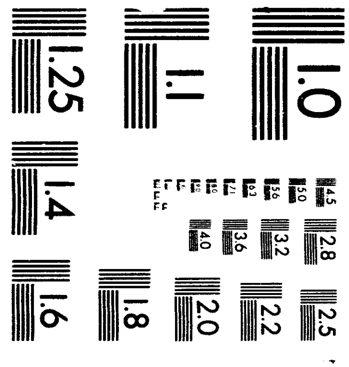

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EFFECTS OF PARTICLE PROPERTIES ON FLUIDIZATION CHARACTERISTICS OF COARSE PARTICLES

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INTRODUCTION

Fluidized-bed reactors can be operated in different flow regimes. Since the gas-solid contacting modes vary as the fluidization regimes change, results obtained from one flow regime for a particulate material can not be applied to another (Grace, 1986). Ascertaining flow regimes of operation is, therefore, essential for design and scaleup of fluidized-bed reactors as well as for reactor performance predictions. During the past decades, considerable research efforts have been spent on flow regime studies for fine particles (Geldart's group A and group B powder). Most of the available data has been obtained with small laboratory-scale units of 15 cm diameter or less. Most fossil energy fluidized bed combustors have used coarse particles as bed material for economic reasons. For example, both bubbling and circulating fluidized bed combustion systems have used coarse coal and limestone particles (Abdulally and Parham, 1989; Fraser, Schaller, and Darguzas, 1991; Boemer, Braun, and Renz, 1993). Despite of all these research efforts, there remain many areas where our understanding of gas-solid fluidization is still far from complete (Grace, 1986). In particular, little is known of the transition regimes between bubbling fluidization and dilute phase flow for fine (Yerushalmi, Cankurt, Geldart, and Liss, 1978; Grace, 1986; and Horio, 1992) and large/heavy particles (Horio, 1990). It is the objective of this study to better understand the various flow regimes and their transitions from incipient fluidization to high velocity fluidization using coarse particles in a 30 cm diameter atmospheric fluidized bed cold model. Results of this investigation on high velocity fluidization process will be helpful in predicting operational characteristics and improving circulating fluidized-bed reactor design and scale up for fossil energy applications.

EXPERIMENTAL

A schematic of the experimental apparatus is shown in Figure 1. It consists of a 0.3 m diameter, 6 m high fluidization vessel with metered air supply; a primary and a secondary cyclone and a metal cloth filter. The fluidization vessel is instrumented with four wall mounted capacitance plates and six sets of differential pressure transducers. The capacitance plates are located at 25 cm, 45 cm, 81 cm, and 100 cm above the air distributor. Differential pressure tappings are located at 10 cm intervals along the vertical axis. Solids captured in the primary cyclone are either weighed to determine the amount carried out of the fluidization vessel or recycled back to the fluidization vessel through an eductor and a non-mechanical "L" valve.

Differential pressure and capacitance signals indicating changes in bed pressure and density were used to characterize the transitions of gas-solid fluidization process as gas velocity eas varied. Both these signals were sampled at 125 Hz. Data reduction and analysis of these signals were performed in the same manner as described previously (Rockey, Mei, Nakaishi, and Robey, 1989; Mei, Rockey, Lawson, and Robey, 1991).

The properties of the coarse particles tested throughout the investigation are given in Table 1. The static bed height was held constant at 46 cm for all the experiments.

