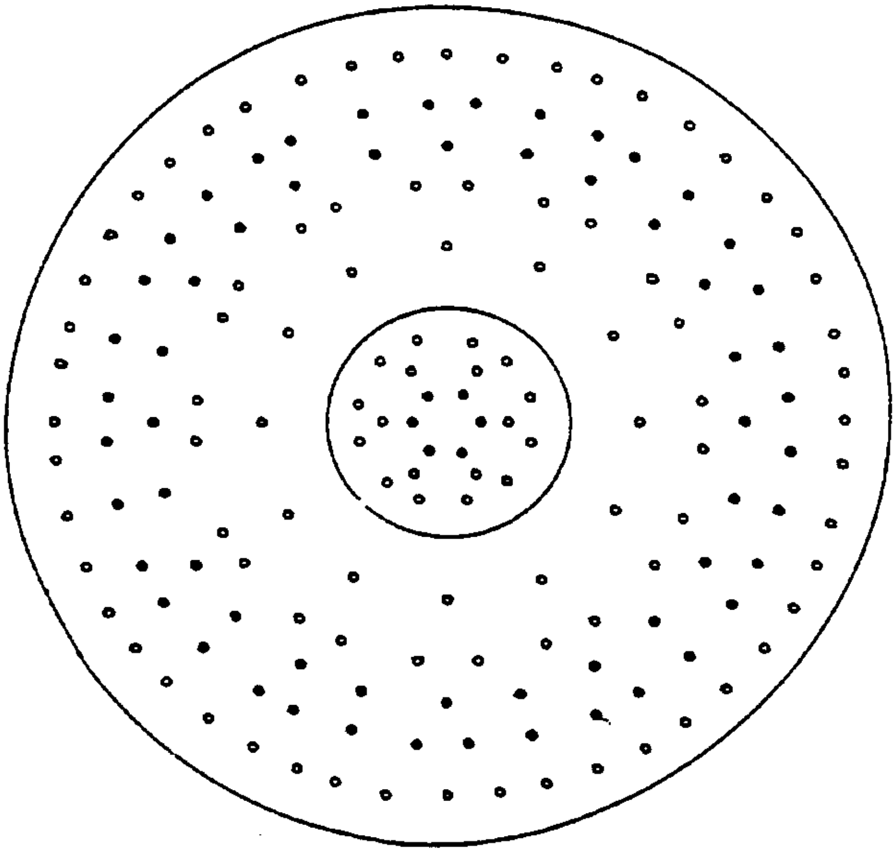

Figure 22. The design of the lower distributor plate.

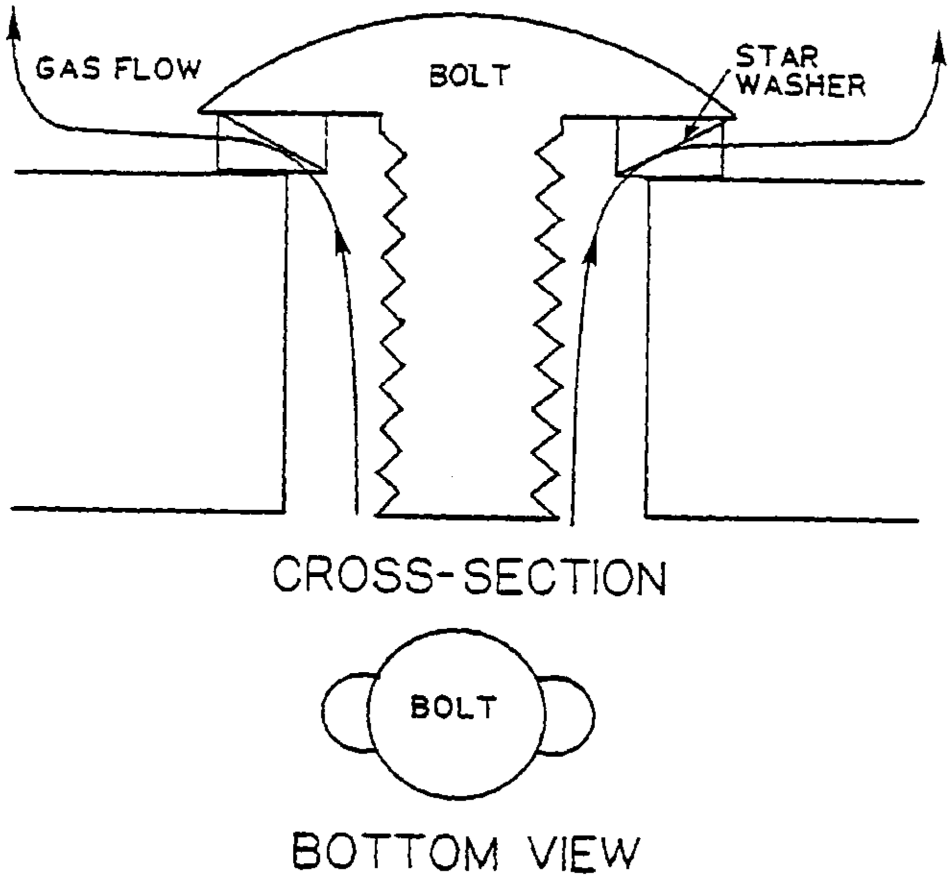

Figure 23. The details of the construction of the orifices in the distributor plate.

