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CITY UNIVERSITY OF NEW YORK, P.H.D., 1979

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TURBULENT AND FAST FLUIDIZATION

(HIGH VELOCITY FLUIDIZATION)

by

DAVID H. LARSON TURNER

A dissertation submitted to the
Graduate Faculty in Engineering in partial fullfillment
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ABSTRACT

As the velocity of a gas flowing across a bubbling fluid bed is raised, the heterogeneous, two-phase structure of the bed first peaks, and then gradually changes giving way to a condition of increasing uniformity culminating in the turbulent state in which on the whole large voids or bubbles are absent. A turbulent fluidized bed has an upper bed surface, though it is considerably more diffused than in a bubbling bed owing to greater particle carryover rates attending operation at higher gas velocities. The turbulent regime extends to the transport velocity. As the transport velocity is approached there is a sharp increase in the rate of particle carryover, in the absence of solid recycle, the bed would empty in short order.

Beyond the transport velocity, solid fed to the vessel will traverse it in transport flow and the concentration of the resulting suspension will depend not only on the velocity of the gas but also on the rate at which solid is circulated through or fed to the vessel. If the solid rate is low, dilute-phase flow will result. If, on the other hand, the solid rate is sufficiently high, then the entrained suspension becomes dense and would be characterized by solid mixing - a state of extreme refluxing of dense strands and clusters of particles. Such a suspension, which may have a solid concentration typically in the range of 5 to 20 per cent (by volume), displays many of the characteristics of the

fluidized state, and it has accordingly been named the fast fluidized bed.

This thesis presents and discusses the results of experiments conducted at room conditions in a two-dimensional (2 x 20 inches) and a round 6-inch I.D. beds, on the transition from bubbling to turbulent fluidization of several solids of different sizes and densities. The fast fluidization characteristics of a fine cracking catalyst in a round 3-inch I.D. bed will also be presented and discussed. Finally, the character and range of turbulent and fast fluid beds (collectively called high velocity fluid beds) will be explored.

The main conclusions of this thesis are:

The transition from bubbling to turbulent fluidization is reflected in the fluctuations of both the dynamic pressure at any point in the bed and the pressure drop across the bed.

This transition can be characterized by two velocities: U_c, the velocity at which the fluctuations begin to dimish from their peak value, and U_k, the velocity at which the fluctuations have decayed to a sufficiently low level and simultaneously begun to level off. U_k marks the onset of the turbulent regime.

Both U_c and U_k generally increase with particle size and density.

The ratios U_c/U_T and U_k/U_T (where U_T is the particles terminal velocity) approach or even exceed 10 for beds of fine powders, while they diminish with size and density, and could become less than 1 for beds of coarse solids.

Experiments with FCC in the 3-inch fast bed has shown that the contribution of the wall-solid friction to the pressure drop was always less than 10%.

In the 3-inch fast fluid bed, slip velocities are an order of magnitude greater than the free fall velocity of the individual particles.

High velocity fluid beds of fine solids enjoy the following advantages over bubbling fluid beds: higher processing capacities per unit volume of reactor space, a more intimate degree of contact between gas and solid, a lesser degree of gas back-mixing, a capability of receiving and handling cohesive solids which might agglomerate; and they may be easier to scale up.

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TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

Page

LIST OF EXHIBITS

NOMENCLATURE

D	column diameter
$\bar{d}$	volume surface mean diameter; $\bar{d} = 1/\sum_i x_{i/di}$
d_i	particle diameter for given size cut
G_s	solid rate, lb/sec per ft^2 of cross sectional area
H	column height
$\Delta P_{measured}$	pressure drop during flow conditions
$\Delta P/\Delta L$	pressure drop during flow conditions per unit length
ΔP_{gr}	pressure drop due to the weight of solid
U_b	absolute rise velocity of slugs
U_c	velocity where the pressure fluctuations across bubbling bed begin to diminish from their peak value
U_g	superficial gas velocity
U_k	velocity where turbulent regime begins
U_{mf}	minimum fluidization velocity
U_T	terminal velocity for an individual particle
X	pressure drop due to friction
x_i	mass fraction for particles of size d_i

Greek Symbols

ρs	particle density
ϕs	particle sphericity

I. INTRODUCTION

The schematic shown in Figure 1 may serve to introduce the subject and outline the scope of this thesis. First, consider a bubbling fluidized bed (Figure 1a). As the velocity of the gas flowing across the bed is raised slowly, the heterogeneous, two-phase character of the bed first peaks, and then gradually changes giving way to a condition of increasing uniformity culminating in the turbulent state (Figure 1b) in which large bubbles or voids are on the whole absent.

In the turbulent bed, there is an upper bed surface which is considerably more diffused than in a bubbling bed because of the greater freeboard activity attending operation at higher gas velocities.

The turbulent regime extends to the so-called transport velocity. As the transport velocity is approached, there is a sharp increase in the particle carryover rate; and in the absence of solid recycle, the bed would empty in short order. Beyond the transport velocity, solid fed to the bottom of the bed or vessel transverses it in transport flow, and the concentration or density of the resulting suspension depends not only on the gas velocity but also on the solid flow rate. If the solid rate is low, dilute phase flow will result. The density of the suspension increases with the solid rate, at a given gas velocity. If the solid rate is sufficiently high, the suspension may become quite dense (solid concentration by volume could typically reach as high as 20-25%) and take on many of the characteristics we have come to associate with the fluidized state. This condition has been dubbed at City College the fast fluidized bed. A typical installation in which the fast bed condition can be achieved involves an external cyclone, and a relatively large standpipe (Figure 1c).

The turbulent and the fast fluidized beds are what we have come to call high velocity fluidized beds.[1] As its name suggests, a high-velocity fluidized bed embodies three essential elements: (1) It operates at gas velocities beyond the bubbling regime, and it is indeed free of any large voids or bubbles. (2) It preserves all the underlying qualities associated with the fluidized state, namely, the ability of maintaining a uniform temperature, the capability to bring a cold gas or solid feed almost instantaneously to bed temperature, etc., and indeed the very quality of resembling the liquid state. (3) A high solid concentration can be maintained even at high gas velocities by circulating solid to the bottom of the vessel at suitably high rates.

High velocity fluidized beds may afford several important advantages over the bubbling bed, and may accordingly suit several processes far better than the latter.

Among the advantages are higher processing capacities, more efficient and intimate contact between gas and solid, and the capability to receive and handle cohesive solids. High velocity fluidized beds may have a role in processes converting coal into gaseous and/or liquid products.

High velocity fluidized beds have received little attention in the literature, and though they have been the subject of some commercial practice, they remain little explored. We have accordingly undertaken the study of their fundamental characteristics in cold model equipment. This thesis presents and discusses data on transition of beds of several solids from bubbling to turbulent fluidization, and on the behavior of a fine cracking catalyst in the fast fluidized bed. This thesis incorporates, for completeness, some experimental results obtained by my colleagues at the City College which bears directly or elucidates the phenomena upon which I have focused. Such work is suitably referenced.

Section II provides technical background describing previous work in the subject area, as well as instances of commercial use of high velocity fluidized beds. Section III describes the experimental equipment. Section IV and V presents and discusses results on the transition from bubbling to turbulent fluidization, and on the behavior of a fast bed of a cracking catalyst, respectively. The last section highlights some of the major conclusions.

II. TECHNICAL BACKGROUND

To our knowledge, Lanneau[2] was the first to recognize the transition from bubbling to turbulent fluidization, and the potential advantages of the turbulent regime. He studied the fluidization characteristics of a fine powder in a bed 15 ft. deep, contained in a column 3 inches inside diameter, as the velocity of the gas was raised to 5 ft/sec. At a given gas velocity, he obtained traces of instantaneous point densities from a small-point capacitance probe inserted in the bed. Lanneau conducted experiments at two pressures, 10 and 60 psig. He noted that as the velocity was raised beyond 1 ft/sec, the heterogeneous, two-phase character of bubbling regime gradually gave way to a condition of increasing homogeneity. At velocities in the range of 3 - 5 ft/sec, the tracings from the probes indicated, in his words, that a condition of "almost uniform or 'particulate' fluidization was approached." Lanneau accordingly drew a rationale to argue that, for a fine powder similar to the one he used, a velocity around 1 ft/sec, where the heterogeneous structure of the bubbling bed is most pronounced, is the worst velocity to use from the standpoint of efficiency of contact between gas and solid. On the other hand, as the velocity is raised from 1 ft/sec toward the "uniform regime" around 3 - 5 ft/sec, the interaction of gas and solid and their efficiency of contact increases.

Because Lanneau worked in steel equipment, it remained for Kehoe and Davidson [3] using transparent apparatus to describe the transition from bubbling (slugging, in fact) to turbulent fluidization, and it remained for Massimilla [4] to provide a laboratory research demonstration of the higher contacting efficiency of turbulent fluidized beds.

As the velocity of the gas flowing through a narrow bed of a fine powder was raised, Kehoe and Davidson witnessed a breakdown of the slugging regime into "a state of continuous coalescence - virtually a channelling state with tongues of fluid darting in a zigzag fashion through the bed." Kehoe and Davidson used the work <u>turbulent</u> to describe this state.

The breakdown of the bubbling regime to a turbulent state was also observed by Carotenuto, et. al. [5] and Thiel and Potter [6].

Observations of the transition from bubbling to turbulent fluidization cited above were all made in beds of essentially fine powders. More recently, a team at General Electric has reported results on the transition in question for coarse particles [7]. The experiments were conducted with two beds of uniformly sized glass beads, 650 and 2600 microns in diameter, in two square vessels, 1' x 1' and 2' x 2', and at pressures up to 10 atmospheres. In addition to direct observations and photography, measurements from capacitance probes of the ratio of the oscillating-to-average steady state bed pressure drop, the bed dynamic pressure drop signal character, and the local void fraction signal character, were also used to follow and identify the changes in fluidization regimes. Staub and Canada [7] report the transition from bubbling to turbulent fluidization in terms of the ratio of transition velocity, U_k, to the terminal velocity, U_T, of the particles in the bed. Accordingly, for the bed of 650 microns glass beads, the transition from bubbling to turbulent fluidization occurred at U_k/U_T around o.6 - o.7, and for the bed of 2600 micron glass beads at a corresponding ratio around o.3 - o.4. These results remained essentially unchanged over the pressure range used, 1 to 10 atmospheres.

As for commercial utilization, the open literature contains precious little information that would permit close identification of the regime in which a given fluid bed operates. One gets the impression that in several commercial fluid beds of fine powders, the fluidizing velocity has been, as a matter of course, slowly increased over the years to the range of 2 - 3 ft/sec and beyond. Such beds might very well be operating in the turbulent regime. The regenerator of the fluid catalytic cracking process is probably a case in point. In the early cat crackers, the regenerator was run at a velocity around 1 ft/sec. The regenerator in many of the modern fluid crackers typically operates at 2.5 to 3 ft/sec, considerable carryover is reported; and with the concentration of solid in the freeboard being appreciable all the way to the cyclones, the upper surface of the bed appears to be rather diffused.[8]

There is, however, little doubt that most commercial fluid bed roasters of pyritic feedstocks operate in the turbulent regime. This arises from the very nature of the roasting process. The feed typically comprises of fine particles which are particularly vulnerable to agglomeration should hot spots develop in the distributor zone where intensely exothermic reactions take place. The designer must accordingly provide for highly turbulent solid mixing throughout the distributor zone. This is accomplished by a suitable distributor design and by employing gas velocities in the range of 3 - 5 ft/sec. At these velocities, particle carryover rates are relatively high; accordingly, if additional solid residence time is needed, the roaster may be expanded toward its top -- and many commercial roasters do indeed have this configuration -- or carryover matter may be recycled. The INCO roaster[9] belongs to the second category. Two 550 ton/day roasters have operated at Copper Cliff, Ontario, Canada. The ore, a Canadian Phyrrotite containing nickel, is a flotation product with a size of 90% under 200 mesh. It is pumped into the roaster where the gas velocity is maintained around 3.5 ft/sec. Apart from few coarse particles which arise from agglomeration, the entire fine matter is carried aloft in the form of what INCO's engineers have termed "a flying bed." Much of the material is circulated through a combined waste heat boiler-cyclone separator, while a small production stream is withdrawn.

Few realize today that the first commercial fluid catalytic cracking plant operated in the fast fluidized bed mode. The plant, whose reactor and regenerator measured respectively 15 and 19.5 feet in inside diameter, was erected at the Baton Rouge refinery of the Standard Oil Company of New Jersey. It was placed in operation on May 25, 1942, "with surprisingly little difficulty."[10] The plant was designed by the Standard Oil Development Company and was the culmination of extensive pilot plant development and several years of research on techniques for contacting gas with a fine solid catalyst. That research enjoyed the pioneering input of Professors Lewis and Gilliland of the Massachusetts Institute of Technology.

Along with the fast fluidized bed, two other options were considered. The now familiar bubbling fluidized bed offered the advantage of a large surface of contact between gas and solids per unit volume of reactor vessel and of uniformity of temperature throughout the bed, but its characteristic low gas velocities spelled correspondingly low-gas-treating capacities. High gas-treating throughputs could be achieved by contacting the solid with a high velocity gas in dilute phase flow, but the low solid concentrations typical of this condition would have required, save in cases of unusual catalyst activity, very tall vessels or the use of multiple-tube coiled reactors. Also, in dilute-phase reactors, endothermic or exothermic reactions would tend to establish a temperature gradient along the flow path.

Operation in the fast bed regime was chosen since, in the words of Lewis and Gilliland,[11] it could "secure in large degree the advantages of both types of operation just described, and at the same time eliminate or greatly reduce the disadvantages." They described the technique as follows:

"If one will operate at a gas velocity sufficient to blow all or substantially all of the solid material out of the reactor in a relatively short time, provided no fresh solid material be introduced during this time, but will feed into the reactor simultaneously solid material at a sufficiently high rate, one can maintain in the reactor a high concentration of solid granules approaching that of the 'liquid state' (of the bubbling fluidized bed). . . and yet be blowing the solid particles out of the top of the reactor at a corresponding rate." Lewis and Gilliland reported the results of experiments in which typically a pulverized clay catalyst (with a settled density of 35 lb/cu ft) was maintained at a fluidized density of 10 lb/cu ft at a superficial gas velocity of about 8 ft/sec.

The name "fast fluidization" was only recently coined by us[12] to describe the above operation. Lewis and Gilliland and Standard Oil's engineers simply regarded it as another mode of fluidization, and referred to it as the "upflow" operation to distinguish it from low-velocity bubbling fluidization which they termed "downflow" operation.

The success of the first fluid bed cat cracker at Baton Rouge led the way to additional installations and by the end of 1943, several upflow plants had gone into production. The operation of these plants, however, was not free of difficulty. From the vantage point of 1978, these plants appear to have been poorly designed. Dust collection and control of solid hold up in the reactor and regenerator posed problems. As a result, industry switched to the use of the bubbling fluidized bed, that is to operation in the downflow mode where dust recovery was simpler, the height of the solid in the reactor could be set at will, and where the mechanical design was on the whole simpler.[10] Those who had earlier expounded the qualities of the upflow plants[13] were now singing the praise of "the new improved" downflow technique.[10]

The advent of the fluid catalytic cracking process must certainly be regarded as one of the most significant landmarks in the history of chemical technology. It made highly visible the fluidization technique which was soon adopted by the process industries toward many and diverse purposes. It is therefore unfortunate that out of the turbulent years of the war, the low-velocity bubbling bed emerged as fluidization's archetype, and that the passage of time has nearly obliterated all records of those early crackers which operated in the upflow mode, and the principles they embodied. As a result, far too much emphasis has been placed upon the bubbling bed both in industry and the academic world.

For all its advantages, the bubbling bed suffers several serious drawbacks: The low gas-treating capacity typical to this regime has already been mentioned; in addition, the efficiency of contact between gas and solid heterogeneous structure of the bed, owing to the presence of bubbles, complicates scale up; and solid mixing, which generally stands in direct proportions to the gas velocity, may not in some applications by sufficiently vigorous, for example, in cases where the particles display a tendency to agglomerate.

It is ironic that while industry in general stuck to the low-velocity bubbling bed, evolution of the fluid cracking process has taken a reverse turn once again in direction of high gas velocities. In today's plants, the cracking reactions are conducted in a transport riser reactor in which gas velocities are typically in excess of 35 ft/sec toward the top;[14] and as we have noted earlier, whereas gas velocity in the early downflow regenerator was around 1 ft/sec, most of today's regenerators operate in the turbulent regime, and at least one company[15] is offering a regenerator operating in the fast bed mode.