RESULTS AND DISCUSSION

Since both the void fraction as well as the standard deviation of bed differential pressure fluctuations have been used to characterize the flow regimes and their transitions, it is interesting to compare the data obtained from both instruments. In Figure 2, both the average void fraction and the standard deviation of bed differential pressure fluctuations are plotted for the 3.18 mm nylon spheres. The pressure fluctuation data were obtained from a pressure transmitter which was located at about the same height (between 38 to 48 cm) as the capacitance plate (at 45 cm). The transition velocity, U_c, which is defined as the gas velocity at which the bed differential pressure fluctuations start to decay from a maximum can be identified at U = 3.00 m/sec from the pressure fluctuation data. The void fraction trace, however, does not provide any clues to identify such transition velocity. As the gas velocity increases, the bed differential pressure fluctuations start to decay as more solids are transferred to the freeboard. The decay in pressure fluctuations is primarily due to the holdup of more solids in the freeboard which is clearly indicated by the increase in void fraction. As the gas velocity reaches 7.71 m/sec, the bed differential pressure fluctuations have decayed to a minimum value and start to level off. The corresponding void fraction is 0.99 which suggests that most of the solids have been transferred to and are held in the freeboard. At the transition velocity, U_k = 7.71 m/sec, the bed height approaches zero. Therefore, the transition velocity, U_k, does not represent the onset of the so-called "turbulent" flow regime. Instead, it signals the onset of refluxing pneumatic transport regime (Rhodes and Geldart, 1986) which is clearly indicated from the increase in void fraction measured with the capacitance probe as well as the decrease in amplitude of bed differential pressure fluctuations obtained from the pressure transducer. The transition gas velocity, U_k, as defined by Rhodes and

Geldart (1986), represents the transfer of solids from bed to freeboard in the case with no net through-flux. For the case with imposed solid flux, this gas velocity increases with solid flux and represents the minimum velocity at which the gas can just entrain all the solids fed into the fluidized bed.

Effects of particle properties on fluidization characteristics of coarse particles of different size, shape, and density are shown in Figures 3, 4, and 5. In Figure 3, bed differential pressure fluctuations of two different particle sizes, the 3.18 and 7.14 mm acrylic chips, are plotted against the gas velocity. As expected, the minimum fluidization velocity and both transition velocities, U_c and U_k, increase with particle size. Furthermore, the increase in standard deviation for the larger particle (7.14 mm Acrylic chips) reflects the larger amplitude of the bed differential pressure fluctuations for larger particles. However, it is of interest to note that the standard deviation of bed differential pressure fluctuation profiles for both small and large particles with the same density and sphericity are similar in shape.

Figure 4 shows the effect of different particle shapes, as indicated by sphericity (0.99 for the nylon spheres, 0.87 for the nylon cylinders, and 0.60 for the acrylic chips), on flow regime transitions. It can be seen from this figure that the transition velocity to refluxing pneumatic transport, U_k, increases with the sphericity. However, both the minimum slugging velocity and the transition velocity U_c are only slightly affected by the particle sphericity. However, the transition velocity to the refluxing pneumatic transport, U_k, increases considerably with particle sphericity. This suggests that a much higher gas velocity is required in order to transport particles with higher sphericity. Among the 3.18 mm particles, the nylon cylinder which has sphericity of 0.87 causes the highest bed differential pressure fluctuations.

Figure 5 shows the effect of particle density on the flow regime transitions. As expected, the transition gas velocity and that the amplitude of differential pressure fluctuations increase with particle density.

Pneumatic transport of coarse particles was established and maintained at various solid recycling rates. Figure 6a, 6b, and 6c plot the pressure gradient against the recirculating solids flux with gas velocity as the parameter for the nylon cylinders, nylon spheres, and the acrylic chips respectively at gas velocity beyond the transition velocity, U_k. Figure 6a presents the data for the 3.18 x 3.18 mm nylon cylinders. At gas velocity of 7.2 m/sec, the void fractions measured at 45 cm above the gas distributor range from a high of 0.95 to a low of 0.7 corresponding to solid flux of 3.5 and 14.0 $kg/m^2/sec$ respectively. These data suggest that a dense bed exists at the bottom of the riser provided that sufficient solids can be recycled. Data for the 3.18 mm coarse particles given in Figure 6a, 6b, and 6c are re-plotted in a phase diagram as shown in Figure 7. Karri and Knowlton (1990) defined the region between the choking velocity (the velocity at which the gas is saturated with solids) and the gas velocity at the minimum point of the pressure gradient as the fast fluidization regime for a given solid mass flux. Since the transition gas velocity, U_k, is similar to the choking velocity and the gas velocity has not yet reached the minimum value of the pressure gradient, the experimental data for the 3.18 mm coarse particles, therefore, fall into the region of fast fluidization in the phase diagram according to such practical definition given by Karri and Knowlton (1990). Figure 7 also shows the influence of the particle sphericity on the fast fluidization regime. For the same solid recirculating rate, higher gas velocity is required to establish the fast fluidization regime for particles with increasing sphericity.