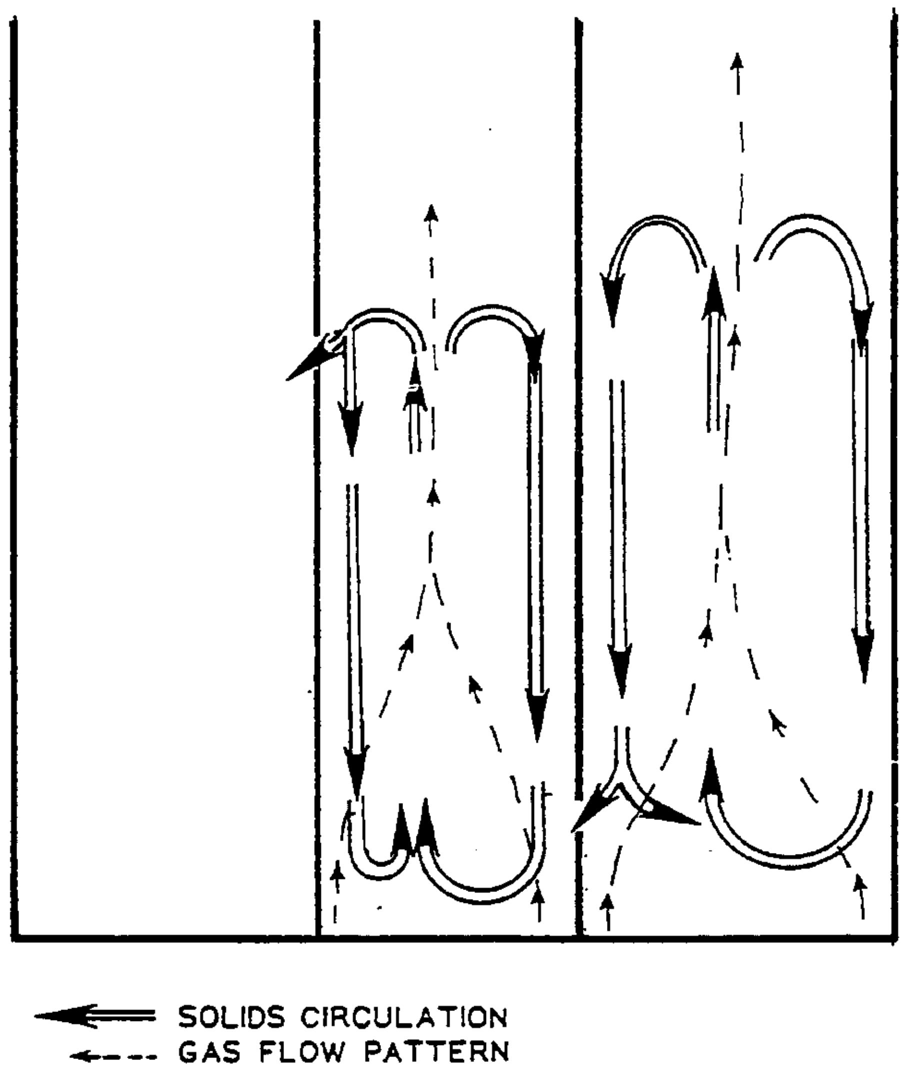

Figure 24. The gas and solids flow patterns expected.

reactor. An 1.27 cm (1/2 in) hole located at the bottom of the combustor was used to remove the solids from the unit when desired. The hole was located approximately where the entrance for the feed in the actual reactor would be located. The top of the combustor was attached to the pyrolyzer. The point of attachment was reinforced with rings of 0.64 cm (1/4 in.) plexiglass. Two 1.91 cm (3/4 in) outlets were provided for the gases exiting the combustor. It was found to be essential for the pyrolyzer and the combustor be maintained at the same pressure in order to prevent crossflow of gases and an uneven circulation of solids. The large outlet holes prevent an excessive pressure drop from occurring in the combustor. Solid transfer tubes were attached to the pyrolyzer over the holes provided for solids circulation. The tubes promote the desired circulation pattern while at the same time making the cross flow of gas more difficult. The combustor is 61.0 cm (24 in.) in diameter while the pyrolyzer is 24.1 cm (9.5 in.) in diameter. The upper transfer hole is located 40.6 cm (16 in) from the bottom of the reactor. The total height of the stage is 61.5 cm (24 in.)

The scrubber entrance introduces the gas from the pyrolysis reactor into the scrubber near the scrubber's wall and directed in such a way as to induce cyclonic flow.

In this way the scrubber makes use of the inertial effects found in a cyclone and the collision effects found in ordinary scrubbers to remove particles from the gas stream. Care must be taken in the design of the entrance to insure that it induces as low a pressure drop as possible. Two layers of water jets are included in the design. These jets were not used in the simulator studies. The water inlet was used to withdraw or add gas in order to simulate both the cooling of the gas and the recycling of Fischer-Tropsch outlet gases.

The scrubber opens directly on the distributor plate for the Fischer-Tropsch stage. The distributor plate is made of 0.32 cm (1/8 in) aluminum sheet just like the lower distributor. The pattern of holes in the Fischer-Tropsch distributor is the same as in the inner section of the pyrolyzer/combustor distributor. The distributor plate limited the Fischer-Tropsch stage to fluidized beds. The Fischer-Tropsch reactor is made of three sections each of which is approximately 152 cm (5 ft.) long. The stage can thus be between 152 cm (5 ft.) and 457 cm (15 ft.) tall. It was found that the best flow characteristics were obtained using a 300 cm (10 ft.) tall reactor. The reactor was 16.5 cm (6.5 in.) in diameter.