Lurgi Chemie and Huttentechnik GmBh of Frankfurt, West Germany, was the first to appreciate the broad commercial potential of the fast fluidized bed. Lurgi realized this potential in its successful process, developed jointly with Vereinitge Aluminum Werke, for calcining aluminum hydroxide to supply cell-grade alumina.[16] Figure 2 shows a schematic of Lurgi's fast bed calciner. Calcination is normally conducted at a temperature between 1000 and 1100°C. Heat is provided by direct injection of oil or fuel gas to the lower, denser portion of the bed; the primary air is sub-stoichiometric, and combustion is accordingly completed in the upper region of the bed. The temperature, however, remains remarkably uniform throughout the bed. The development of Lurgi's calciner suggests that the fast bed might prove considerably easier to scale up than the bubbling fluid bed. Lurgi's scale up was made in two steps; a 5-inch I.D. unit was followed by a pilot 3 ft. in inside diameter and 26 ft. tall. Both the pilot and the first commercial plant (which was 10.5 ft. I.D.) achieved design operation within a few weeks of startup. At any rate because of the fast bed's high capacity, it might often not be imperative to scale the equipment to uncomfortably large dimensions.

In the early forties, a major effort was launched to develop a bubbling fluid bed process for Fischer-Tropsch synthesis, using powdered iron as catalyst.[17] A large plant, the so-called Carthage Hydrocol, was subsequently built in Brownsville, Texas. The effort met with failure owing, among other things, to conversions well below those anticipated on the basis of laboratory experiments. In today's only Fischer-Tropsch plant in Sasolburg, South Africa, the synthesis reaction is carried out in a circulating high-velocity fluid bed (the Synthol reactor). The plant has seen successful operation since 1954,and large expansion has recently been announced.

III. EXPERIMENTAL

The City College test stand for the study of high velocity fluidization includes three systems: the so-called two-dimensional, the 3-inch, and the 6-inch systems.

Figure 3 is a schematic of the 3-inch system. The 3-inch bed, 28 feet tall, is on the right. On the left is a 11.5-inch I.D. round companion bed in which the test solid is maintained in low-velocity, bubbling fluidization. This so-called "slow" bed doubles as a standpipe and storage for the solid. Fluidizing air for the 3-inch bed is provided at its bottom. Solid can be maintained in the 3-inch bed at any desired fluidization condition through independent control of the gas velocity, and of the solid rate from the slow bed to the bottom of the 3-inch bed. The solid rate is controlled by a butterfly valve installed just below the bottom of the slow bed. Solid carried over from the 3-inch bed is separated from the gas via cyclones. The gas leaving the cyclones flows to a bag filter. Dust losses are usually very small. The test column is constructed in alternating glass, plexiglass and copper sections to permit visual observations and the introduction of instrumentation. Pressure taps have been installed liberally in many locations around the system.

Solid circulation rates are measured with the aid of a sintered-plate butterfly valve installed in the middle of the 11.5-inch bed. At any given moment, the butterfly valve can be closed, and the rate of descent of the bed level immediately below the valve timed. Solid, returning to the companion bed through the cyclone diplegs, forms a fluidized bed on top of the butterfly valve, which acts as a distributor for this upper bed. It should be appreciated that, apart from some additional pressure introduced by the porous butterfly valve, the solid head in the entire companion bed remains constant throughout this procedure.

The two-dimensional and the 6-inch systems are assembled in a similar manner to the 3-inch system. The two-dimensional bed is 2 x 20 inches in cross section and is 23 feet tall. The 6-inch bed is 28 feet tall.

Fluidizing air for the 6-inch bed is provided in a manner similar to that in the 3-inch bed. Fluidizing air for the two-dimensional bed is provided by a sintered plate and two separate manifolds into the bed with six ports each. The sintered plate along the bottom of the bed allows the superficial gas velocity to be varied from about 0.5 to 3.0 ft/sec. The first manifold (3 inches above the porous plate), provides for a range from 3.0 to about 8.0 ft/sec; and the second manifold (6 inches above the porous plate) allows superficial gas velocities up to 20 ft/sec.

Table 1 lists the solids which were tested, and their properties. The three finer solids (A,B and C) belong to group A of Geldart's classification;[18] the coarser solids all fall into group B. Particle sphericities were determined from optical image analysis provided by DuPont Company, Experimental Station, Wilmington, Delaware. A "shape factor number" defined as $(\text{perimeter})^2/(\text{projected area})$ is optically determined for a number of particles using proprietary computer programs. The particle sphericity is then calculated as the inverse of the "shape factor number."

IV. TRANSITION FROM BUBBLING TO TURBULENT FLUIDIZATION

To study the transition from bubbling to turbulent fluidization, a bed of a given solid was established at the bottom of either the 6-inch bed or the two-dimensional bed. The behavior of the bed was observed as the fluidizing air was raised slowly. To compensate for solid carried out of the bed, the valve controlling the flow of solid from the slow bed to the bottom of the two-dimensional or 6-inch bed was kept fully open.

One cannot overemphasize the role of direct visual observation in detecting changes in the character of the fluidization as the velocity across the bed is raised. In addition, the transition from bubbling to turbulent fluidization in the 6-inch bed was followed by recording the fluctuations in the pressure drop measured across the bed. In the two-dimensional bed, the tracing of the dynamic pressure at a point midway in the bed was obtained by means of a high sensitivity pressure transducer, trade name: Piezotron, manufactured by Sundstrand Data Control, Inc. This pressure transducer has a rise time, 10 to 90%, of 3 microseconds.

IV. I Results and Discussion

The first experiment was conducted with the fluid cracking catalyst (solid B) in the 6-inch bed.[19] For this solid, bubbling is evident at a gas velocity around 0.1 ft/sec, and slugging is in force at 1 ft/sec. As the velocity of the gas is slowly raised beyond 1 ft/sec, the slugs lengthen, and the expansion and contraction of the bed increases in violence. This is reflected in the fluctuations of the manometer that serves to measure the pressure drop across the bed (Figure 4). At a superficial gas velocity around 2 ft/sec, the slugs appear to break down giving way to a turbulent state in which large voids or bubbles are absent and which the naked eye indeed perceives as rather uniform in structure. Solid mixing is vigorous, and extensive refluxing of large dense packets of particles is evident. The apparent sharpness of this trnasition and the greater degree of homogeneity of the turbulent regime that lies beyond it are also reflected in the pressure drop fluctuations shown in Figure 4. Figure 4 also shows that if, having reached a velocity around 5 ft/sec, the gas velocity is now slowly reduced, one retraces essentially the same fluidized states.

The behavior of FCC in the 6-inch bed turned out to be rather unique. Whereas in this case the transition from bubbling to turbulent fluidization is very sharp, in all the other experiments we have conducted, both in the 6-inch and in the two-dimensional beds, the transition in question occurred gradually over a velocity range whose boundaries and length depend on the properties of the test solid. Figures 5-7 demonstrate the corresponding results, obtained in the 6-inch bed, for solids A, C, D and E, respectively (see Table 1).

The transition may be characterized by two velocities: U_c, the velocity at which the fluctuations begin to diminish from their peak value, and U_k, the velocity at which the fluctuations have decayed to a sufficiently low level and simultaneously have begun to level off. The values of U_c and U_k for the results shown in Figure 4-7 are given in Table 2 which also provides a summary of the results of all our experiments.

Inasmuch as the experiments in the 6-inch bed were dominated by slugging behavior, we repeated some of them, and conducted experiments with additional solids in the two-dimensional bed. The 2-inch thick, two-dimensional bed does give rise of course to wall effects peculiar to this geometry, but at least laterally the two-dimensional column placed less constraint on bubble growth and its transparency afforded better visualization of the transition from bubbling to turbulent fluidization.

A series of photographs, depicting the 2-D bed while the superficial gas velocity is increased, are shown in Exhibits 1-8. HFZ-20 (Solid C) was the test solid.

Exhibits 1 and 2 show the bubbling fluid bed at 0.51 ft/sec. At this low gas velocity, there is hardly any particle carryover. Exhibit 2 shows a small bubble going up the left side. Exhibits 3-5 show the bubbling fluid bed at 3.0 ft/sec. Here, the bubbles are larger, and there is more carryover. Exhibits 6 and 7 show the turbulent fluid bed at 3.9 ft/sec. Here, the upper bed (Exhibit §) shows denser carryover, and the lower bed (Exhibit 7) appears more homogeneous. To complete the story, Exhibit 8 shows a rather dilute fast fluid bed at 11.5 ft/sec. The two bright spots, from strong direct back lighting, indicate cluster formation.

The first series of photographs (Exhibits 1-5) follows the dynamic pressure fluctuations caused by bubbles passing through the bed, beginning with small infrequent bubbles (Exhibits 1-2) and progressing to larger more frequent bubbles (Exhibits 3-5).

The second series of photographs depicts the stabilization (at a low value) of the dynamic pressure fluctuations ue to the increased homogeneity of the high velocity fluid bed (Exhibits 6&7 - turbulent bed; Exhibit 8 - fast bed). The dymamic pressure fluctuations were measured by the pressure transducer inserted in the 2-D bed.

Also as noted earlier in the two-dimensional bed, the transition in question was followed by tracing the fluctuations of the dynamic pressure in the middle of the bed by means of a high sensitivity pressure transducer. A series of such tracings, photographed from an oscilloscope during an experiment with HFZ-20 (Solid C) are shown in Figure 8. From these tracings, the mean peak-to-peak oscillating component at a given superficial gas velocity was measured and then divided by the fluidized density of the bed at a given velocity. This is plotted in Figure 9. This curve and the one shown in Figure 6 which gives the relative fluctuations recorded for the same solid in the 6-inch bed, are quite similar in outline. But the transition in the two-dimensional bed occurred earlier; that is, both U_c and U_k recorded in the two-dimensional bed are lower than the corresponding values for the 6-inch bed.

We should note that in all the experiments, save those with the three finer solids, strong wall effects were clearly evident in our rather small equipment. Slugging was particularly intense in both units, and channelling arose in the two-dimensional bed during experiments with coarse solids. As a result the corresponding data often suffered considerable scatter. In this Thesis, I have presented only data we have judged to be sufficiently reliable to serve as the basis for the conclusions arising from our work. I have in fact, conducted experiments with several solids coarses than the coarsest solid listed in Table 1. But owing to strong wall effects, we cannot report the results with any degree of certainty. A study of beds of large particles which parallels ours ought to be conducted in equipment of considerably larger scale.

Several observations can be drawn from the results summarized in Table 2:

(1) Both U_c and U_k generally increase with particle size and density, and may be conveniently plotted as functions of $(\rho_s \bar{d})^{1/2}$ -- see Figure 10-- where ρ_s is the particle density and d is the column-surface mean diameter of the solid. $\bar{d}$ is defined by

$$\bar{d} = \sum_i 1/(x_i/d_i)$$

where x_i is the mass fraction of particles of size d_i.

In addition to the results for solid C, noted above, other results also appear to indicate that the transition from bubbling to turbulent fluidization occurred in the two-dimensional bed earlier than in the 6-inch bed; that is, either or both U_c and U_k recorded in the two-dimensional bed are lower than the corresponding values for the 6-inch bed. Thiel and Potter[(6)] have conducted experiments with a cracking catalyst, similar to our solid B, in beds of 2, 4 and 8.5 inches in inside diameter, and have reported that the transition from bubbling to turbulent fluidization was strongly dependent on bed diameter. They report their results in terms of a single transition velocity, and note that this velocity decreased sharply with bed diameter. Although our two-dimensional bed is smaller in cross sectional area than the 6-inch bed, it does afford growth of bubbles larger than 6 inches; therefore, the two-dimensional bed could be viewed as effectively "larger" than the 6-inch bed. In that sense some of our results, corroborate, at least qualitatively, the results of Thiel and Potter. Further, since the transition in question involves change in the bubbly structure of the bed and since, for a given solid, the size and rise velocity of a bubble does depend upon bed diameter within some range typical to laboratory equipment, it therefore stands to reason, that over that size range, the transition from bubbling to turbulent fluidization may show some dependence on bed diameter. In sufficiently large beds, the behavior of the bubbles will no longer depend on bed diameter; therefore, the same ought to hold for the transition to the turbulent regime.

(2) The range of gas velocities over which the transition from bubbling to turbulent fluidization occurs generally increases with particle size and density.

(3) Plots of U_c/U_T and U_k/U_T (where U_T is the terminal velocity of a particle of size $\bar{d}$) versus $(\rho_s \bar{d})^{\frac{1}{2}}$ appear in Figure 11 and are very revealing. These ratios approach or even exceed 10 for the finer solids, and decrease sharply with increasing particle size and density. As noted previously, some experiments have been conducted with particles larger than those of solid E; however in our equipment, beds of such solids were subjected to severe wall effects and we cannot report quantitative results with any degree of certainty. Therefore, the results of Staub and Canada[(7)] (obtained in relatively large beds 1' x 1' and 2' x 2') show that for larger particles (650 and 2600 microns glass in this case) U_k/U_T becomes smaller than unity.

The greater than unity ratios of Figure 11 arise from a property of fine solids, .i.e., their tendency to form clusters or agglomerates. The structure of a turbulent fluidized bed of fine solids is governed by the strong interaction of two phases: A dense phase consisting of closely packed clusters or aggregates and streamers of particles whose motion is mostly downwards, and a lean dispersed phase in which individual particles and small clusters zigzag their way upwards. As a result, the behavior of the bed as a whole reflects the properties of not only the individual particles of the lean phase but also the clusters of the dense phase. From the ratios of Figure 11 it appears that the effective size of these clusters, relative to the size of the individual particle, stands in inverse proportion to the size of the individual particle.

The existence of clusters and the concept which attaches to them some effective size is useful in understanding the range of the turbulent regime for a fine solid. First, however, consider a fluidized bed of closely-sized coarse particles, such as those in the experiments of General Electric.(7) Both the existence and the boundaries of the turbulent regime are fairly clear in this case. Take for example, the 2600 microns glass particles. For this solid, General Electric reports U_k to be around (0.3 - 0.4) U_T. The turbulent regime accordingly extends from (0.3 - 0.4 U_T to a velocity at least equal to U_T. Over this range, since the gas velocity obviously lies below the terminal velocity of the individual particle, no particle carryover will in fact be experienced provided the freeboard is sufficiently high.

A somewhat analogous situation holds for a turbulent bed of a fine solid, except that here the focus has shifted from large individual particles to large clusters of small particles. The effective size and density of these clusters, at any gas velocity in the turbulent range, is such that their terminal velocities remain greater than the velocity of the gas. That is why a bed of fine particles can be maintained at gas velocities that are 10 to 20 times the terminal velocity of its median particle. Entrainment of particles will of course take place owing to the individual particles being swept up by the rising gas from the lean phase near the top of the bed, the erosion of clusters by the rising of the phase, and other conceivable mechanisms. As the velocity of the gas is raised across the turbulent range, the carryover rate increases, the solid hold up in the freeboard increases, and the upper surface of the bed becomes considerably more diffused. At the same time, the fluidized density of the bed decreases, as does the effective size of the clusters.

The turbulent regime extends to a gas velocity at which the effective cluster size has diminished to a point where its terminal velocity becomes comparable to the velocity of the gas. As the gas velocity approaches this point, there is a sharp increase in the rate of solid carryover from the bed, and in the absence of solid recycle, the bed would empty in short order. This point shall be designated the <u>transport velocity</u>. Beyond the transport velocity, solid fed to the bottom of the bed traverses it in transport flow. The density or concentration of the resulting suspension depends not only on the velocity of the gas but also on the solid flow rate. If the solid rate is low dilute-phase flow results. If, on the other hand solid is fed to the bed at a sufficiently high rate -- for example, by circulating solid carried over from the bed to its bottom via a cyclone and a stand-pipe -- then it is possible to maintain in the bed a large solid concentration typical to the fast bed condition. This is in contrast to the turbulent and bubbling fluidized beds, where the fluidized density depends primarily on the gas velocity and only slightly on the rate at which solid may be fed or withdrawn from the bed. Also, given a sufficient freeboard space, the concentration of solid above a bubbling and a turbulent fluidized bed operating at a given gas velocity decays through the transport disengaging height and approaches a constant level. If solid were fed to the bed at a rate beyond the saturation carrying capacity, the bed level would rise until the new solid rate is matched by solid carried out of the bed. The gas exit point would now lie within the transport disengaging height. But unless the slip velocity changed significantly, the fluidized density of the bed itself would remain virtually the same.

Therefore, the transport velocity divides the vertical gas-solid flow regimes into what might be called the "stationary" states, including the bubbling and the turbulent fluidized beds, where, save for relatively small particle carryover, the bed on the whole experiences no net flow and remains at the bottom of the holding vessel; and the transport regime which encompasses a wide range of states from dilute-phase flow to the fast bed condition.

For the fluid cracking catalyst (solid B), the transport velocity lies around 4-5 ft/sec. Referring back to Figure 4, we may accordingly note that the turbulent regime, for this solid, stretched from 2 ft/sec to around 4-5 ft/sec. Over this range, although the fluidized density decreases and the particle carryover rate increases, the vertical pressure drop profile (see Figure 12) and the naked eye perceives a dense bed with an upper bed surface which is of course more fuzzy than that of a bubbling bed. Figure 4 attests to the rather high fluidized densities measured at the bottom of the bed throughout the turbulent range.