CONCLUSIONS

1. The so-called "turbulent" flow regime was not experienced with the tested coarse particles. Instead, when the bed differential pressure fluctuations decayed and leveled off at a gas velocity, U_k, most of the bed material had been transferred to the freeboard or in the solids separation and return system. When U_k was reached, the condition in the fluidization vessel was equivalent to pneumatic transport at low velocities with refluxing of solids.

2. The transition to the onset of refluxing pneumatic transport for the coarse particles, characterized by the gas velocity, U_k, is signalled by a decrease in amplitude of bed differential pressure fluctuations and an increase in void fraction.

3. The transition gas velocity for the tested coarse particles, U_k, increases with increasing particle size, sphericity, and density.

4. The phase diagram for the 3.18 mm particles suggests that fast fluidization flow regime can be established and maintained for coarse particles.

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TABLE I

Physical Properties of Tested Solids in the Experiments
and the Transition Gas Velocities

Solid	Particle Size mm	Particle Density gm/cm^3	Particle Sphericity	Minimum Slugging Velocity m/sec	Transit. Velocity U_k m/sec
Nylon Spheres	3.18 7.14	1.10 1.10	0.99 0.99	1.15 1.90	7.71 ----
Nylon Cylinders	3.2x3.2 4.76x6.4	1.10 1.10	0.87 0.86	1.1 1.35	6.85 8.95
Acrylic Chips	3.18 7.17	1.19 1.19	0.60 0.60	0.84 1.28	5.57 6.92
Ceramic Spheres	3.18 7.14	2.40 2.40	0.99 0.99	1.46 2.50	---- ----
Limestone Chips	3.18 7.14	2.6 2.6	0.60 0.60	1.17 2.15	9.04 ----