Compressed air was used to simulate the gas flow for all reactor stages. The air was regulated to 239 kPa

(20 psig). This pressure is well below the maximum of 928 kPa (120 psig) predicted by considering the yield strength of the plexiglass however the glued joints tended to fail when 239 kPa was approached. The gas flow rates were controlled using Brooks rotameters. Each rotameter was scaled to read a maximum of 11.33 l/s (24 scfm).

Simulator Experiments

The experiments with the simulator had two goals. The first was to examine the operating characteristics of the reactor in order to determine the feasibility of the design used. In addition, any oversights in the original design would be located and corrected. The experiments involved in attaining this goal were qualitative with the results being used to modify the simulator system.

The second purpose was to examine those parameters describing the behavior of the bed that could be tested without extensive instrumentation. The effects of varying bed height and gas flow rate on bed expansion, average visible bubble rise velocity, minimum fluidization velocity, critical gas velocity, bubble distribution at the exit of bed exit, height of jetting region, and solids circulation were determined. Table 5 shows the experiments carried out in the simulator.

Table 5. The experiments performed in the simulator.

Independant Variable	Dependant Variable
Initial gas flow rate	Visible bubble velocity Expanded bed height
Bed height	Visible bubble velocity Expanded bed height Minimum fluidization velocity Bubble distribution Tracer concentration Gas velocity required to fluidize the entire bed
Run time	Tracer concentration
Reactor	Height of the jetting region

Two sets of experiments were conducted using the plexiglass simulator. The first set was conducted on the original reactor prior to its modifications and the second was conducted using the modified reactor shown in Figure 18. The only operational characteristic that varied significantly between the two sets of experiments was the circulation of the solids. In the original design there was very little transfer from one reactor to the other. In the modified version there was considerable transfer.

One of the most important items to be considered when conducting experiments with fluidized beds is the nature of the particles to be fluidized. The particles used in the simulator studies were alumina. The bulk density of alumina is 1.60 g/cm^3 while the apparent density of the alumina is 2.60 g/cm^3. (Hsu, 1979) The particles were examined under a microscope at a magnification of 2x. Since the average particle size used was 0.025 cm, this magnification was sufficient to allow a detailed analysis of the particles shape but was not high enough to provide any surface detail. The microscopic examination demonstrated that the alumina was made up of two different types of particles. One type of particle was dark and nearly spherical. These particles had a sphericity of 0.67. The other type of particle was translucent and resembled flat plates. They had a sphericity of 0.14.

RESULTS AND DISCUSSION

SIMULATOR EXPERIMENTS

The results of the simulator studies were largely as expected. The heights of the jetting regions were all around 4 cm. This value is only accurate to within 1 cm due to the difficulty in determing the boundry of the jetting region. The experimentally determined values for the minimum fluidization velocity in each of the three reactors are shown in Figure 25. The velocities for the pyrolyzer and the Fischer-Tropsch reactor are independent of bed height. The values for the minimum fluidization velocity can be predicted using several different correlations available in the literature. The expression developed by Leva (Hsu, 1979) was the most accurate for the the pyrolyzer and the Fischer-Tropsch reactor.

The behavior of the combustor was considerably different and somewhat unexpected. The velocity required to fluidize the bed increased linearly with bed height. Repeated experiments verified the trend. One likely explanation for this phenomena stems from the fact that the bed in the combustor has a much higher surface area to volume ratio than either of the other two beds. This means