As the transport velocity is approached, there is a sharp increase in the rate of solid carryover from the bed, and if it were not for solid flowing from the slow bed through the transfer line the bed would empty in short order. This may be demonstrated by the results, shown in Figure 13, of a simple experiment in which we suddenly closed the solid slide valve and measured the time it took the bed to empty at a given gas velocity. Figure 13 gives the time, in seconds, for the pressure drop across the bed to decay 1 inch of water. Accordingly, at a velocity around 4-5 ft/sec., the bed would empty in few seconds.

Beyond this velocity, solid traverses the bed in transport flow, and the concentration of the resulting suspension depends not only on the velocity of the gas but also on the solid flow rate. If the solid rate is sufficiently high, one can produce the condition typical to the fast bed regime. This is the subject of the following section.

IV. 2. Concluding Remarks

The mechanism responsible for the transition from bubbling to turbulent fluidization is not clear. The results for the fluid cracking catalyst, shown in Figure 4, and visual observations of the 6-inch bed as well, indicate an actual and sharp breakdown of the gas slugs. This was not observed in all our other experiments where the change appeared to be gradual.

As noted earlier, Thiel and Potter [6] reported results for a fluid cracking catalyst showing that the transition velocity (they report only a single transition velocity, U_{trans}) is very dependent on bed diameter. The table below summarizes their observations.

bed diameter, inches	U_{trans} ft/sec
2	1.35
4	0.66
8.5	0.082

Thiel and Potter do not clearly characterize their U_{trans}, and in view of our own results for essentially the same solid (B), we question the rather low transition velocities they report for the 4 and especially the 8.5-inch bed. If correct, their results imply that a commercial bed of a fine solid comparable to a cracking catalyst never operates in the bubbling regime.

If correct, at least in outline, the results of Thiel and Potter may also suggest that the breakdown of the slugging regime is somewhat related to the absolute rise velocity of the slugs. Using the familiar expression

$$U_b = U_g - U_{mf} + 0.35\ (gD)^{\frac{1}{2}} \qquad (2)$$

the value of U_b may be computed as 2.15, 1.79 and 1.75 ft/sec corresponding to the transition velocities shown above for increasing bed diameter. These values could be regarded as comparable, and may help to explain why increasing the bed diameter results in a smaller gas velocity at the transition to the turbulent state.

Our own observations indicate that the transition to turbulent fluidization is generally gradual, suggesting a simpler mechanism. Namely, as the velocity is raised beyond U_c the bed has expanded to a point where it can no longer support a structure marked by distinct bubbles. As the velocity is raised further the fluidized density decreases as well, and the bed structure gradually collapses toward a condition of increasing homogeneity where the two phases -- the dense or cluster phase, and the lean phase -- can no longer be viewed either as continuous or discontinuous.

Observations of a turbulent fluidized bed in the transparent two-dimensional bed, with a strong light directed from behind, show the two phases to be strongly interacting -- a state of continuous disintegration and coalescence of both solid and gas. Occasionally, a relatively large void or bubble will zigzag quickly upwards through the bed.

In a turbulent bed of the finer solids, the de-mixing of gas and solid appears on a relatively fine scale. It is considerably coarser in a bed of larger particles, with the rising gas tracing narrow spiraling corridors, and the solid aggregating in elongated and twisted streamers. In turbulent beds of both fine and coarse solids, both vertical and lateral solid mixing appear very vigorous.

V. THE FAST FLUIDIZED BED

V. 1. The Fast Bed Condition

Fast fluidization is a technique for bringing a high velocity gas into intimate contact with a fine solid in what is essentially an entrained, dense suspension. The technique is primarily oriented toward gas-solid reactor applications, catalytic and noncatalytic. The solid in the fast fluidized bed may typically occupy up to 25% of the bed volume and is in a state of extreme turbulence marked by extensive refluxing of dense strands and packets of particles.

We have viewed the operation of the two-dimensional fast bed by means of high-speed photography. The air velocity was held fixed at 12 ft/sec., and the solid throughput was varied. At low solid throughputs, the solid conveyed upward in dilute-phase transport, but contrary to the impression created by many discussions on this subject, fine particles are not streaming upward discretely. Even at very low solid loadings (0.1 lb/ft^3) some segregation is apparent. Relatively denser clouds of particles go up surrounded by a more dilute environment. At higher solid loadings, though still within the dilute phase regime, solid segregation becomes more pronounced. Particles throng into vertical streamers which move upward surrounded by a faster moving leaner phase. Some of the streamers or strands weave a bit or even halt momentarily. At a solid loading around 2 lb/ft^3, solid backmixing gradually comes into play, and the fast bed is established.

The fast bed can be regarded as essentially a dense suspension marked by vigorous and intensive backmixing of solid. At loadings between 3 and 5 lb/ft^3, the solid at any moment appears distributed in two phases. Dense strands and ribbons rise and fall and drift from side to side at high speeds, while the bulk of the column is occupied by particles moving rapidly upward in a more dilute environment. Solid interchange between the two phases appears rapid and extensive; dense strands of solid break apart, some gradually and some in an explosive fashion, as new strands form.

At loadings beyond 5 lb/ft^3, observation of the details of the structure of the fast bed becomes difficult but suggests that each of the two phases present becomes on the whole continuous. We surmise that the strands and ribbons of the dense phase become linked in a reticulated net of rapidly circulating material that includes many vortices resembling tiny tornadoes. The impression is that gas-solid interaction is on a fine scale. The fast bed appears to the eye to afford intimate contact between a gas at high velocity and a large inventory of solid surface per unit volume of bed. There is no substantial change in the appearance of the fast bed over the 23 ft height of our equipment.

V. 2. Results and Discussion

In the remainder of this section, data for FCC is presented obtained in the 3-inch system. Figure 14 shows the pressure gradient across a middle section of the fast bed (extending from an elevation of 7.5 to 18.5 ft from the bottom of the bed) as a function of the solid rate at different superficial gas velocities. The question arose immediately as to the extent of the contribution of the solid-wall friction to the measured pressure gradients. The answer to this question was sought by conducting a series of experiments described below.

The experiments were conducted in the 3-inch system modified as shown schematically in Figure 15. A test section, approximately 4 feet in length, was installed about 10 ft from the bottom of the bed. Also, a 3-inch bypass was installed. To measure the contribution of solid-wall friction to the pressure drop across the test section, quick-action slide valves close both ends of the test section, and the flow of gas and solid is diverted by the same action to the bypass. The solid hold up in the test section is then determined and the static pressure corresponding to it is compared to the pressure drop measured across the test section just before the valves are closed. The frictional contribution to the pressure drop is reported as

$$X\% = \left(\frac{\Delta P_{measured} - \Delta P_{gr}}{\Delta P_{gr}}\right) \times 100$$

where P_{gr} is the pressure drop corresponding to the weight of solid in the test section (i.e., the static pressure).

Table 3 summarizes the results. Experiments were conducted at five different flow conditions. At each different flow condition, many runs were made. The average values for X are shown in Table 3, and it is rather puzzling that for five different flow conditions (reckoned in terms of the gas and solid flow and the resultant pressure drop) the averages were close, falling between 6.28 and 7.34%. Because of the relatively low values of X, and owing to fluctuations in the bed behavior, there was considerable variation in the data as the standard deviations indicate. Accordingly, these results cannot serve for any quantitative correlations of the contribution of wall-solid friction to the pressure drop in the 3-inch fast bed, and we may accordingly only conclude that the contribution in question is less than 10%. Since friction effects might be expected to decrease with diameter, their contribution to the pressure drop in larger beds would be practically negligible.

In viewing and treating the data shown in Figure 14, we shall, for convenience, ignore the contribution of wall-solid friction, and take the pressure gradient to reflect the fluidized density in the section of the bed in question. Whether frictional effects are included or not, Figure 14 attests to the high solid loadings that can be achieved in the middle of the column. A fluidized density of about 13 lb/ft^3 corresponds to a solid concentration by volume of about 20%, or a gas voidage around 80%. This can be achieved at a gas velocity of 10 ft/sec, and that is a gas velocity an order of magnitude greater than the velocities normally employed in a bubbling fluidized bed of the same solid.

Pressure gradients are normally higher at the bottom section of the bed, and, within the range of gas and solid rates used in our experiments, are lower toward the top of the bed. Figure 16 depicts pressure gradients measured at the bottom of the bed. Figure 17 shows pressure gradient profiles along the height of bed for three fluidizing gas velocities and several solid rates.

Slip velocities are high in the fast fluidized bed. Those corresponding to the data of Figure 14 are shown in Figure 18. It should be appreciated that the theoretical free-fall velocity of the median particle in our test solid (50 μ in diameter) is merely 0.20 ft/sec, and that of the largest particle (130 μ) is about 1 ft/sec. Yet, Figure 18 records slip velocities approaching 10 ft/sec. The large slip velocities in the fast bed are a measure of the high degree of backmixing of solid. More precisely, the high slip velocities arise from the characteristic structure of the solid in the fast bed. A dense packet of particles would naturally have an effective free-fall velocity considerably larger than that of a single particle in isolation. If the packet is sufficiently large, it cannot be sustained by the rising gas, it will fall back, and will subsequently undergo disintegration in one manner or another.

V. 3. Concluding Remarks

The fast bed regime represents a natural extension of bubbling and the turbulent fluidization regimes. In the bubbling fluidized bed, gas is contacted with solid at high concentration. The gas velocity is normally held down to avoid excessive carryover. In the turbulent bed, solid concentrations are somewhat lower, but gas throughputs are higher. The design of a turbulent bed, however, must allow for considerable carryover. In the fast fluidized bed, gas velocities are sufficiently high to effect total entrainment of the particles in a short time, but solid concentrations are still fairly high provided the material is recirculated at a suitable rate. As Figure 14 illustrates, for a given gas velocity, the solid concentration in the bed increases with solid rate, and in principle it might be possible to achieve solid concentrations comparable to those in the bubbling bed. In practice, however, very high solid rates might not be feasible, and solid densities in the fast bed would generally be smaller (though not necessarily by a large factor) than in a bubbling bed. The lower density of the fast bed need not mean a lower <u>effective</u> inventory, in view of the better gas-solid contacting of the fast regime. Moreover, the solid inventory of the fast bed may be varied over a range of values, by increasing the height of the bed, without the deterioration in gas-solid contacting efficiency commonly experienced in the case of deeper bubbling beds. For some applications, the ability of an operator to vary a fast bed's solid inventory, by changes in solid recirculation rate, may be advantageous: for example, to vary catalyst inventory to compensate for loss of catalyst activity where a degree of conversion must be closely held, or to vary carbon inventory in a fast bed gasifying carbon by steam or carbon dioxide in order to obtain a handle on bed temperature.

In bubbling, turbulent, and fast beds, gas is brought into contact with a large inventory of solid surface per unit bed volume. In certain cases it might be desirable to contact a gas with only a small amount of solid surface, as in the case of gas reactions in the presence of a very active catalyst, or it might be necessary to minimize the duration of contact between the gas and solid. In such cases, operation in a dilute-phase transport reactor proves advantageous. In such reactors, gas velocities are typically very high (50 ft/sec, say) and the solid loading is correspondingly low, often below 1 lb/ft^3. Temperature is nonuniform if a large heat effect is present.

It is also important to appreciate the distinction between fast fluidization and the riser reactor used in modern petroleum catalytic cracking art.[8] Gas velocity in the riser increases with height, but typically ranges in excess of 35 ft/sec in the upper portions of the riser. Solid rates through commercial risers range from 100 to as much as 300 lb/ft^2-sec and solid loadings are in the neighborhood of 10 lb/ft^3. But the demixing[14] of the gas and catalyst in the riser reactor is gross in scale by comparison with the fine-scale demixing of the fast bed condition. A probe traversing the fast bed would pass every inch or so from a lean void into a strand of solid or vice versa. In contrast, density contours across the riser reactor reveal extreme segregation; high density zones, most likely representing solid flowing downwards, appear along the walls, while the core remains dilute.[(8)] The riser reactor, unlike fast fluidization, does not appear as a natural extension of bubbling and turbulent fluidization. It will be interesting to explore the way in which the gross demixing of the riser reactor evolves from fast fluidization upon increase in gas velocity.

VI. CONCLUSIONS

This study leads to the following conclusions:

1. As the velocity of the gas flowing across a bubbling fluidized bed is raised, the heterogeneous, two-phase character of the bed peaks, and then gradually changes giving way to a condition of increasing uniformity culminating in the turbulent state in which large bubbles or voids on the whole are absent.

2. The transition from bubbling to turbulent fluidization is reflected in the fluctuations of both the dynamic pressure at any point in the bed and the pressure drop across the bed.

The transition can be characterized by two velocities: U_c, the velocity at which the fluctuations begin to diminish from their peak value, and U_k, the velocity at which the fluctuations have decayed to a sufficiently low level and simultaneously begun to level off. U_k marks the onset of the turbulent regime.

3. The range of gas velocities, $U_c - U_k$, over which the transition occurs, increases with particle size and density.

4. Both U_c and U_k generally increase with particle size and density.

5. The ratios U_c/U_T and U_k/U_T (where U_T is the terminal velocity of the median particle) approach or even exceed 10 for the beds of fine powders, while they diminish with particle density and size, and could become less than unity for beds of coarse solids. The large ratios for the fine powders provide an indication of the tendency of fine powders to form clusters. The behavior of a turbulent bed of fines is accordingly governed by the interaction of the gas not only with individual particles but also, and perhaps mostly, with relatively large clusters of particles.

6. In a turbulent fluidized bed, an upper bed level is present (though it is more diffused than that of a bubbling fluidized bed) owing to greater particle carryover rates. As in the bubbling bed, the fluidized density in a turbulent bed depends primarily on the gas velocity and only slightly on the rate at which the solid may be fed or withdrawn from the bed.

7. The turbulent regime for a fine solid extends to the transport velocity which marks the onset of concurrent, transport flow of gas and solid. The transport velocity can also be viewed as the point where the terminal velocity of the effective cluster which characterizes the structure of the turbulent bed becomes equal to the velocity of the gas.

8. The fast bed condition is typically obtained at gas velocities beyond the transport velocity, and at sufficiently high solid rates. The fast fluidized bed is essentially an entrained, dense suspension characterized by aggregation of the particles in clusters and strands, extensive backmixing of solid, and slip velocities that are an order of magnitude greater than the free fall velocity of the individual particles.

9. The contribution of wall-solid friction to the pressure drop measured in the 3-inch fast bed was always less than 10% over a fairly wide range of conditions.

10. Advantages of high velocity fluidized beds of fine solids. High velocity fluidized beds (that is, the turbulent and fast beds) embody all the qualities of bubbling fluidization: e.g., uniform temperature, capability of bringing a cold solid, gas or liquid feed almost instantaneously to bed temperature, high heat transfer rates to immersed surfaces,[20] etc. But high velocity fluidized beds also afford several potential advantages over the bubbling fluidized bed. Chiefly, higher processing capacities per unit

volume of reactor space, a more intimate contact between gas and solid, a lesser degree of gas backmixing in the turbulent bed, and no gas backmixing in the fast bed,[21] and a capability to receive and handle cohesive solids or particles which have a tendency to agglomerate. High velocity fluidized bed, especially the turbulent bed, may be easier to scaleup than a bubbling fluidized bed.

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TABLE 1 - EXPERIMENTAL SOLIDS

Solid	Symbol Used	Size Range (Microns)	$\bar{d}$ (Microns)	ρ_s (g/cm^3)	ϕ_s	Additional Comments
Dicalite 4200	A	0-160	33	1.67	0.4	Dicatomaceus earth Grefco, Inc., Los Angeles, California
Fluid Cracking Catalyst (FCC)	B	0-130	49	1.07	1.0	Slica-alumina Catalyst. W. R. Grace & Co., Baltimore, Maryland
HFZ-20	C	0-130	49	1.45	1.0	Slica-alumina catalyst. Houdry Division, Air Products & Chemicals, Wayne, Pennsylvania
Hydrated alumina	D	40-200	103	2.46	1.0	Hydrated alumina, H-30. Kaiser Aluminum & Chemical Corp., Baton Rouge, Louisiana
No. 85 Sand	E	80-670	268	2.65	0.82	
No. 2024 Glass	F	105-210	157	2.42	1.0	

Table 2. Summary of Turbulent Transition Velocities

Solid	$\bar{d}$ (microns)	ρ_s (g/cm^3)	$(\rho_s \bar{d})^{1/2}$ (g/cm^3 x μm)$^{1/2}$	ϕ_s	U_T (ft/sec)	U_c (ft/sec)	$\left(\frac{U_c}{U_T}\right)$	U_k (ft/sec)	$\left(\frac{U_k}{U_T}\right)$
A(6" bed)	33	1.67	7.42	0.4	0.0745	1.74	23.4	3.51	47.1
B(6" bed)	49	1.07	7.24	1.0	0.255	2.00	7.84	2.00	7.84
C(6" bed)	49	1.45	8.43	1.0	0.347	2.98	8.58	4.49	12.9
C(2D bed)	49	1.45	8.43	1.0	0.347	2.00	5.75	3.51	10.9
D(6" bed)	103	2.46	15.9	1.0	1.91	4.40	2.10	9.00	4.71
D(2D bed)	103	2.46	15.9	1.0	1.91	4.49	2.35	8.33	4.38
E(6" bed)	268	2.65	26.6	0.82	3.81	9.00	2.36	18.0	4.74
E(2D bed)	268	2.65	26.6	0.82	3.81	4.98-7.02	1.31-1.84	Uncertain	--
F(2D bed)	157	2.42	19.5	1.00	3.12	4.98-7.02	1.59-2.24	Uncertain	--

Table 3. The contribution of wall-solid friction to the pressure drop across the test section. Experimental results.