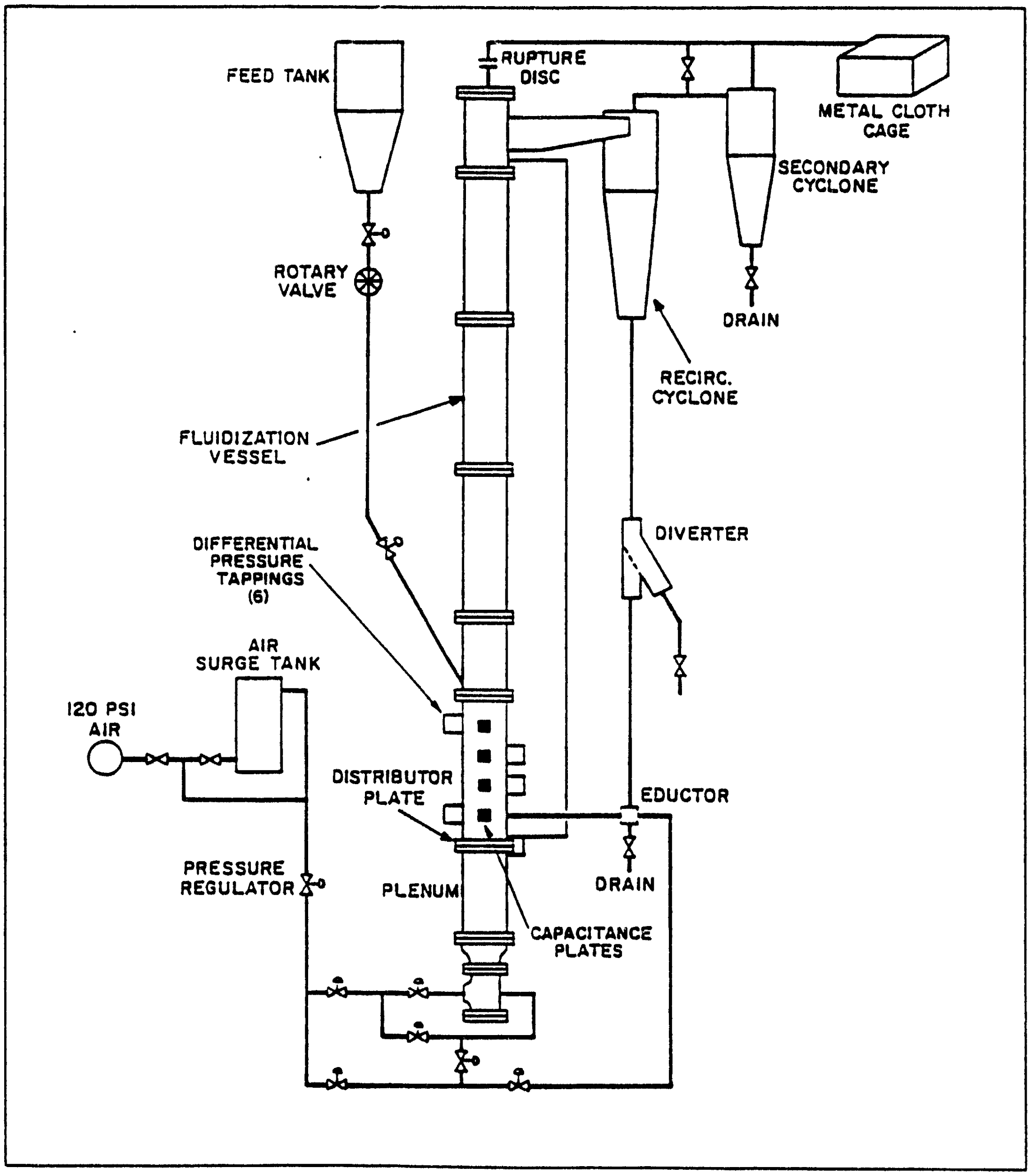

Fig.1 Schematic Diagram for the
30 cm Diameter Fluidization Facility

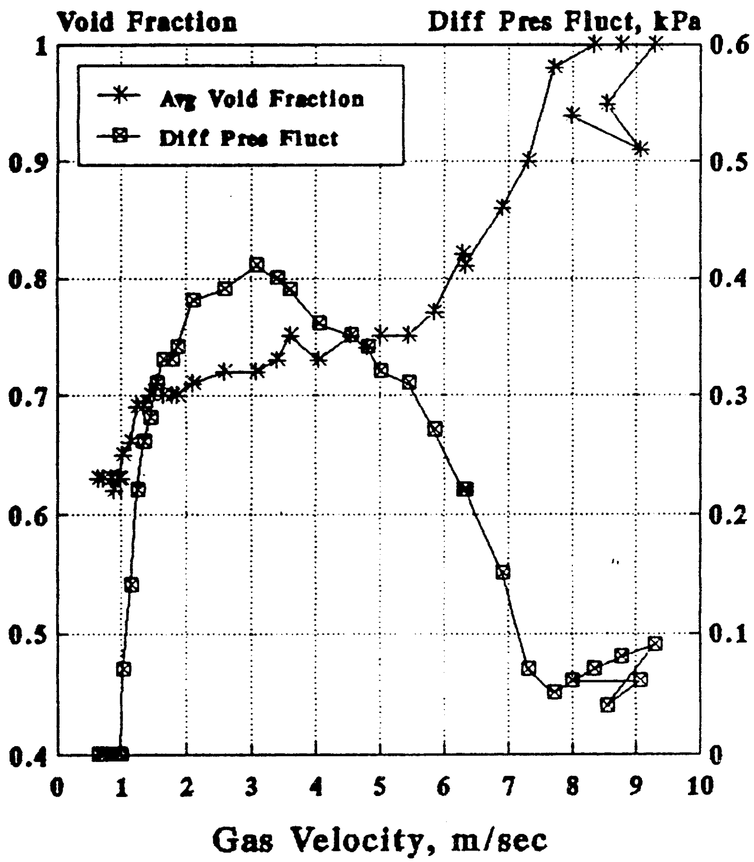

Figure 2. Void fraction and standard deviation of bed differential pressure fluctuations versus gas velocity