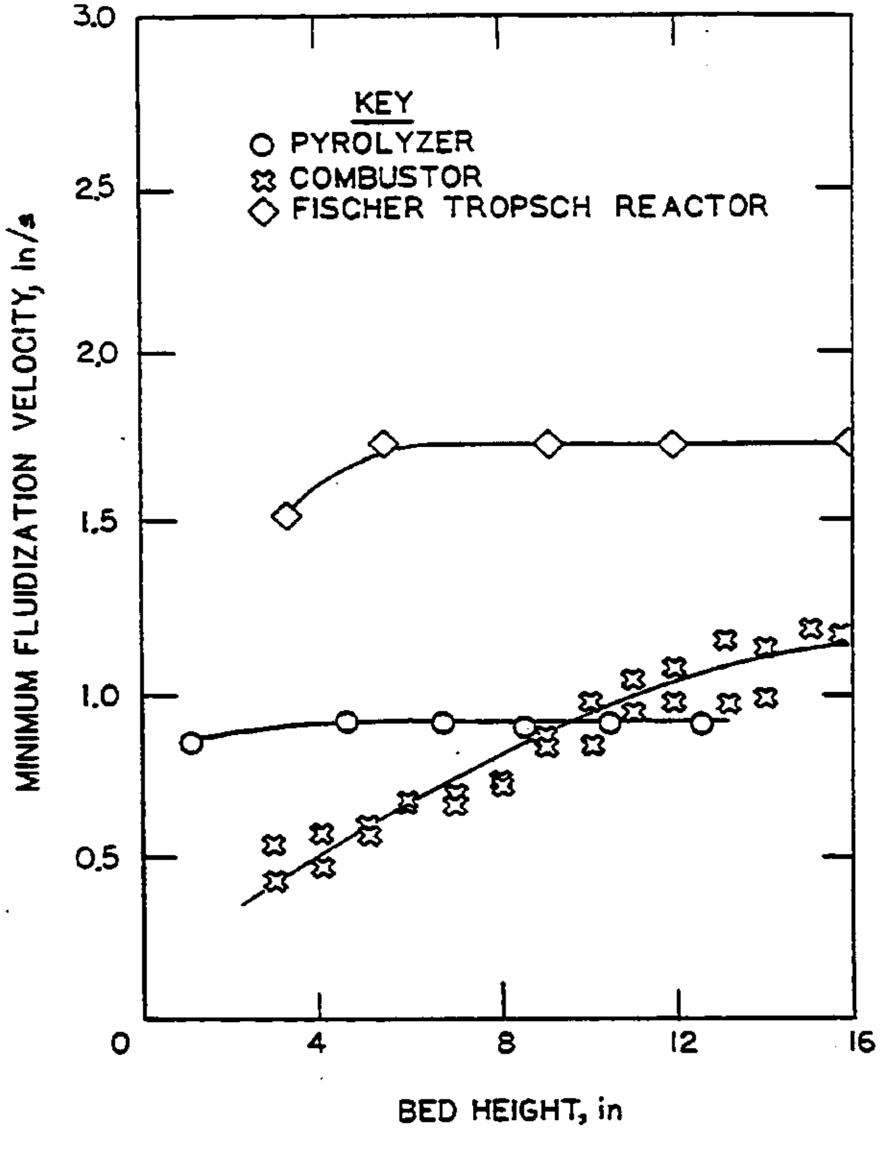

Figure 25. The minimum gas flow rate required to fluidize each of the three reactors at various bed heights.

that the effects that occur at the walls of the reactor have a much greater effect on the behavior of the combustor than on the other beds. The plexiglass tends to build and hold static electricity when moving particles slide along its surface. This static will tend to stabilize the bed and prevent it from fluidizing. The resisting force would increase with the height of the bed.

The combustor also exhibited another unexpected characteristic. Instead of moving directly from fixed bed behavior to fluidized bed behavior, one section of the bed would fluidize first. As the gas flow rate was increased, the fluidized area would expand. Eventually the velocity would reach a critical value where the entire bed was fluidized. The gas velocity could then be decreased while the bed remained fluidized.

This behavior is not unprecedented. Baskakov, Tuponogov, and Philipposky (1985) found this phenomena in several large fluidized beds. The cause is most likely the nonuniformity of the bed. The packing density varies minutely from spot to spot in the bed. If there is one path up through the bed that is slightly easier than any of the others, the gas will take it. The resulting gas flow pattern will fluidize one part of the bed before the rest. The critical velocity for the combustor was found to be 6.1 cm/s. This value was independent of the height of the bed and the minimum fluidization velocity. The fact that the critical velocity is constant while the minimum fluidization velocity increases is consistent with the explanation put forth for the variation in the minimum fluidization velocity. By the time the majority of the bed is fluidized, the force exerted by the gas would be so much larger than the force exerted by the static charge that the charge would have no significant effect.

The distribution of bubbles in the bed at different heights was investigated in the pyrolyzer. A guide marked so as to delineate three concentric regions of the bed without interfering with the bed was constructed and placed over the pyrolyzer. The number of bubbles bursting in each of the areas was observed at various bed heights. Figure 26 shows the distribution of bubbles between the three regions as a function of bed height. The middle third of the bed exhibited the highest number of bubbles at any given height above 3 in. and the inner third showed the fewest. However, if the area of each region is accounted for, the inner third of the bed and the outer trade relative positions as is shown in Figure 27. In any case, the fact that the middle third of the bed exhibited the largest number of bubbles demonstrated that the tendency of the bubbles to cling to the walls had been overcome .