Series	Gas Velocity (ft/sec)	Solid rate; lb/ft^2-sec	Number of runs	Mean $\Delta P/\Delta L$; lb/ft^3	Mean X%	Standard Deviation (%)
1	5.8	4.7	68	10.1	7.09	3.05
2	5.8	10.1	72	12.8	6.98	2.95
3	8.0	11.5	100	9.0	6.28	3.63
4	14.8	11.4	69	1.1	6.62	1.56
5	14.8	18.8	33	3.5	7.34	2.77

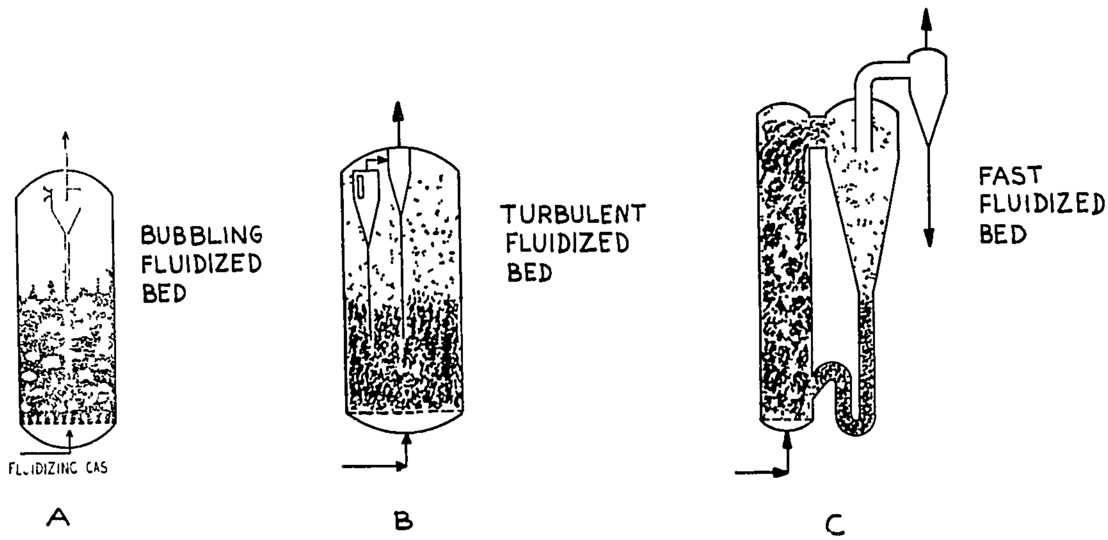


Figure 1
Bubbling, Turbulent and Fast Fluidized Beds

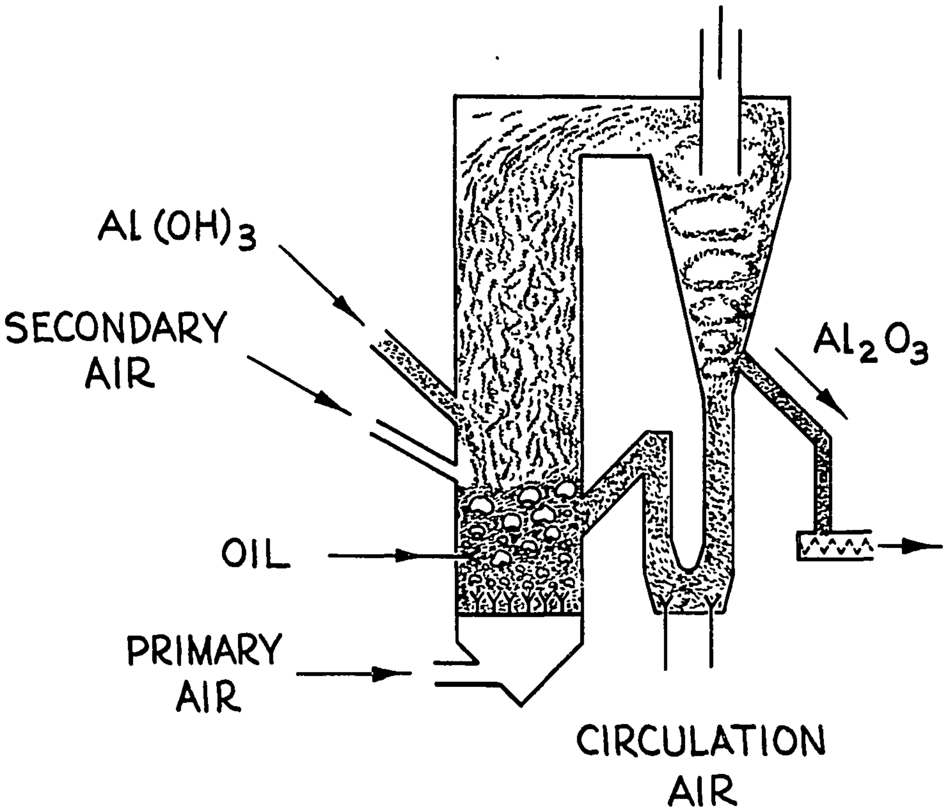


Figure 2
Schematic of Lurgi's Fast Bed Calciner

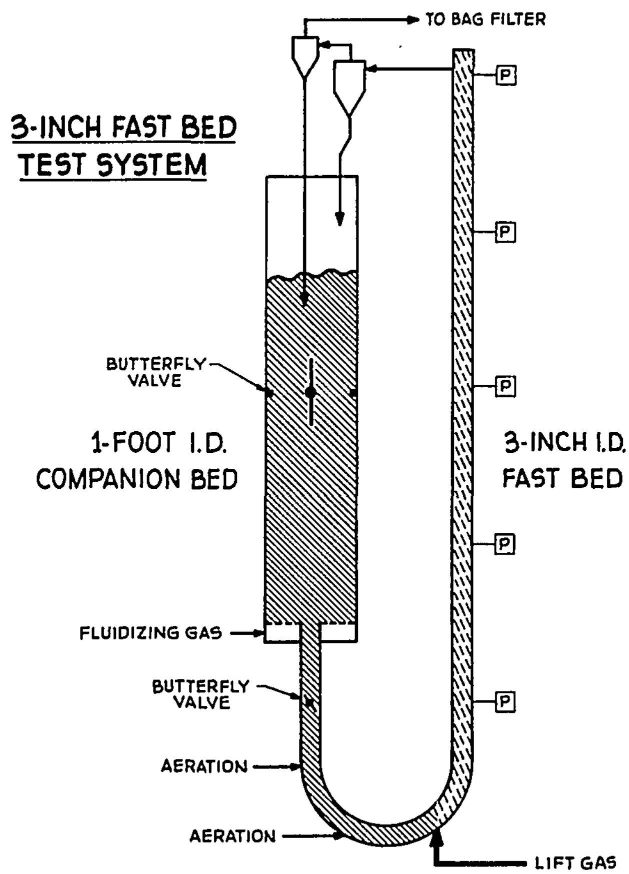


Figure 3
Schematic of the 3-Inch System

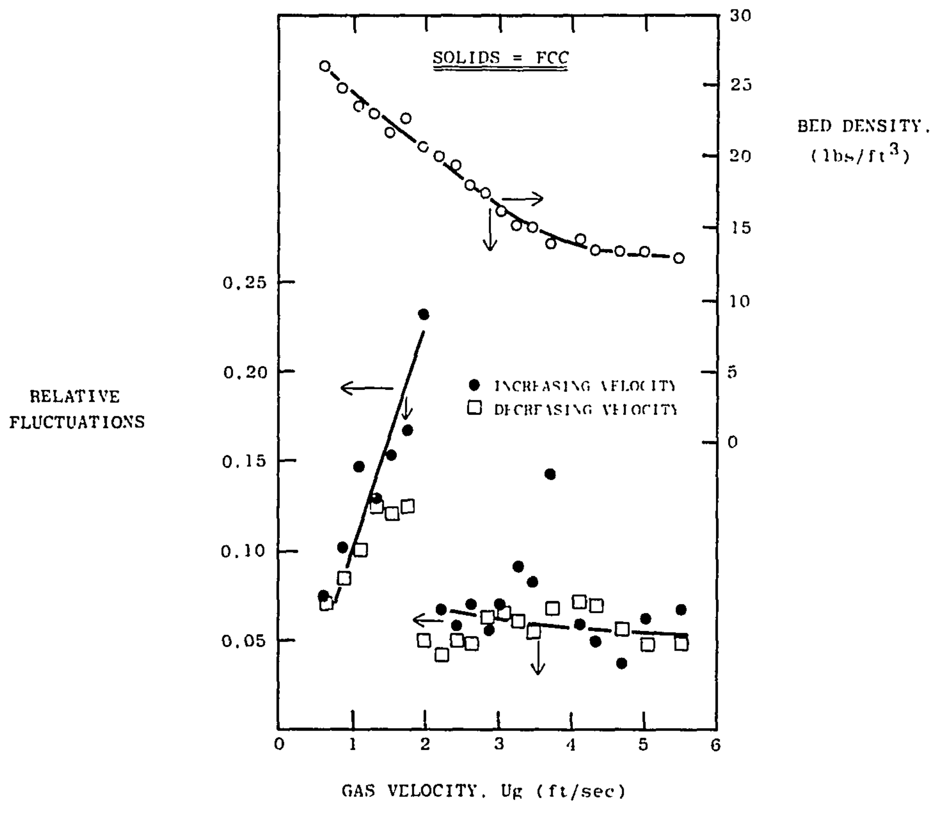


Figure 4
Pressure Fluctuations Divided by the Mean Pressure Drop Across a Bed of FCC (Solid B) vs. Superficial Gas Velocity. The upper curve indicates the bed density. The experiments were conducted in the 6-inch bed.

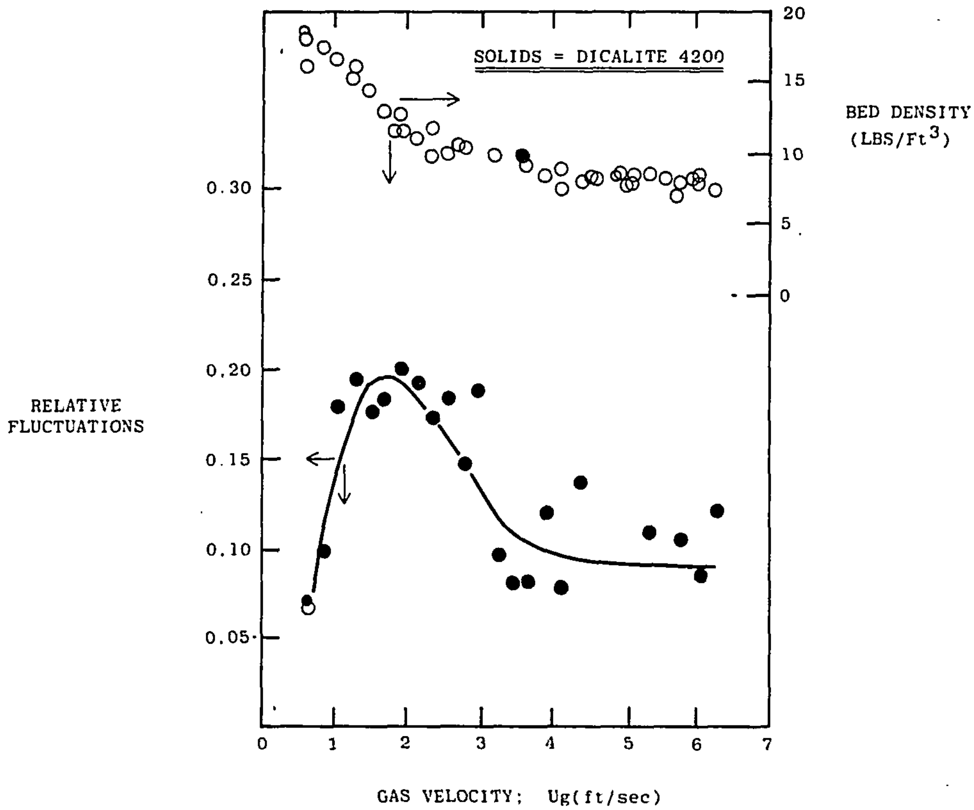


Figure 5
Pressure Fluctuations Divided by the Mean Pressure Drop Across a Bed of Dicalite 4200 (Solid A) vs. Superficial Gas Velocity. The upper curve indicates the bed density. The experiments were conducted in the 6-inch bed.

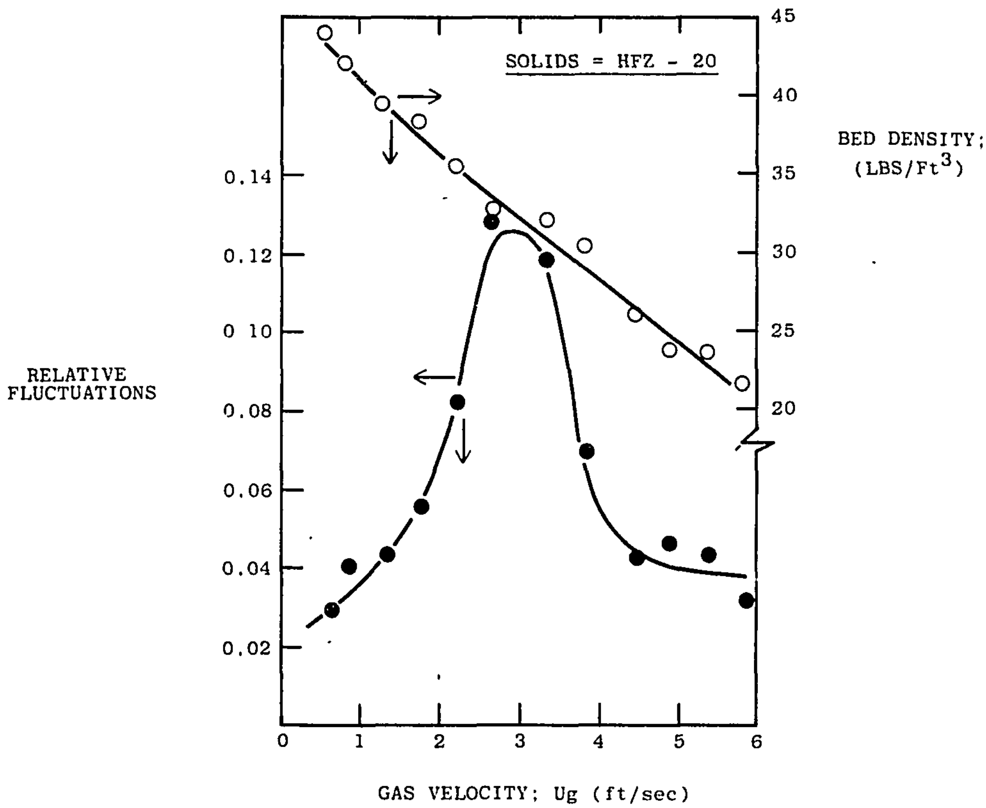


Figure 6
Pressure Fluctuations Divided by the Mean Pressure Drop Across a Bed of HFZ-20 (Solid C) vs. Superficial Gas Velocity. The upper curve indicates the bed density. The experiments were conducted in the 6-inch bed.

