MIXING

A major feature of gas fluidization of particles of uniform density is the high degree of solid movement that occurs in the freely bubbling bed. Experimental findings show that, the channeling of the rising gas bubbles is the main agents causing particles to move with the overall effect of gross circulation of particles in the bed [Kunni and Levenspiel, 1991]. When two different types of solid particle are fluidized in a bed, solid mixing or solid segregation may occur, the degree of which depend on solid density, shape and particle size [Kunni and Levenspiel, 1991]. For this study air-dried bagasse particles were used to investigate their mixing behavior with sand bed particles. The size range of bagasse particles were 0-300μm, 300-600μm and 600-1180μm with bulk density of 145.91, 177.65 and 176.53 kg/m^3 respectively. Bagasse particles of around 2% weight of the bed were introduced at the surface of the sand bed prior to fluidization. The behavior of the fluidized solids was observed as the fluidizing velocity was increased. It was found from this study that a smooth fluidization condition, where all bed particles were observed to move in a gross circulating motion was achieved at fluidizing velocity between $2U_{mf}$ and $3.5U_{mf}$. A full and rapid solids mixing, with no apparent solid segregation between sand and bagasse particles were observed at this fluidizing velocity range. However, bagasse particles started to be elutriated from the bed at 13.80cm/sec that is about $U_f \geq 2.8U_{mf}$. The mixing ability of 0-300μm bagasse particle was very poor but they entrained and elutriated very easily from the bed.

CONCLUSIONS

- From this experimental investigation it is found that a good fluidizing behavior and solid mixing can be achieved at the gas velocity in between 9.53cm/sec to 17.32cm/sec for sand particle size of 150-212μm and 212-300μm with bed height of 6cm.
- The elutriation of bagasse particles and sand particles are possible at 13.80cm/sec and 39.75cm/sec respectively.

- Fine sand particles of size range 150-212μm and 212-300μm are suitable to be used in the fluidized bed reactor, since bigger particles require higher gas flow rate to achieve smooth fluidization conditions.
- The distributor plate of 9 nozzles is suitable and the selected bed height is of 4cm or 6cm.
- Feed particle size of 300-600μm shows good mixing behavior and may be used for fluidized bed system.
- Cylindrical reactor is better one than the rectangular reactor in the fast pyrolysis system.

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