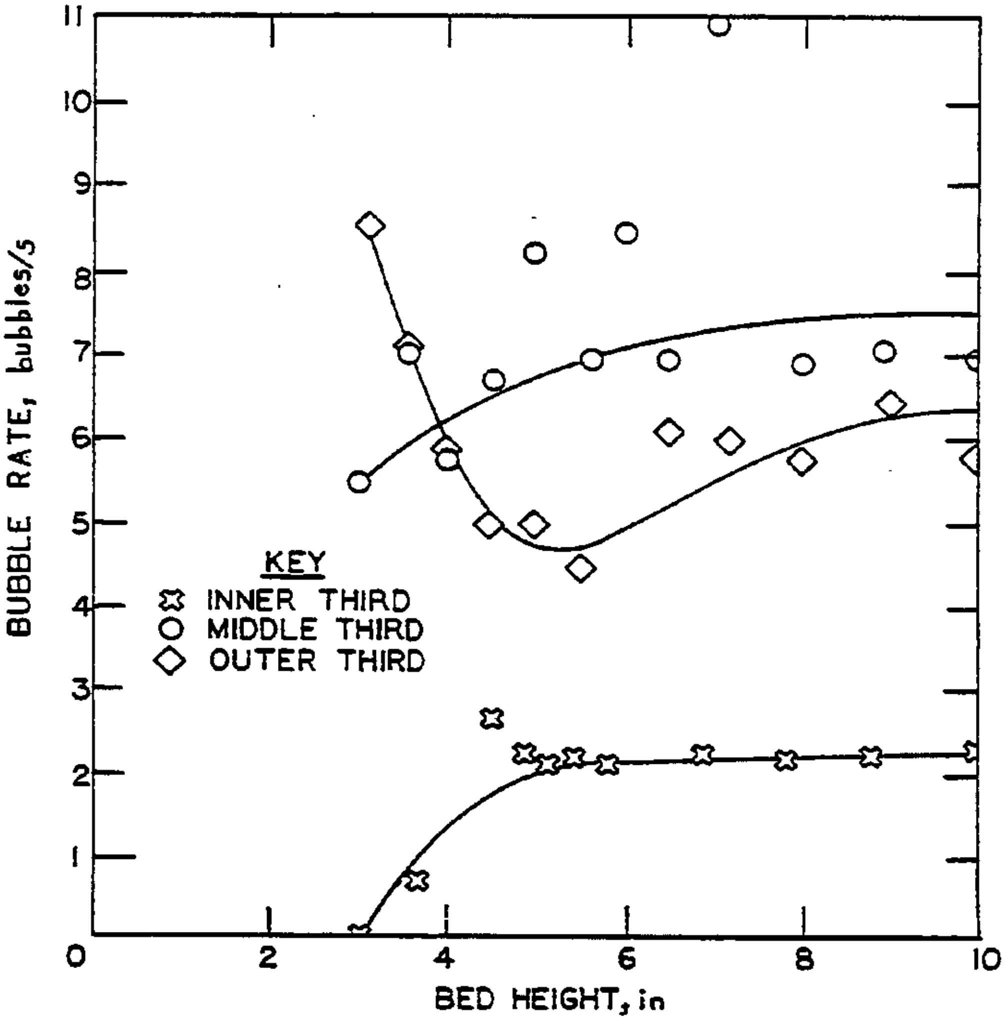

Figure 26. The rate bubbles appeared in each of three concentric areas of the bed at different heights.

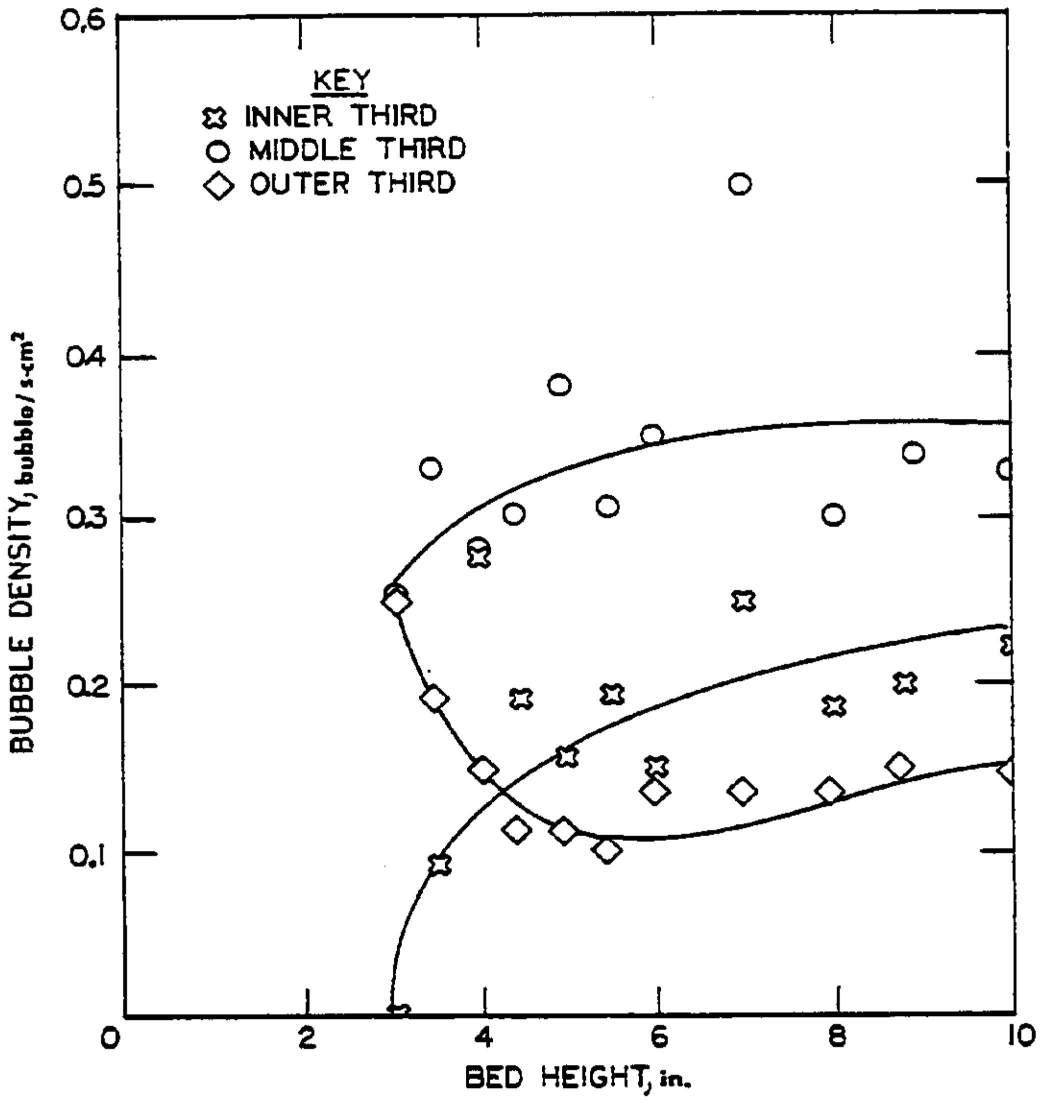

Figure 27. The density of bubbles appearing in each of three concentric areas of the bed.

The circulation of the solids was measured before and after the reactor modifications were made. Figure 28 shows the concentration of the tracer particles at various heights in both the combustor and the pyrolyzer while Figure 29 shows the concentration at various times for the initial configuration. The concentration does not vary significantly with height even after very short time periods. However there appears to be little transfer between the pyrolyzer and the combustor. Figure 30 shows the concentration of tracer in the reactor after the modifications were made.