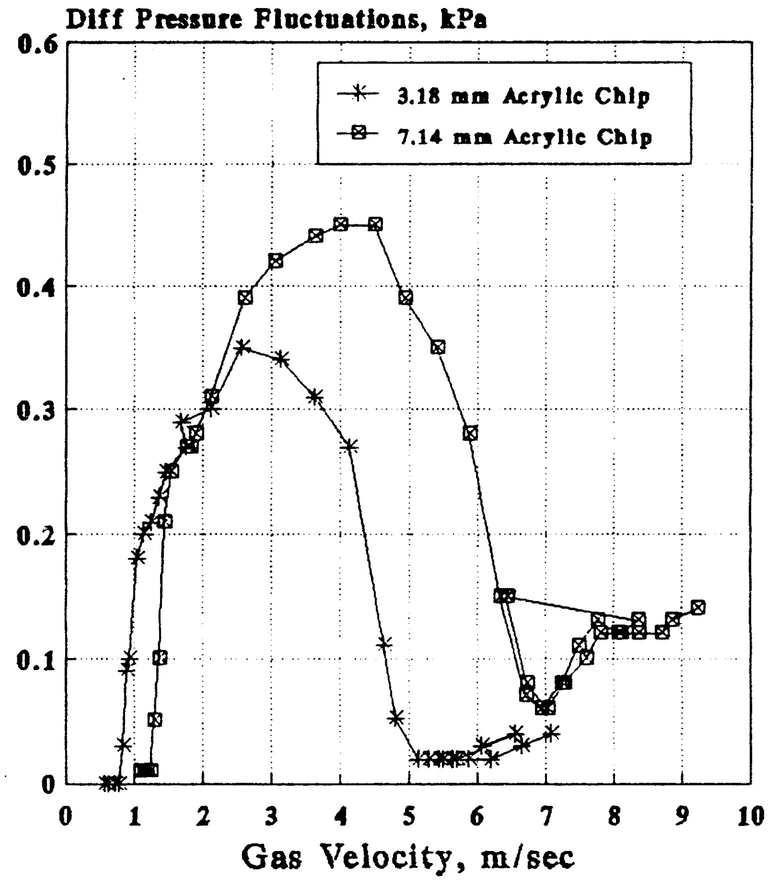

Figure 3. Effect of particle size on flow regime transitions

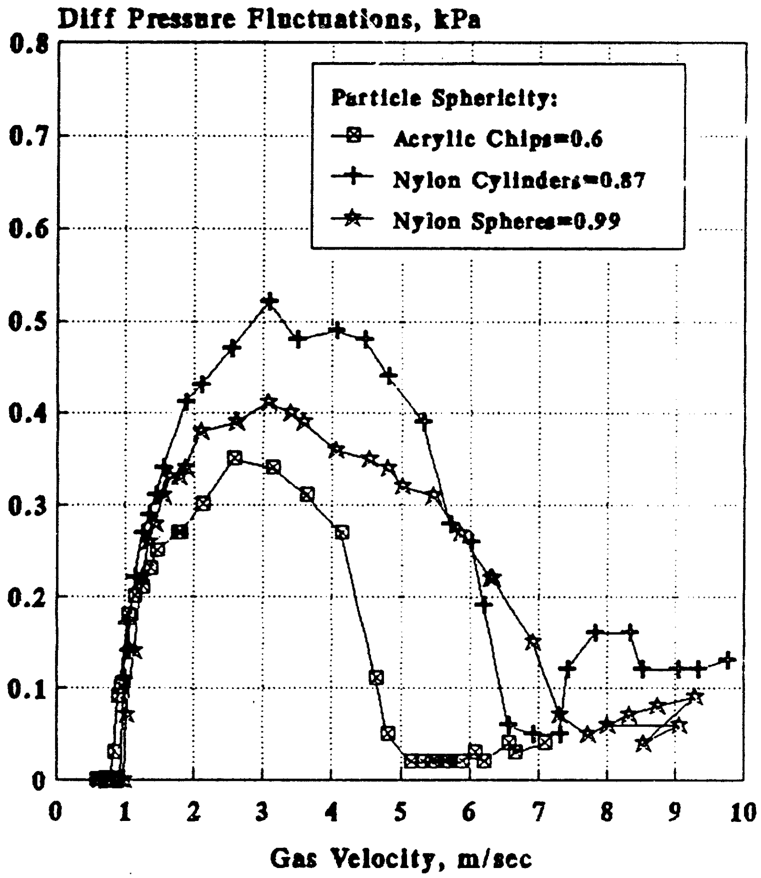

Figure 4. Effect of particle sphericity on flow regime transitions