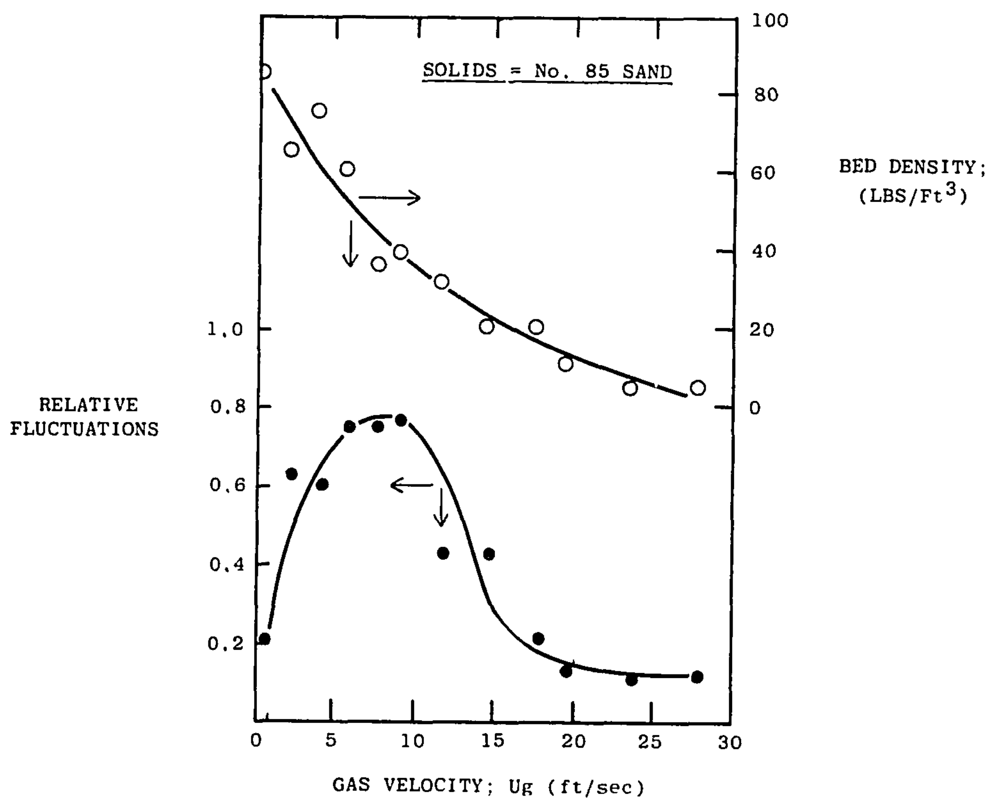
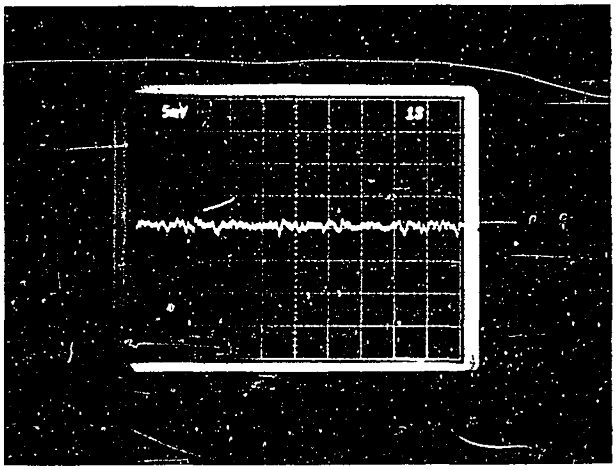
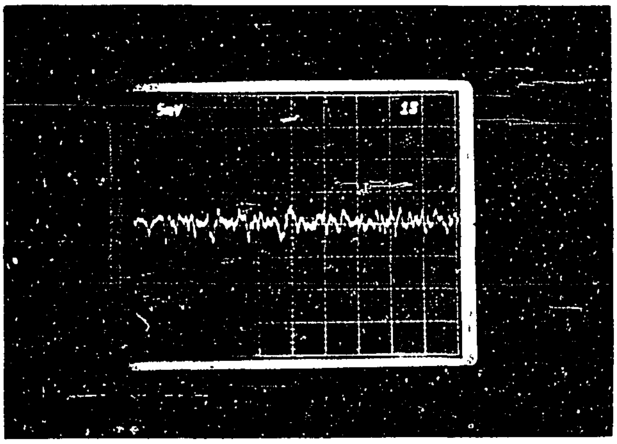


Figure 7
Pressure Fluctuations Divided by the Mean Pressure Drop Across a Bed of No. 85 Sand (Solid E) vs. Superficial Gas Velocity. The upper curve indicates the bed density. The experiments were conducted in the 6-inch bed.

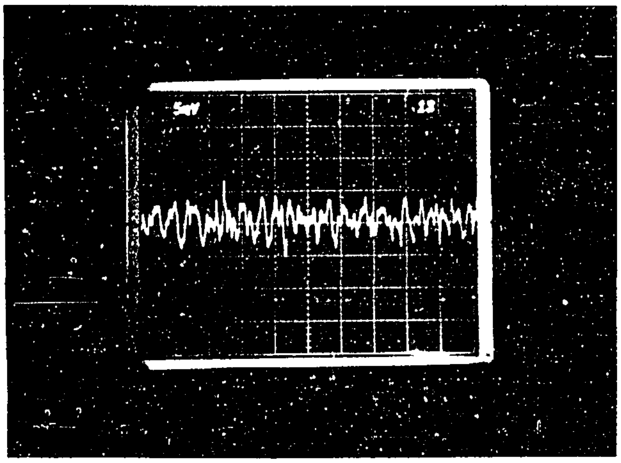


Superficial Gas Velocity = 0.25 ft/sec
Bed Density = 47.9 lb/ft^3

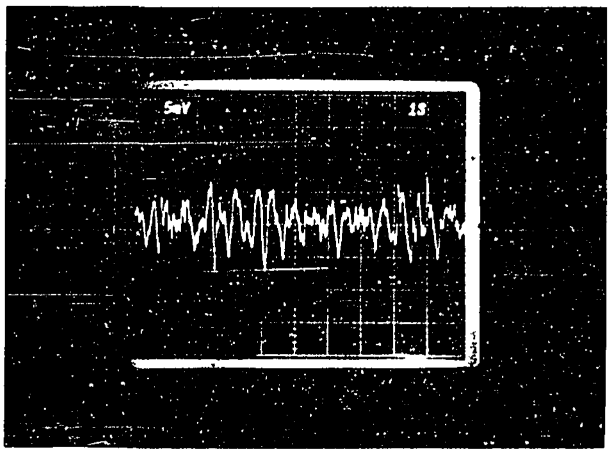


Superficial Gas Velocity = 0.51 ft/sec
Bed Density = 45.0 lb/ft^3

Figure 8
Traces of the Dynamic Pressure Fluctuations Measured by a High Sensitivity Pressure Transducer in the Two-Dimensional Bed Using HFZ-20 (Solid C). Superficial gas velocities in ft/sec are indicated below each tracing. Each square of vertical axis equals five millivolts.

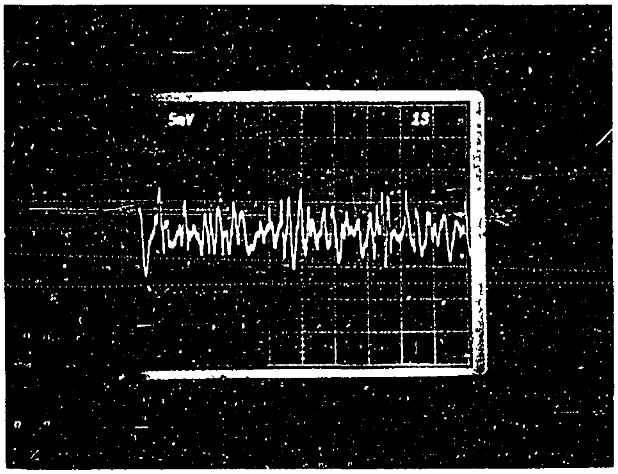


Superficial Gas Velocity = 1.1 ft/sec
Bed Density = 44.2 lb/ft^3

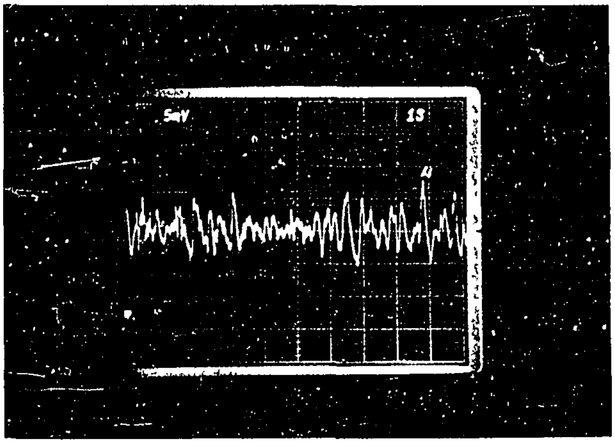


Superficial Gas Velocity = 1.7 ft/sec
Bed Density = 43.4 lb/ft^3

Figure 8 (cont'd)

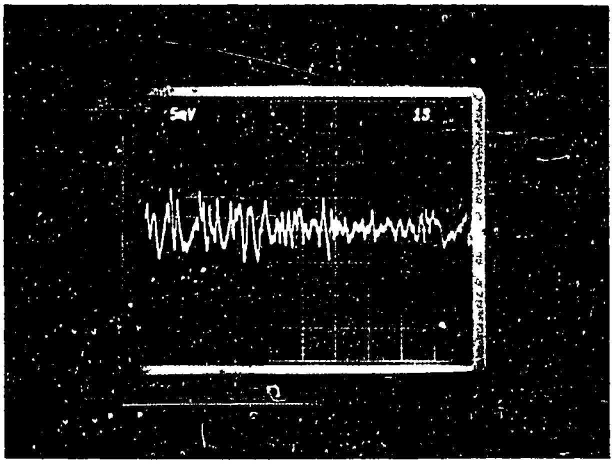


Superficial Gas Velocity = 2.2 ft/sec
Bed Density = 39.7 lb/ft^3

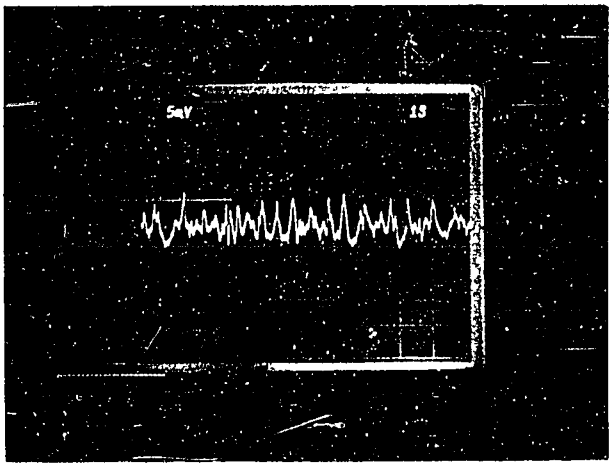


Superficial Gas Velocity = 2.7 ft/sec
Bed Density = 37.6 lb/ft^3

Figure 8 (cont'd)

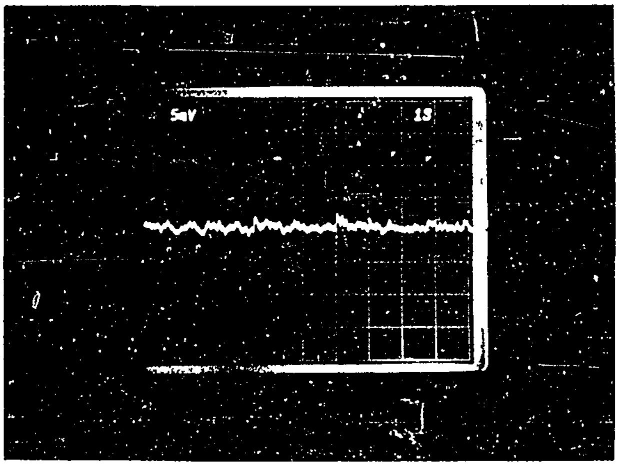


Superficial Gas Velocity = 3.1 ft/sec
Bed Density = 39.1 lb/ft^3

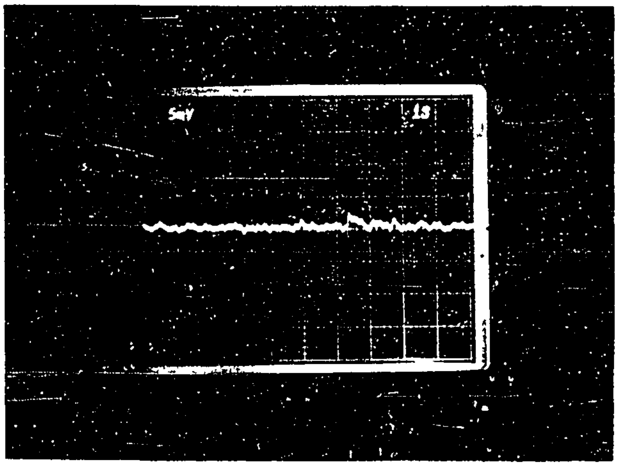


Superficial Gas Velocity = 3.6 ft/sec
Bed Density = 34.6 lb/ft^3

Figure 8 (cont'd)

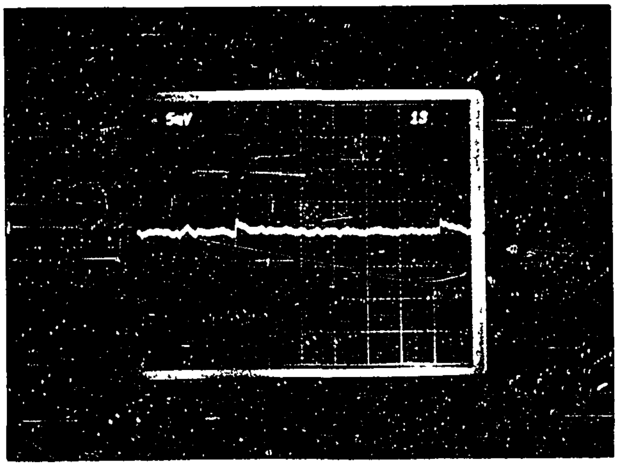


Superficial Gas Velocity = 4.1 ft/sec
Bed Density = 24.2 lb/ft^3

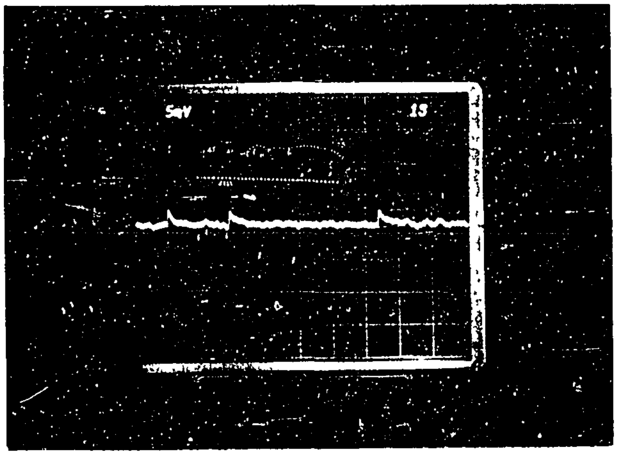


Superficial Gas Velocity = 4.6 ft/sec
Bed Density = 21.2 lb/ft^3

Figure 8 (cont'd)



Superficial Gas Velocity = 5.2 ft/sec
Bed Density = 14.6 lb/ft^3



Superficial Gas Velocity = 5.6 ft/sec
Bed Density = 16.4 lb/ft^3

Figure 8 (cont'd)

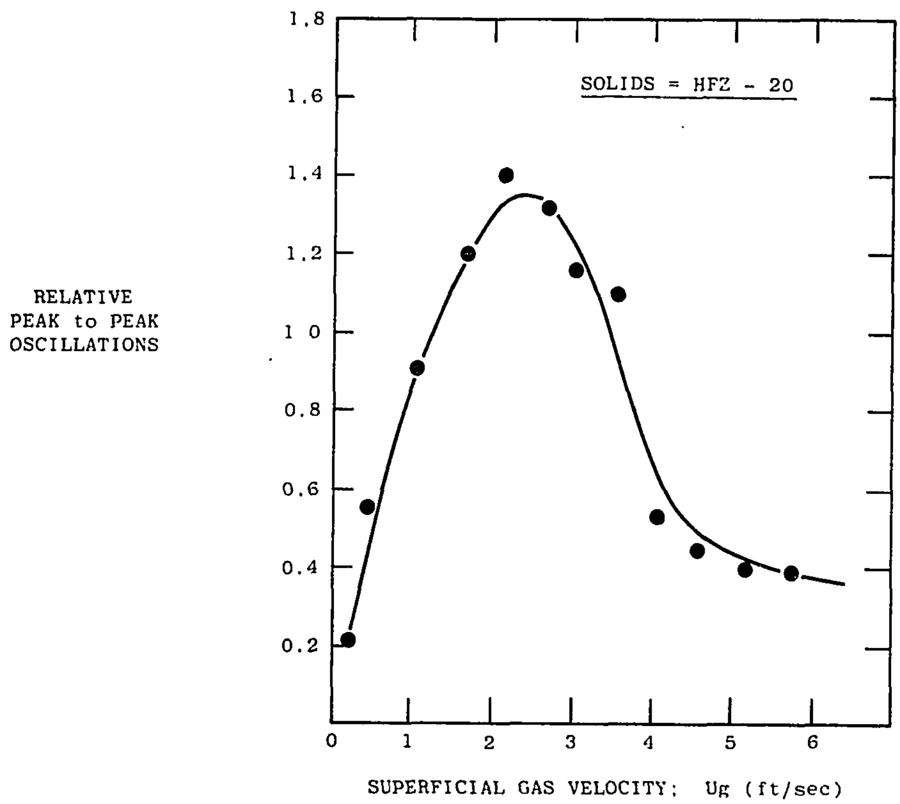


Figure 9
Ratio of the Mean Peak to Peak Oscillations (from Figure 8) to the Bed Density as a Function of the Superficial Gas Velocity. Solid is HFZ-20 (Solid C).