The final characteristic of the beds that could be measured was the visible bubble rise velocity. Previous studies have shown that this velocity is slower than the gas velocity. (Sit and Grace, 1981) However, it is indicative of the rate at which the gas is moving through the bubble phase. The rise velocity was measured at different overall gas flow rates and different bed heights. Figure 31 shows the bubble rise velocity at various bed heights. Figure 32 shows the effect of the gas flow rate on the bubble velocity. The correlation used in the models to predict the bubble velocity was developed by Davidson and Harrison. (1982) A comparison of the results of the simulator studies with the predictions of the correlation is also shown in Figure 31. the correlation is very good

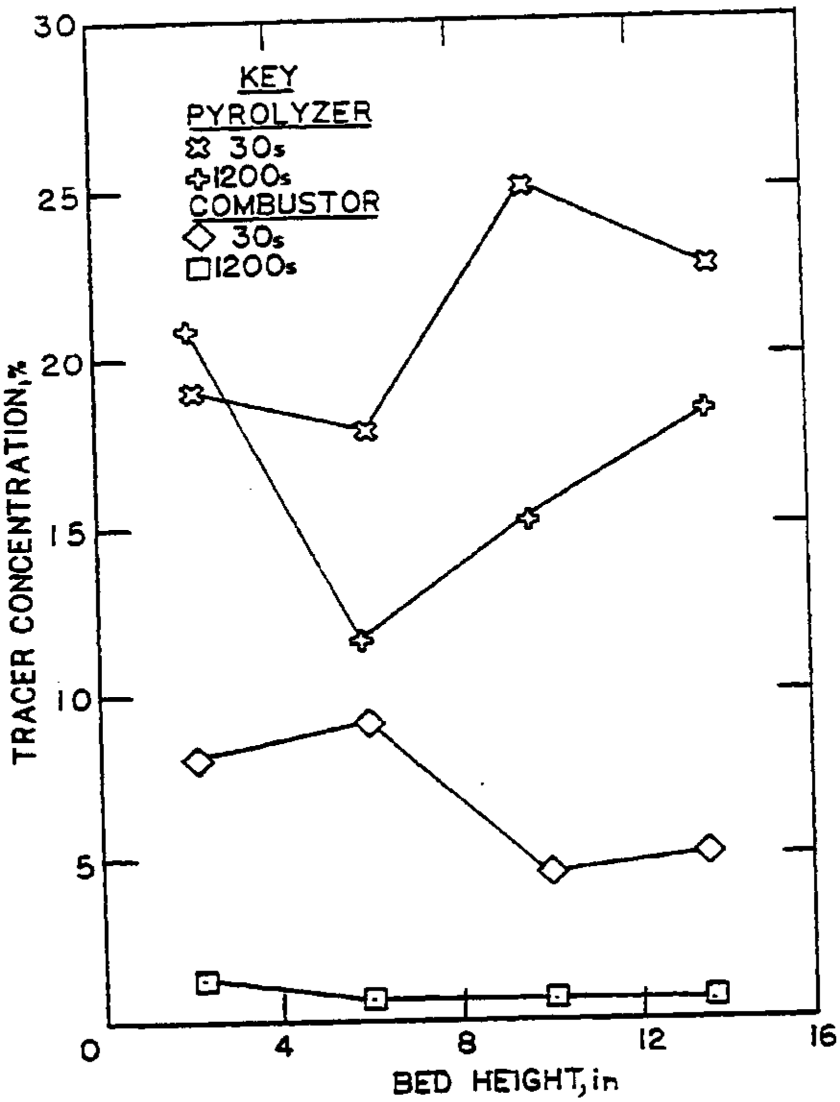

Figure 28. The concentration of tracer particles in number percent as a function of bed height.