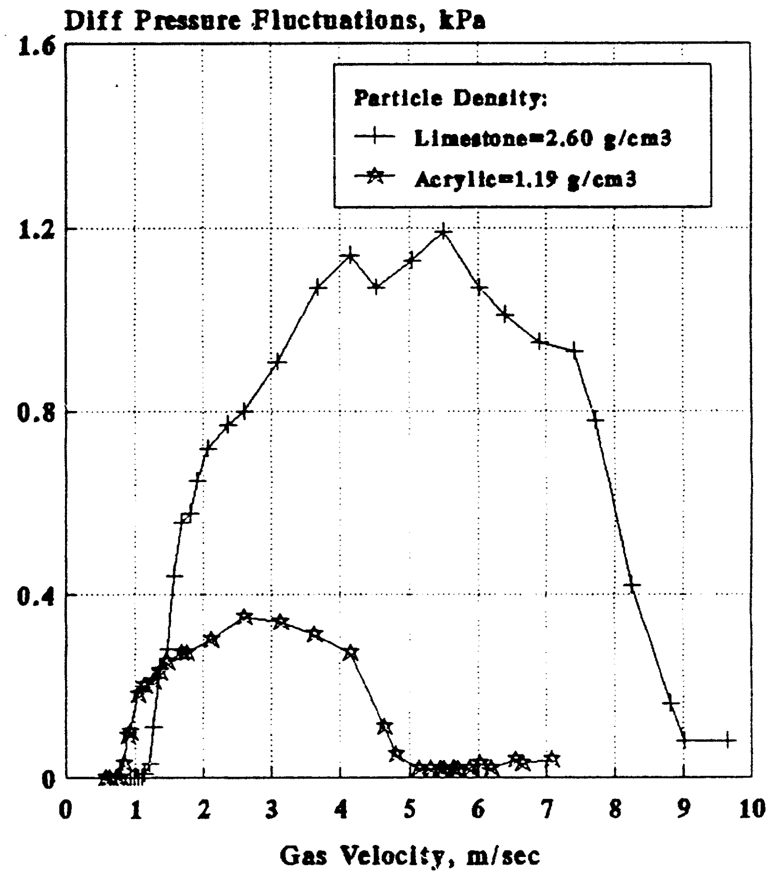

Figure 5. Effect of particle density on flow regime transitions

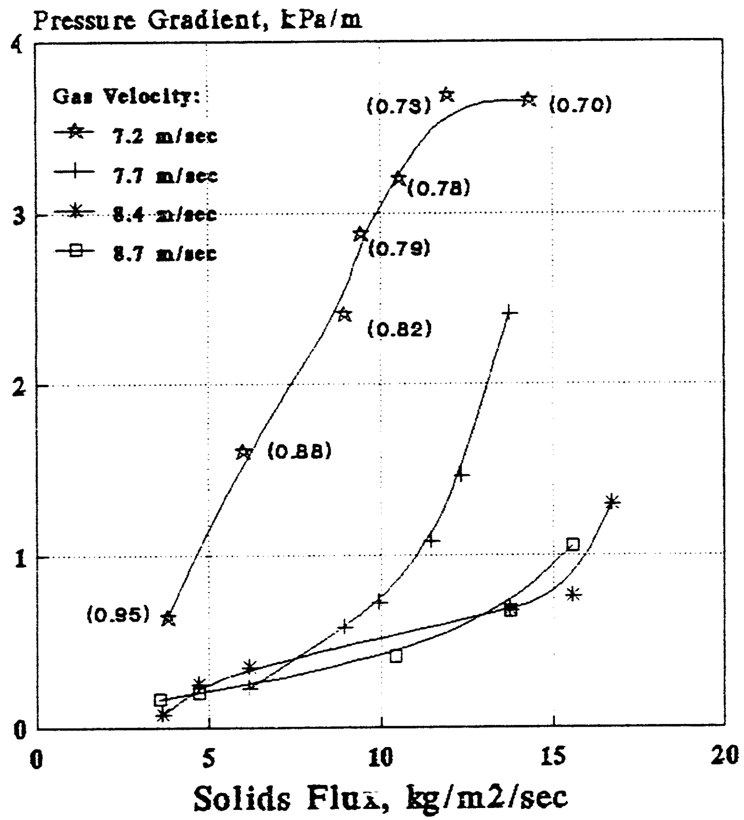

Figure 6a. Pressure gradient measure between 0.38 m and 