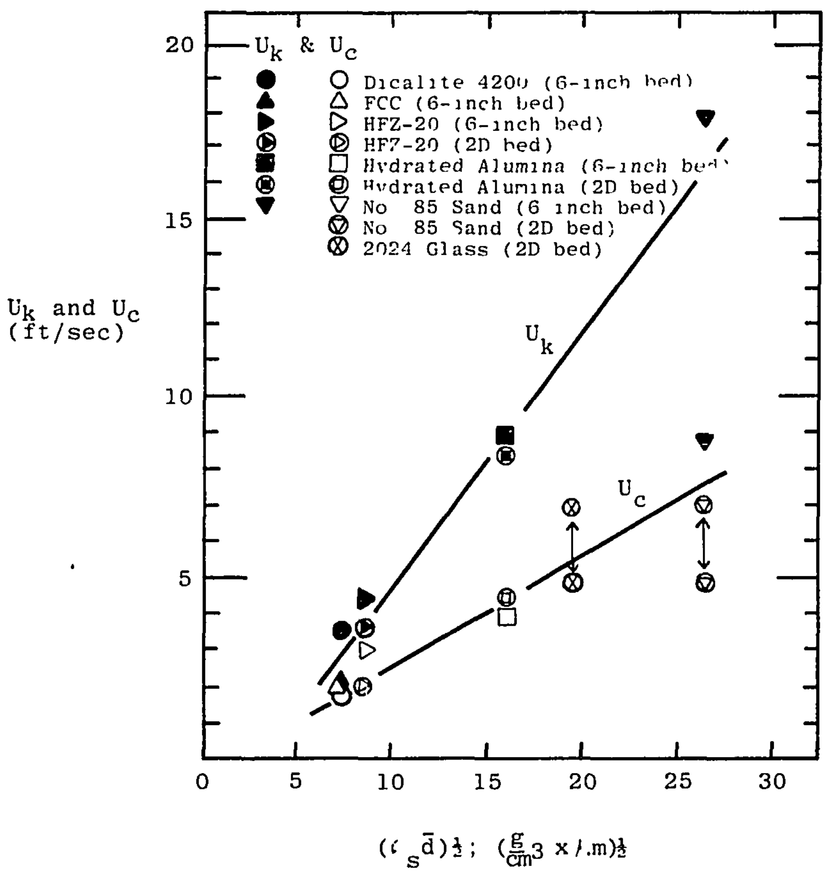


Figure 10
U_k and U_c vs. $(\rho_s \bar{d})^{1/2}$

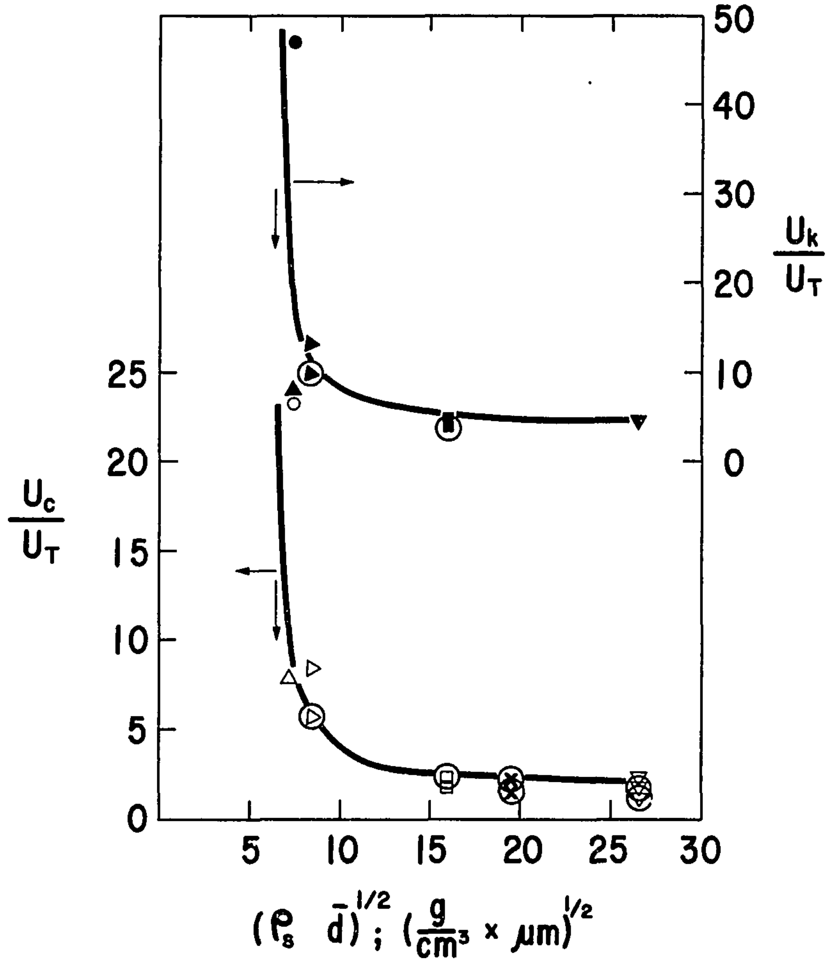


Figure 11
U_c/U_T and U_k/U_T vs. $(\rho_s \bar{d})^{1/2}$

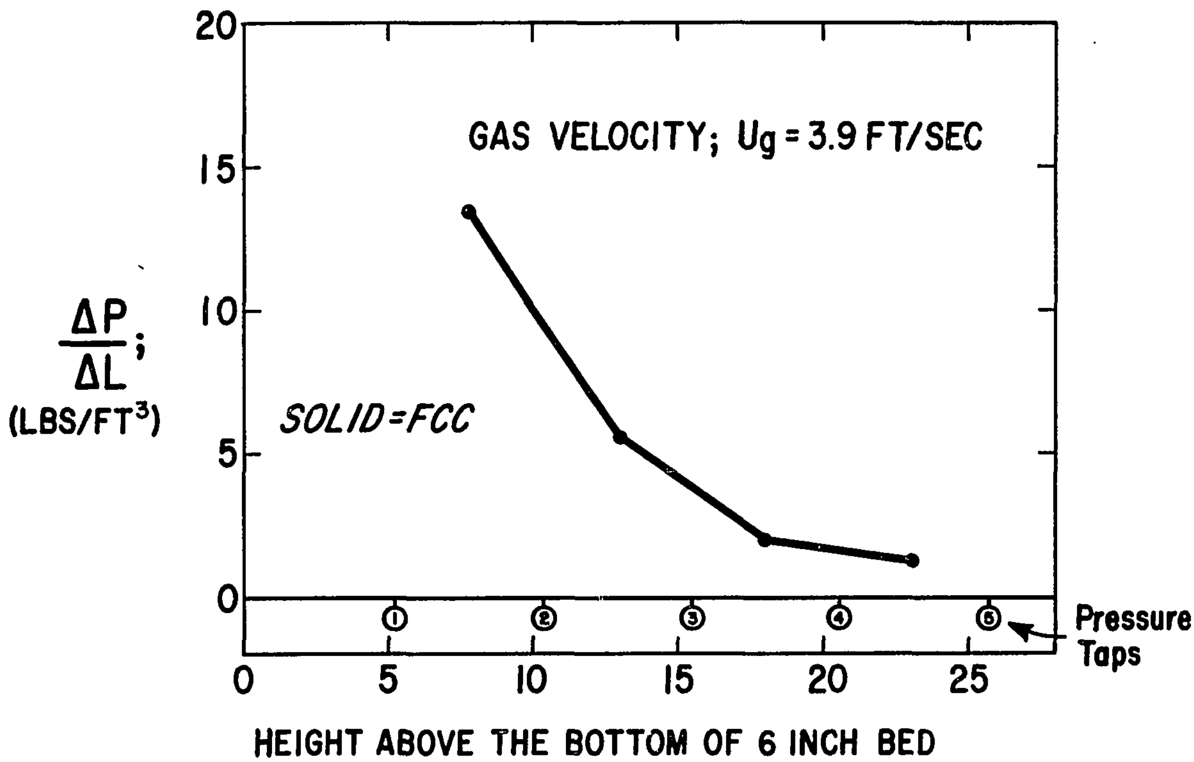


Figure 12
Vertical Pressure Drop Profile Across the Six - Inch
Column in Which a Turbulent Bed of FCC (Solid B) is Maintained
at a Gas Velocity of 3.9 ft/sec

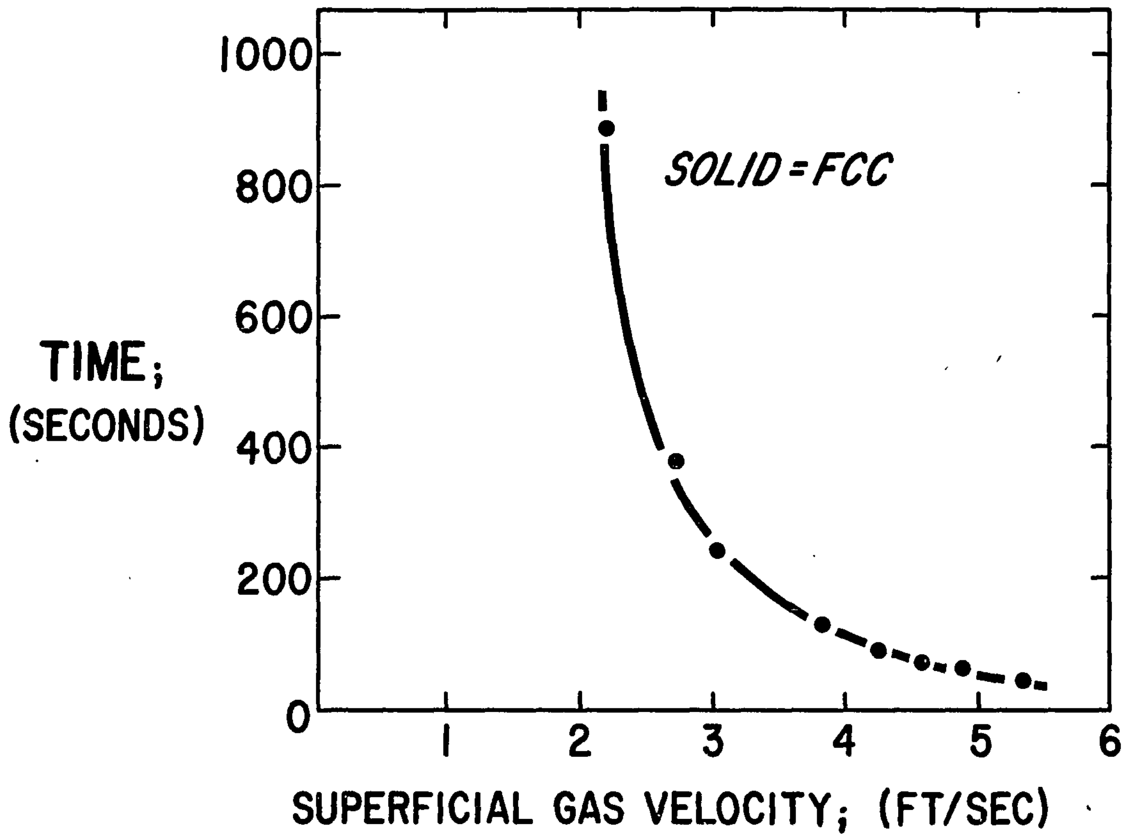


Figure 13
Time Required for $\rho_1 - \rho_2$ to Decrease to 1.0 Inches of Oil (sp. gr. = 0.827) vs. Superficial Gas Velocity in the 6-Inch Bed. Solid is FCC (Solid B).

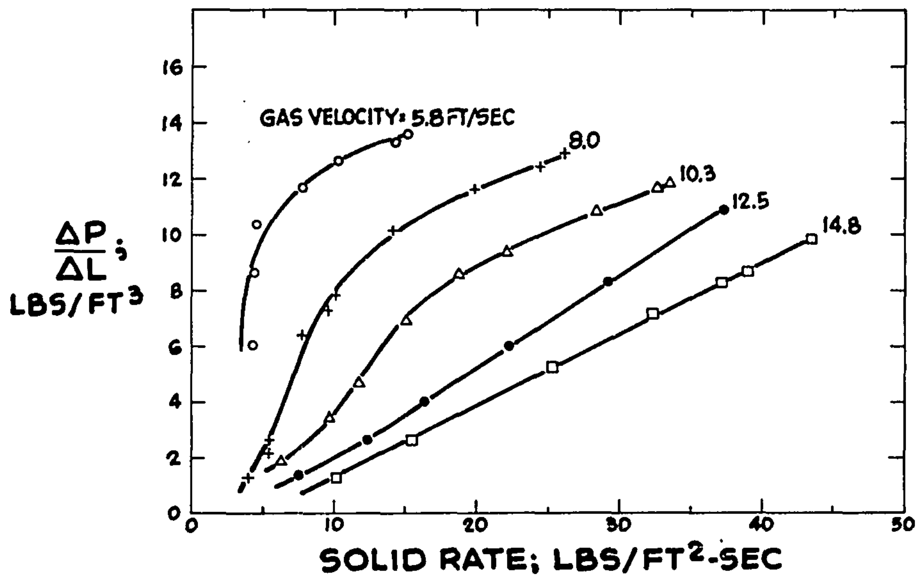


Figure 14
Pressure Gradient (across a section extending from 7.5 to 18.5 ft above the bottom of the 3-inch fast bed) vs. Solid Rate at Different Fluidizing Gas Velocities.

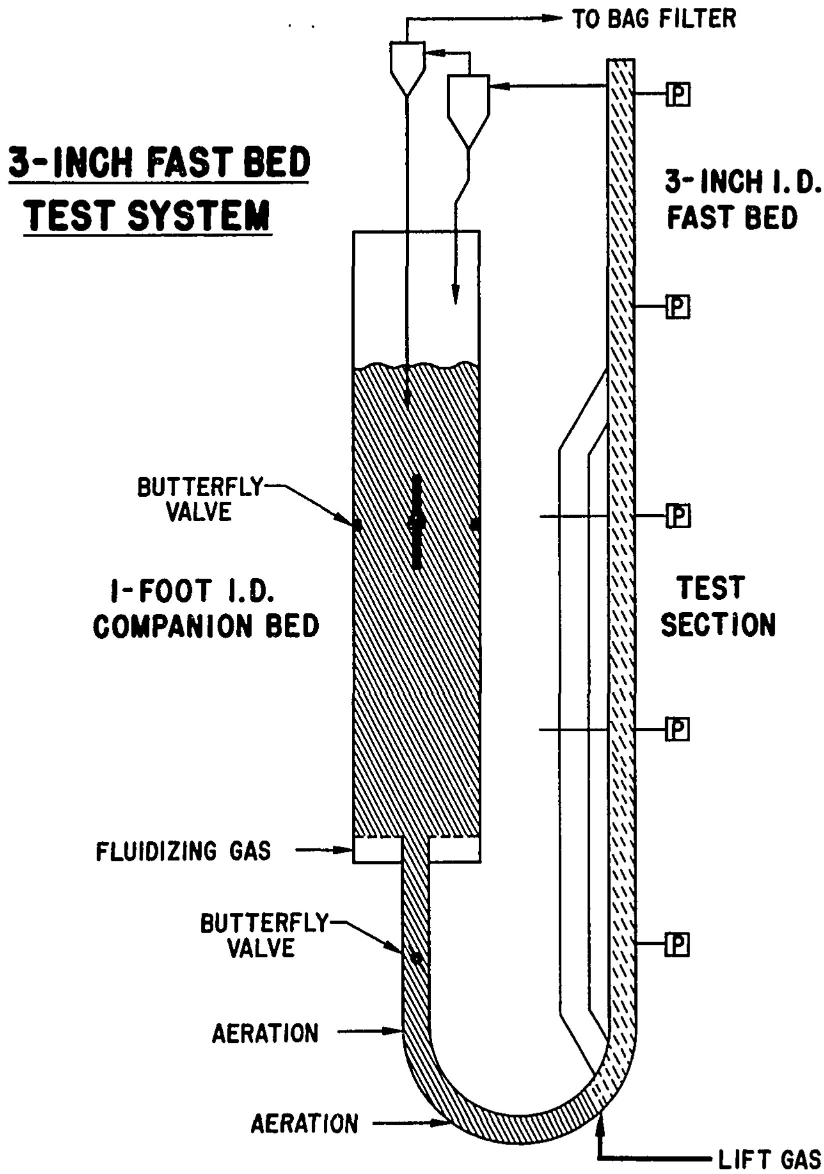


Figure 15
Schematic of the 3-Inch Fast Bed System and the
Modification Introduced to Study the Solid Hold Up

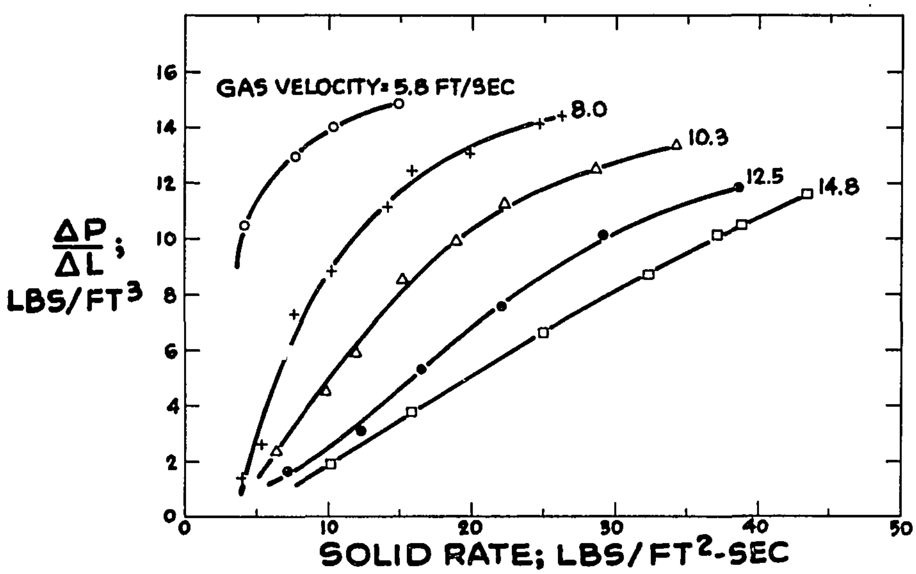


Figure 16
Pressure Gradient (across a section extending from 2 to 7.5 ft above the bottom of the 3-inch fast bed) vs. Solid Rate at Different Fluidizing Gas Velocities.