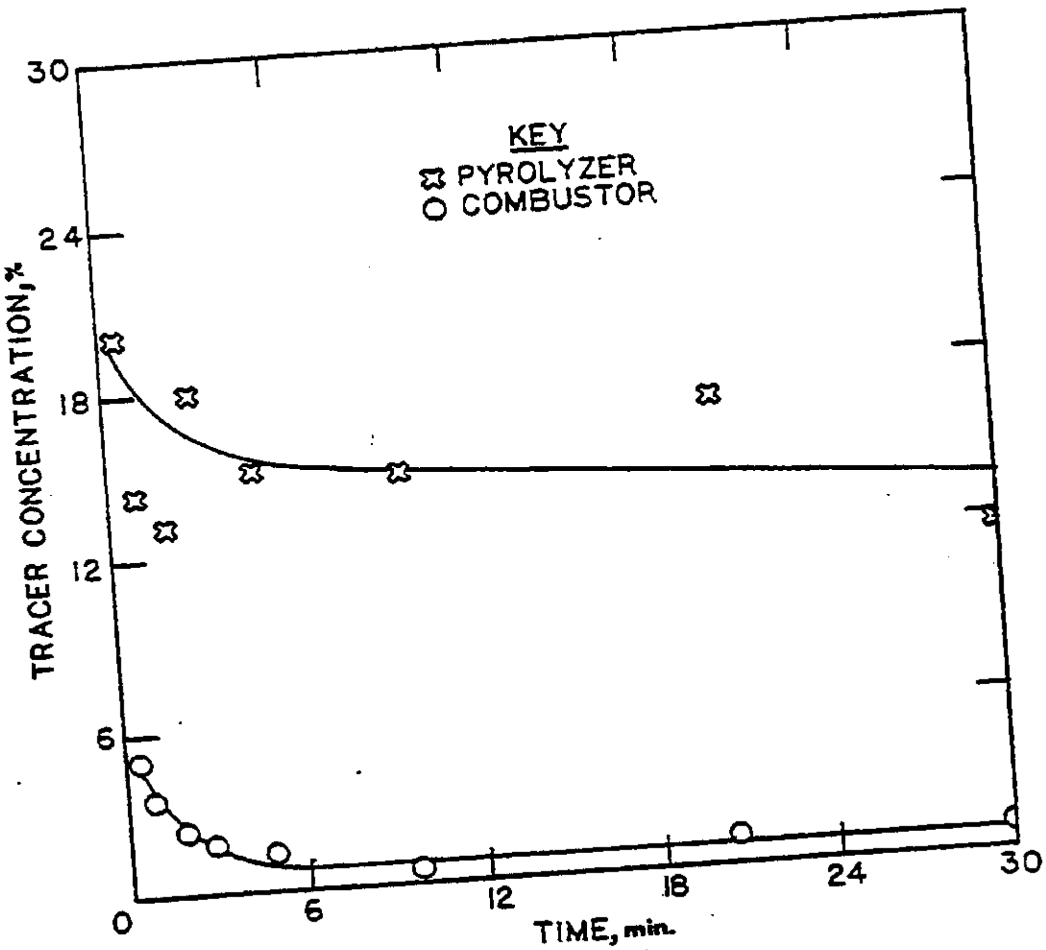

Figure 29. The average concentration of tracer particles in the combustor and pyrolyzer at various times.

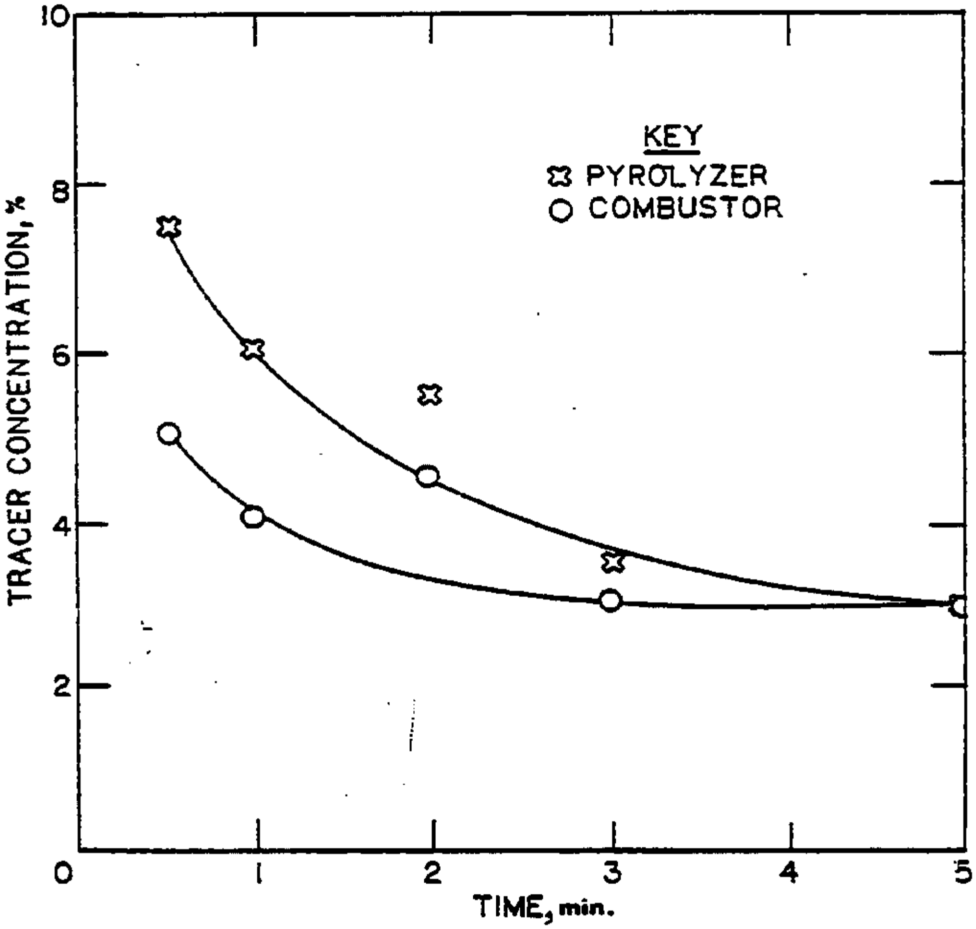

Figure 30. The average concentration of tracer particles in the combustor and pyrolyzer in the modified reactor.