0.48 m above gas distributor for 3.18 mm nylon cylinders

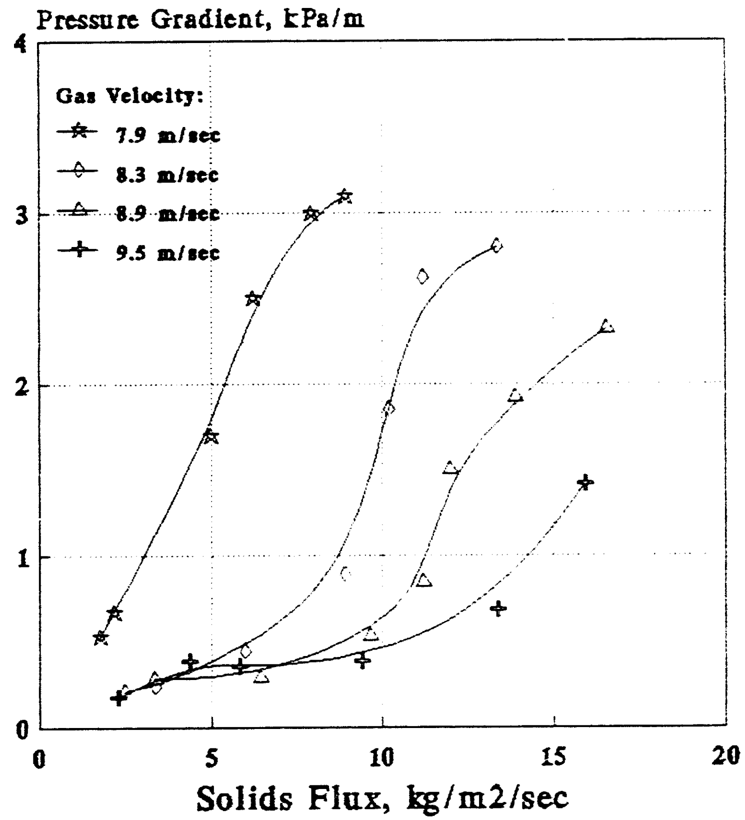

Figure 6b. Pressure gradient measured between 0.38 m and 0.48 m above gas distributor for 3.18 mm nylon spheres