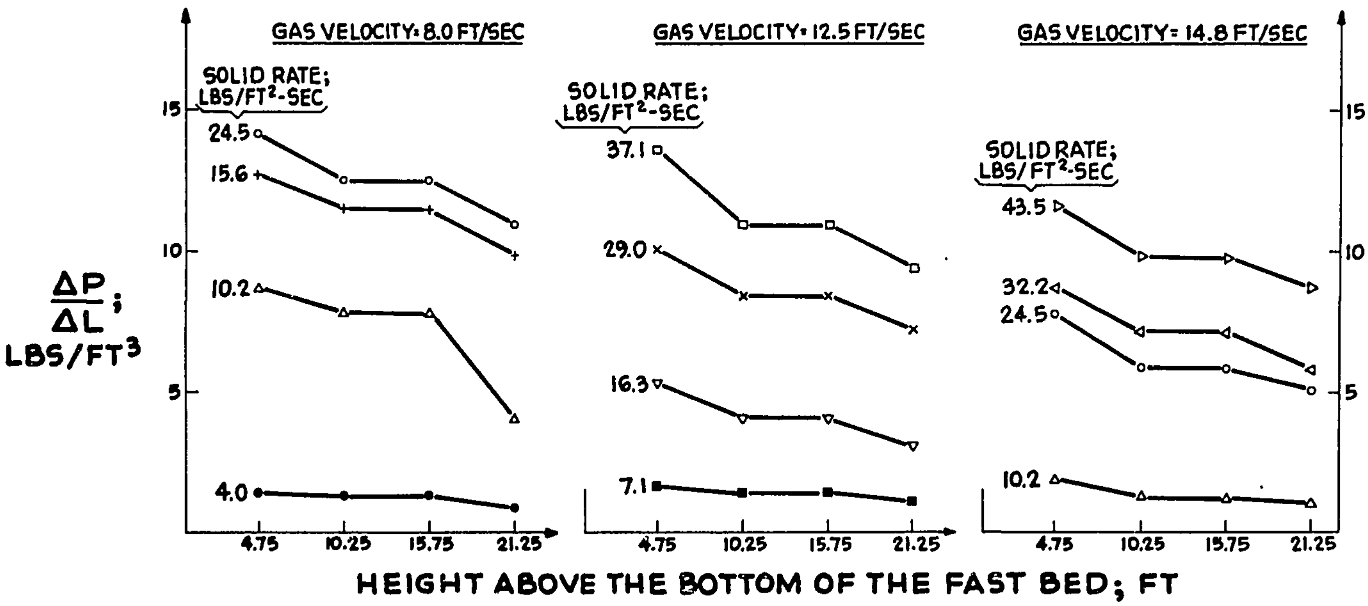


Figure 17
Pressure Gradient Profiles in the 3-Inch Fast Fluidized Bed

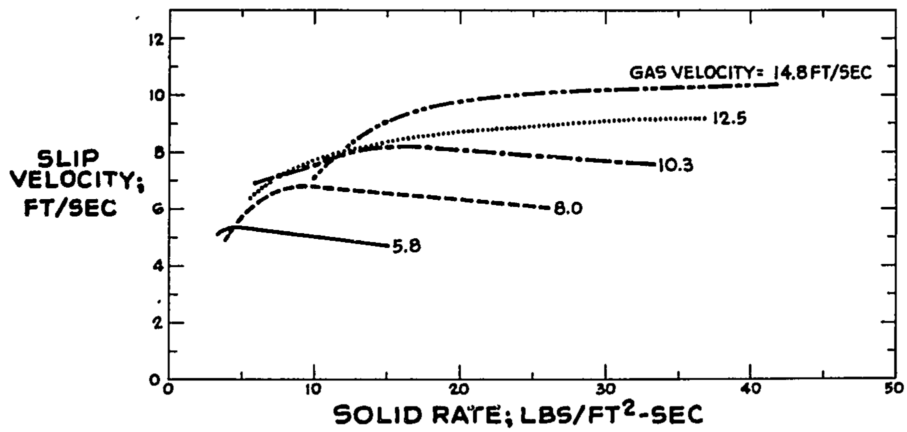


Figure 18
Slip Velocities Corresponding to the Data Shown in Figure 14

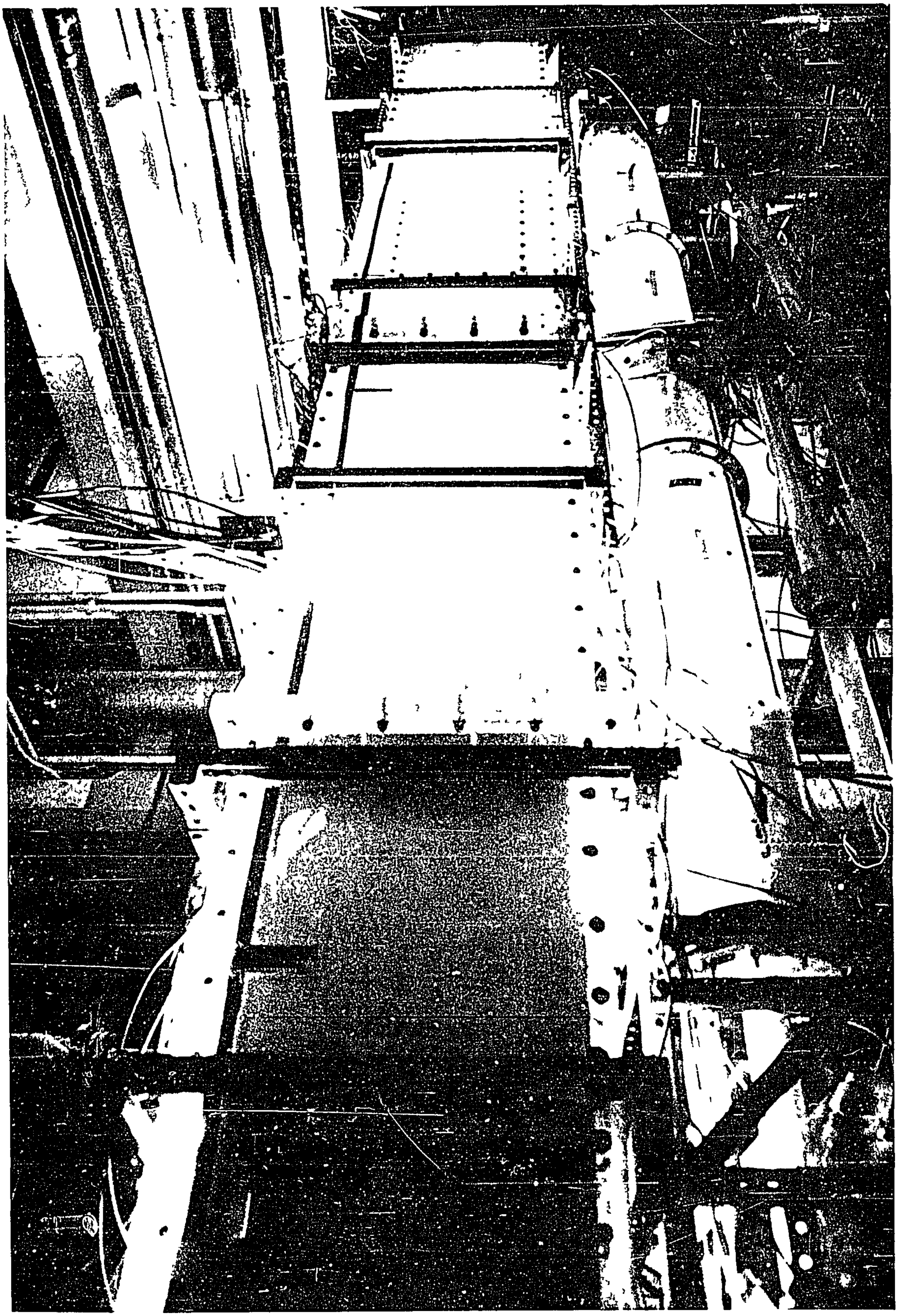


Exhibit 1
Bubbling 2-D Bed at a Superficial Gas Velocity of 0.51 ft/sec. Solid is HFZ-20 (Solid C).

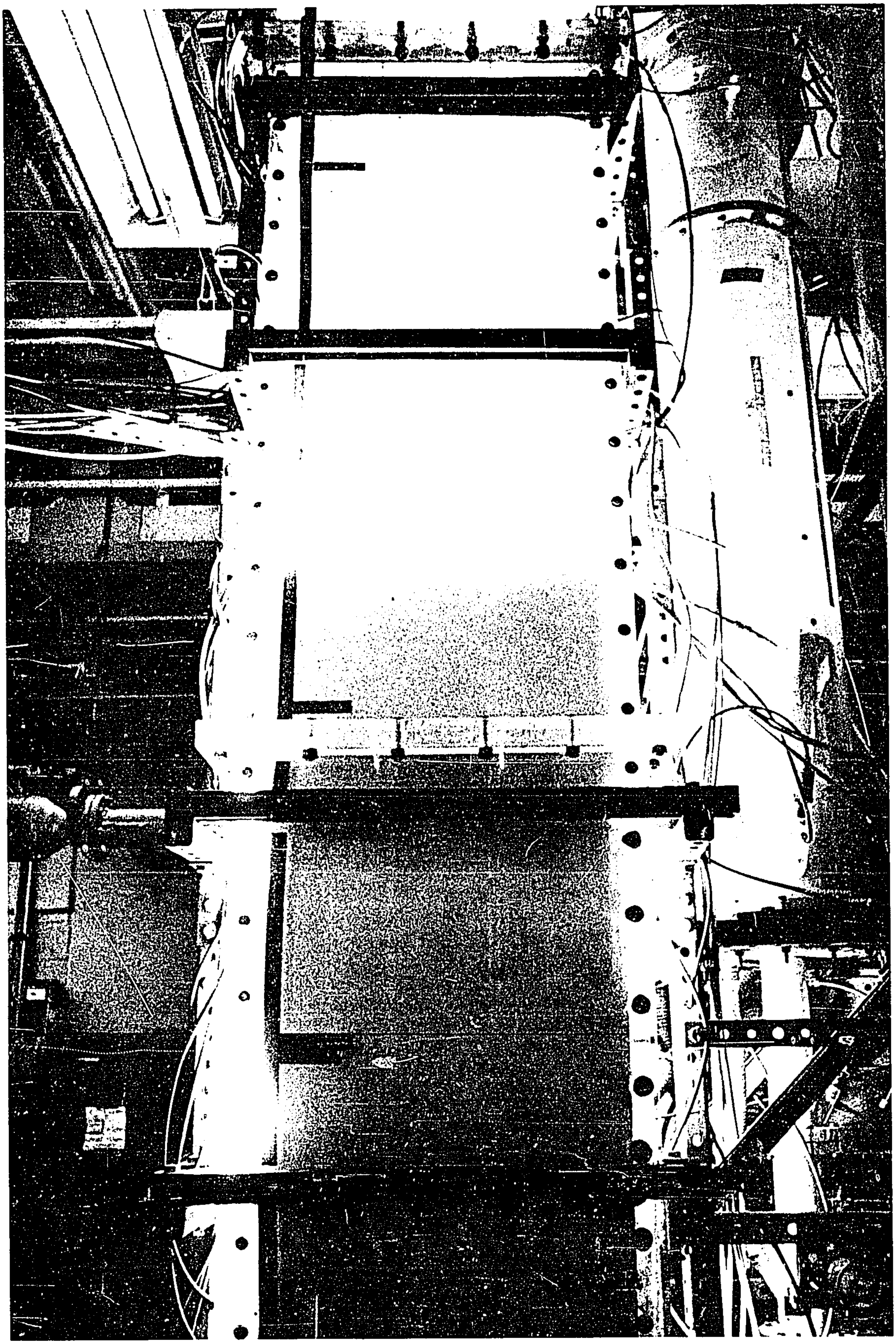


Exhibit 2
Bubbling 2-D Bed at a Superficial Gas Velocity of 0.51 ft/sec.
Solid is HFZ-20 (Solid C).

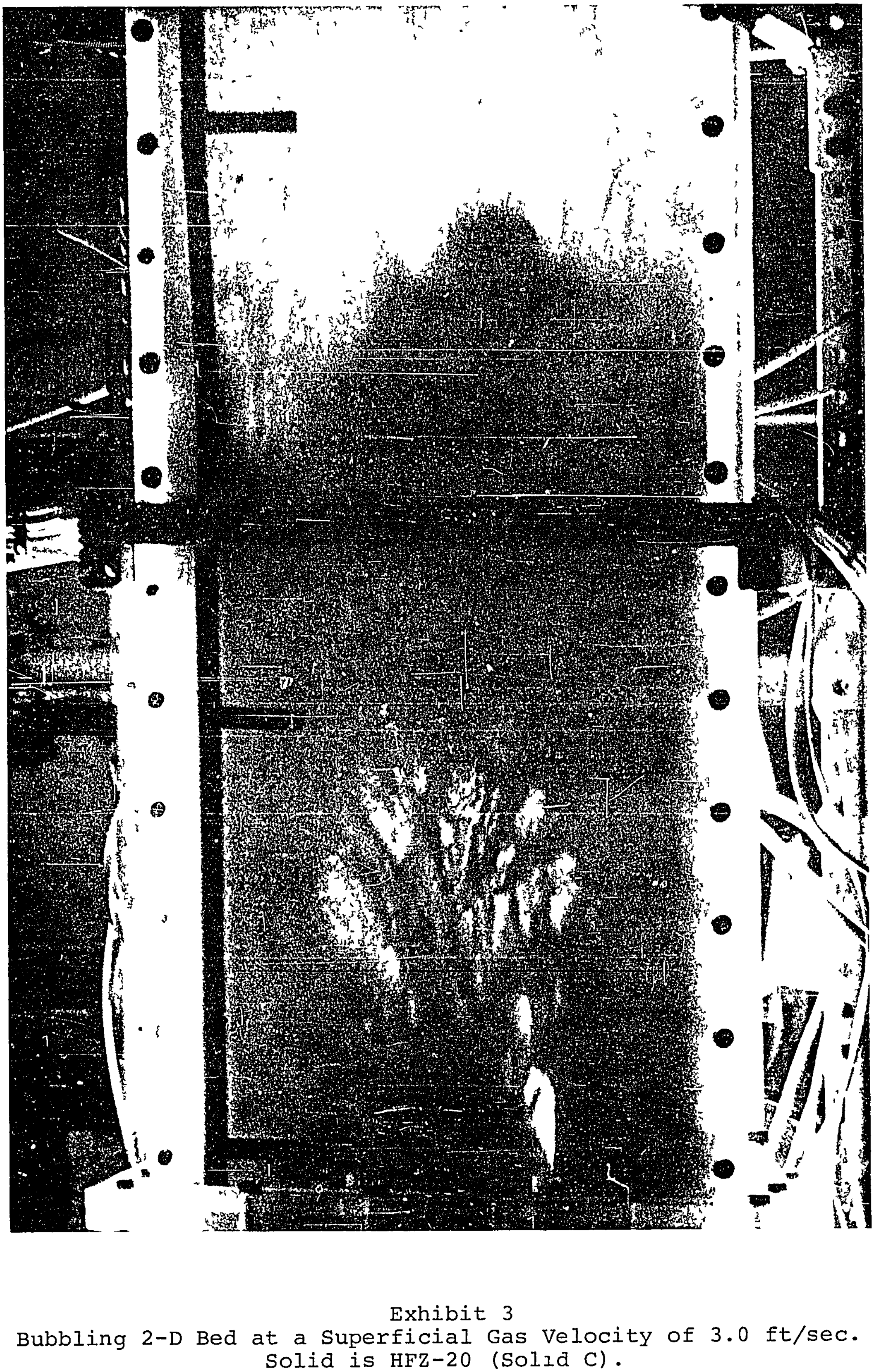


Exhibit 3
Bubbling 2-D Bed at a Superficial Gas Velocity of 3.0 ft/sec.
Solid is HFZ-20 (Solid C).