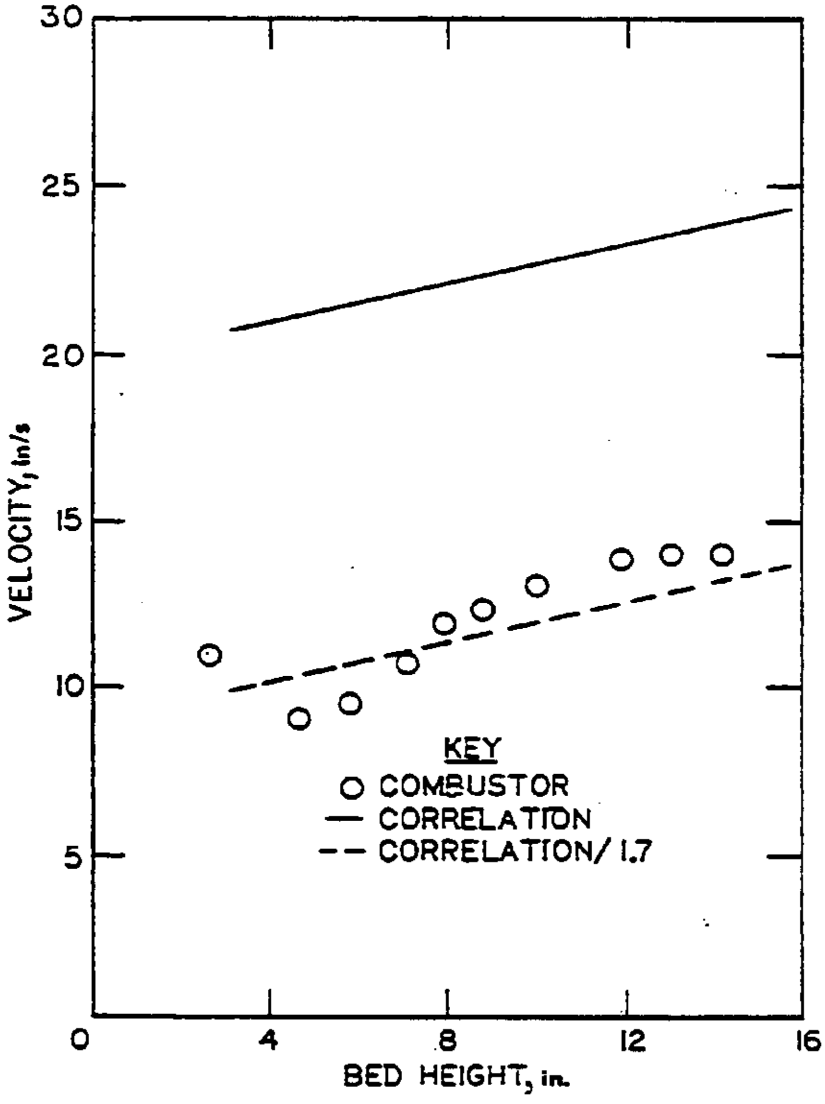

Figure 31. The visible bubble velocity at several bed heights.

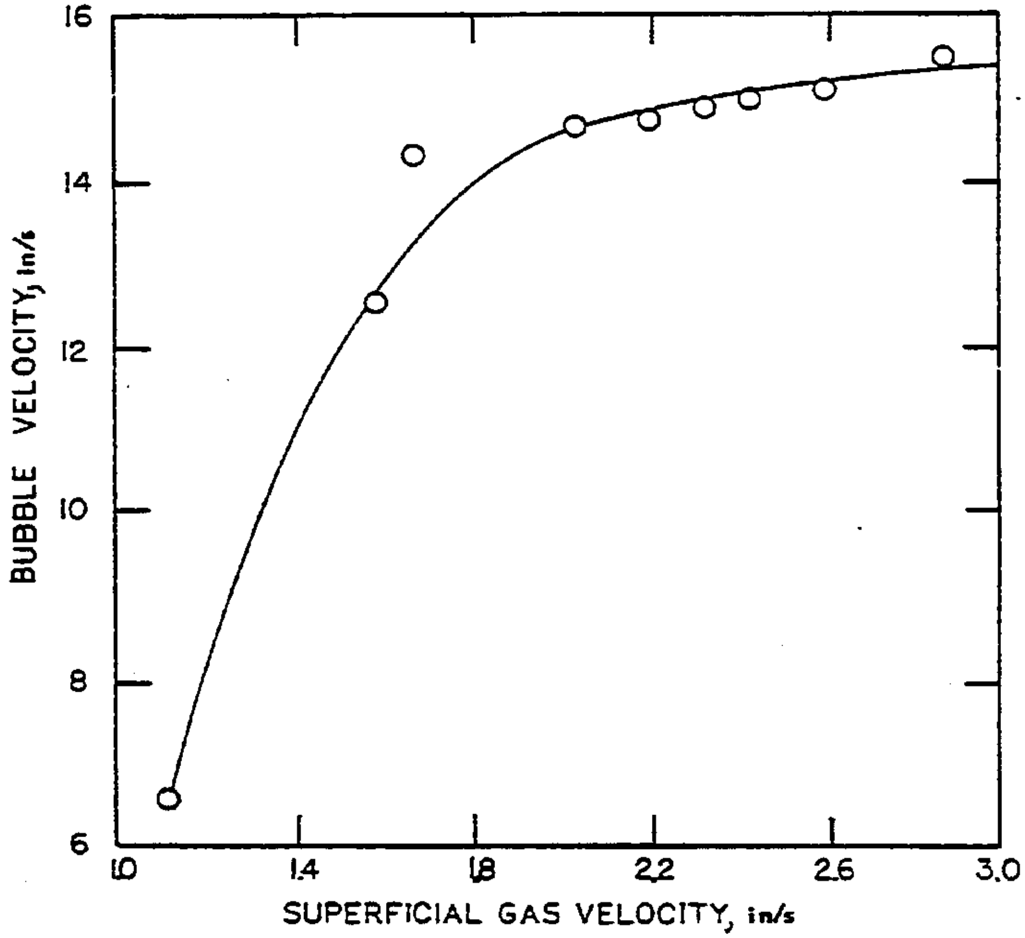

Figure 32. The effect of the fluidizing gas flow rate on the visible bubble velocity.