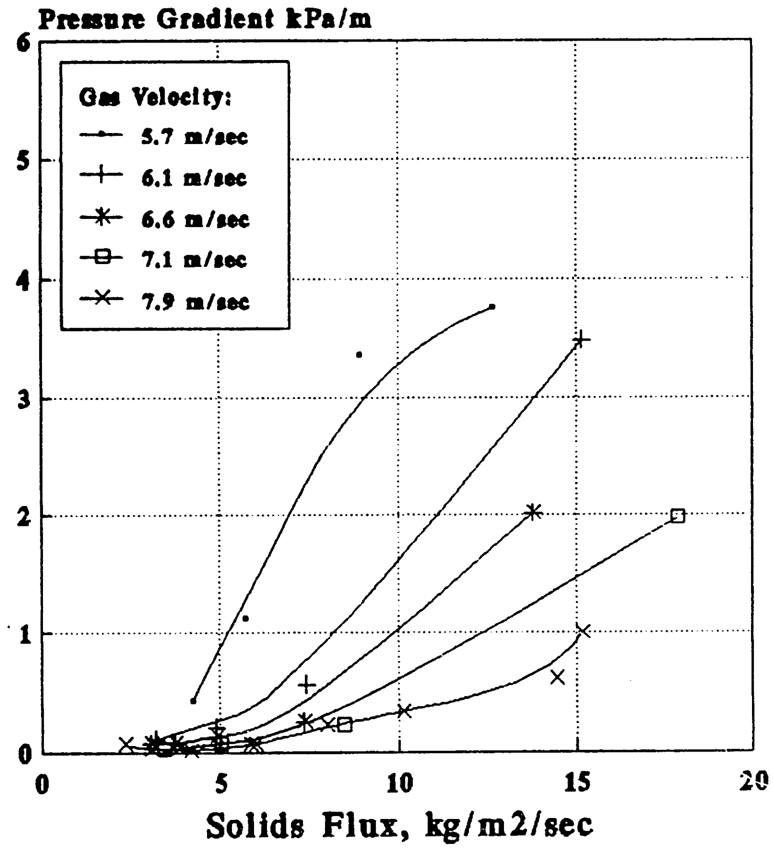

Figure 6c. Pressure gradient measured between 0.38 m and 0.48 m above gas distributor for 3.18 mm acrylic chips

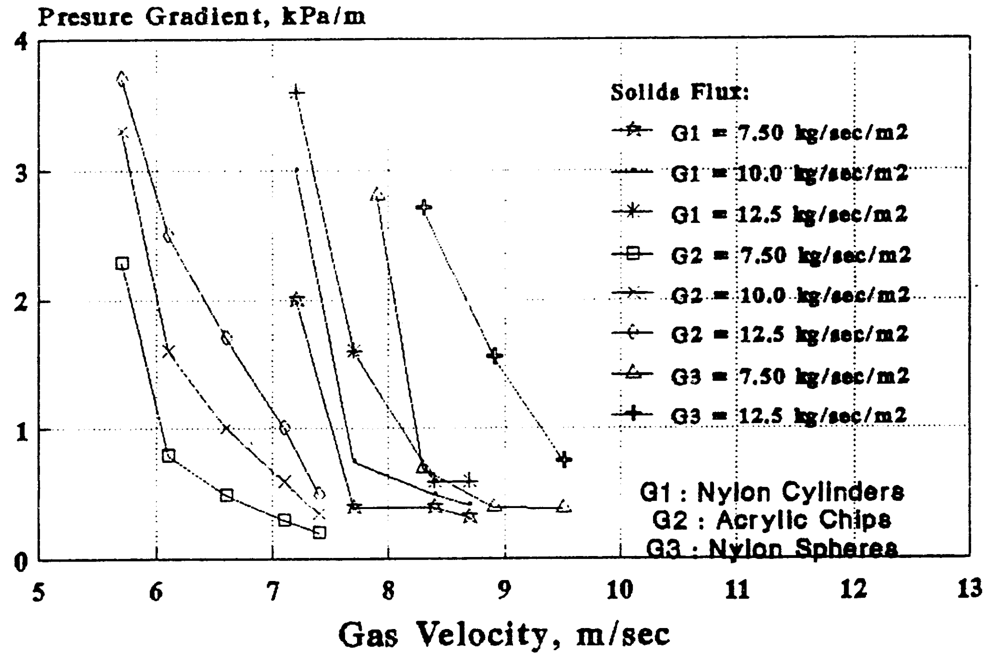

Figure 7. Phase diagram for 3.18 mm particles with different sphericities

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