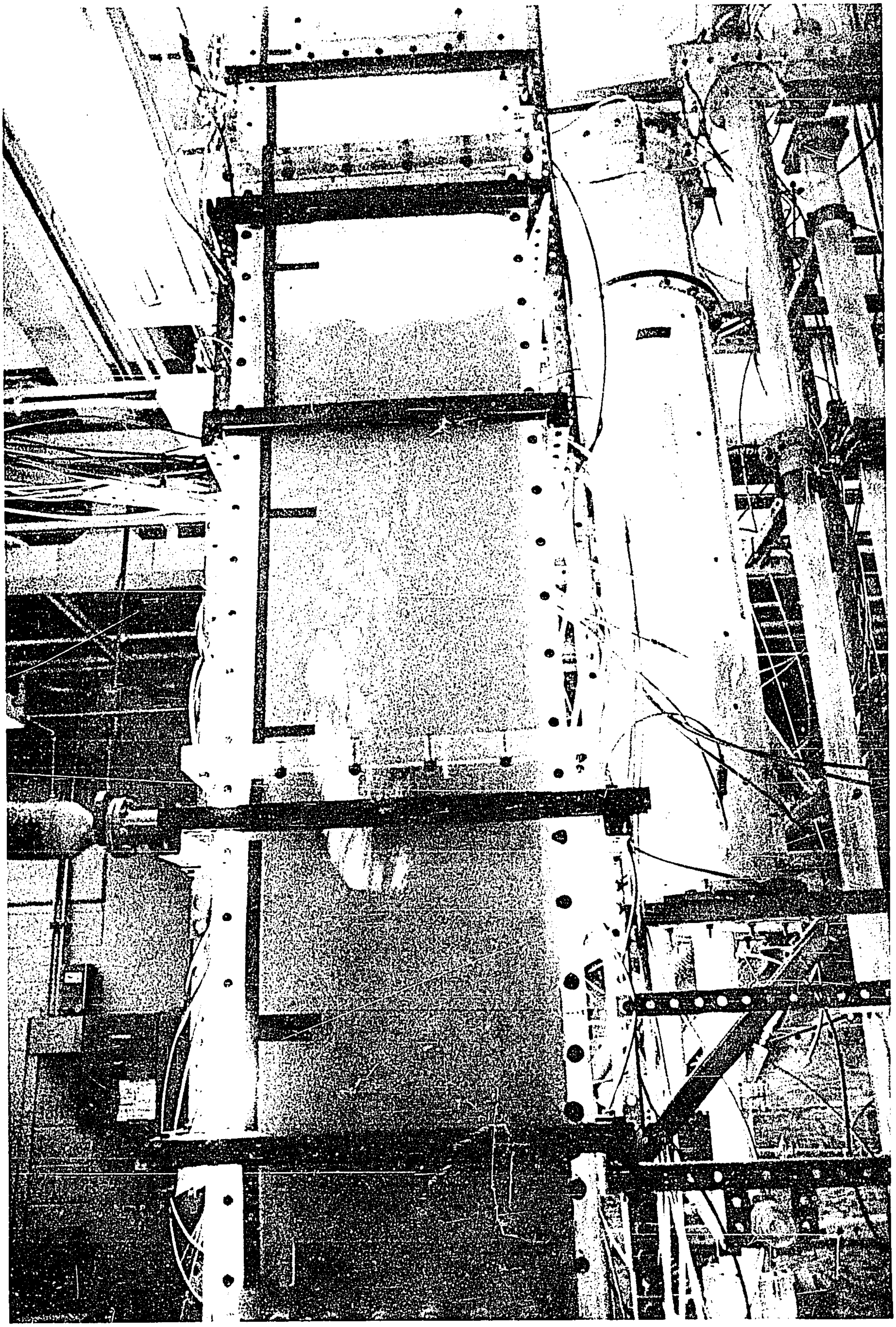


Exhibit 4
Bubbling 2-D Bed at a Superficial Gas Velocity of 3.0 ft/sec.
Solid is HFZ-20 (Solid C).

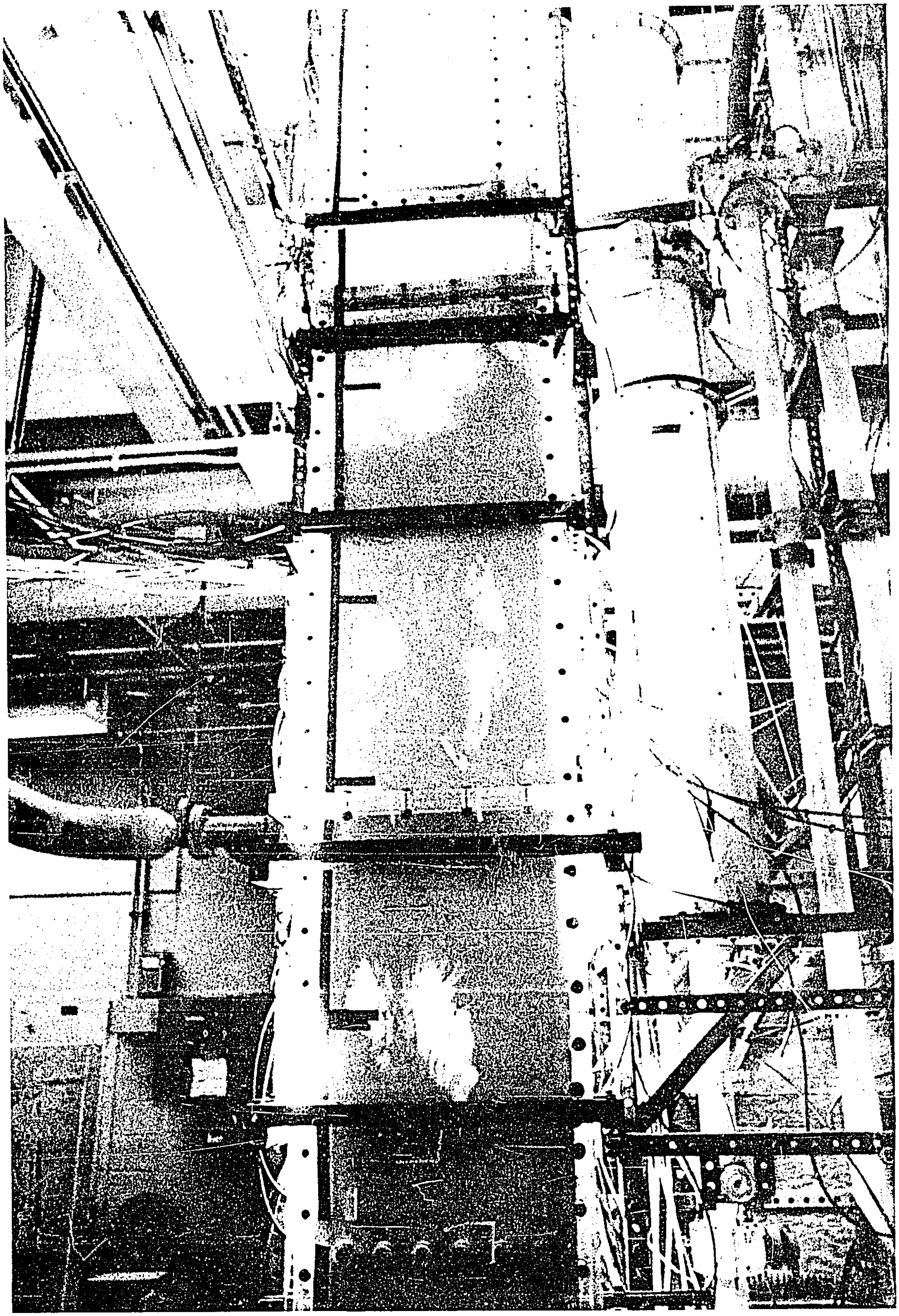


Exhibit 5
Bubbling 2-D Bed at a Superficial Gas Velocity of 3.0 ft/sec.
Solid is HFZ-20 (Solid C).

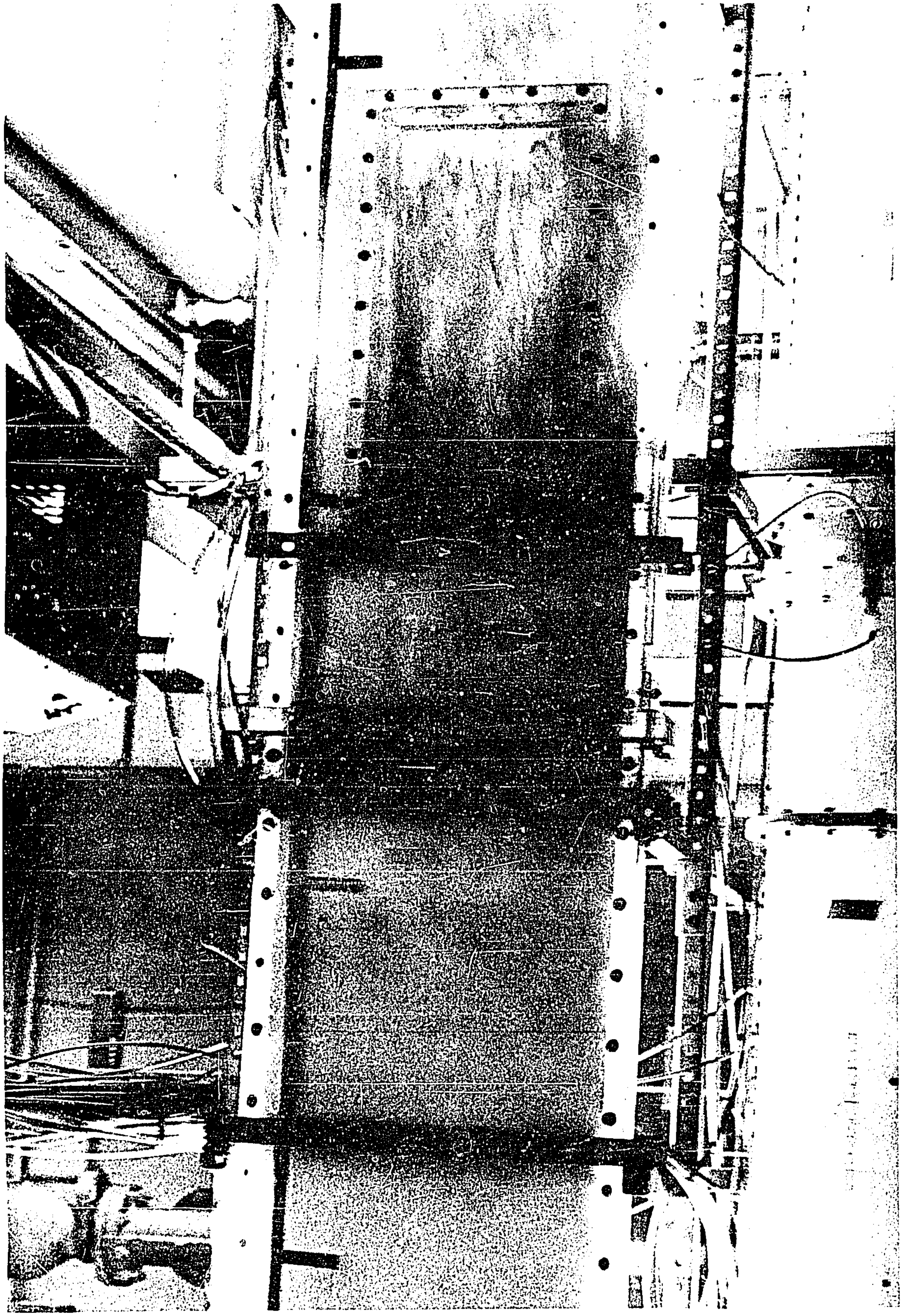


Exhibit 6
Turbulent 2-D Bed at a Superficial Gas Velocity of 3.9 ft/sec.
Solid is HFZ-20 (Solid C).

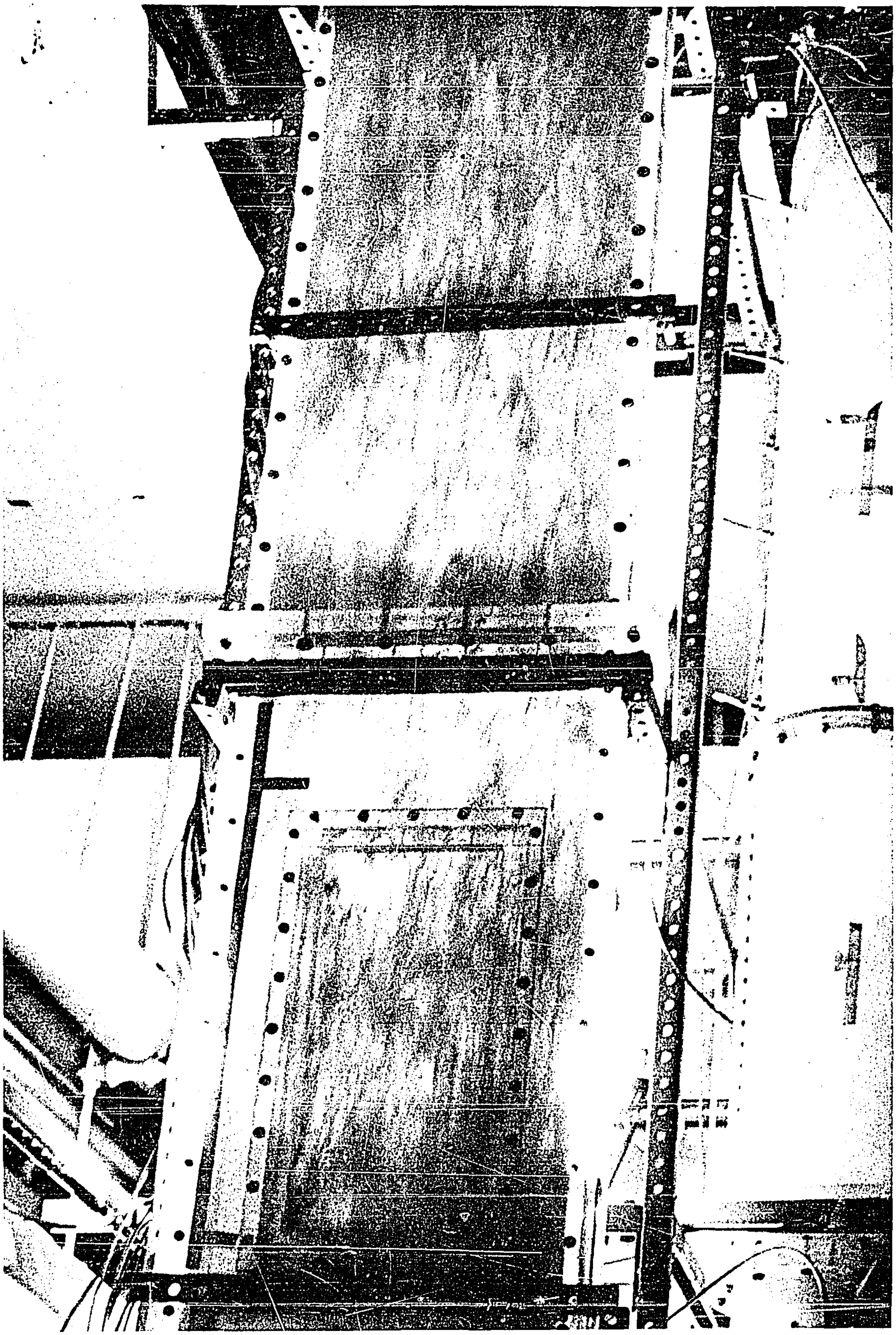


Exhibit 7
Turbulent 2-D Bed at a Superficial Gas Velocity of 3.9 ft/sec.
Solid is HFZ-20 (Solid C).

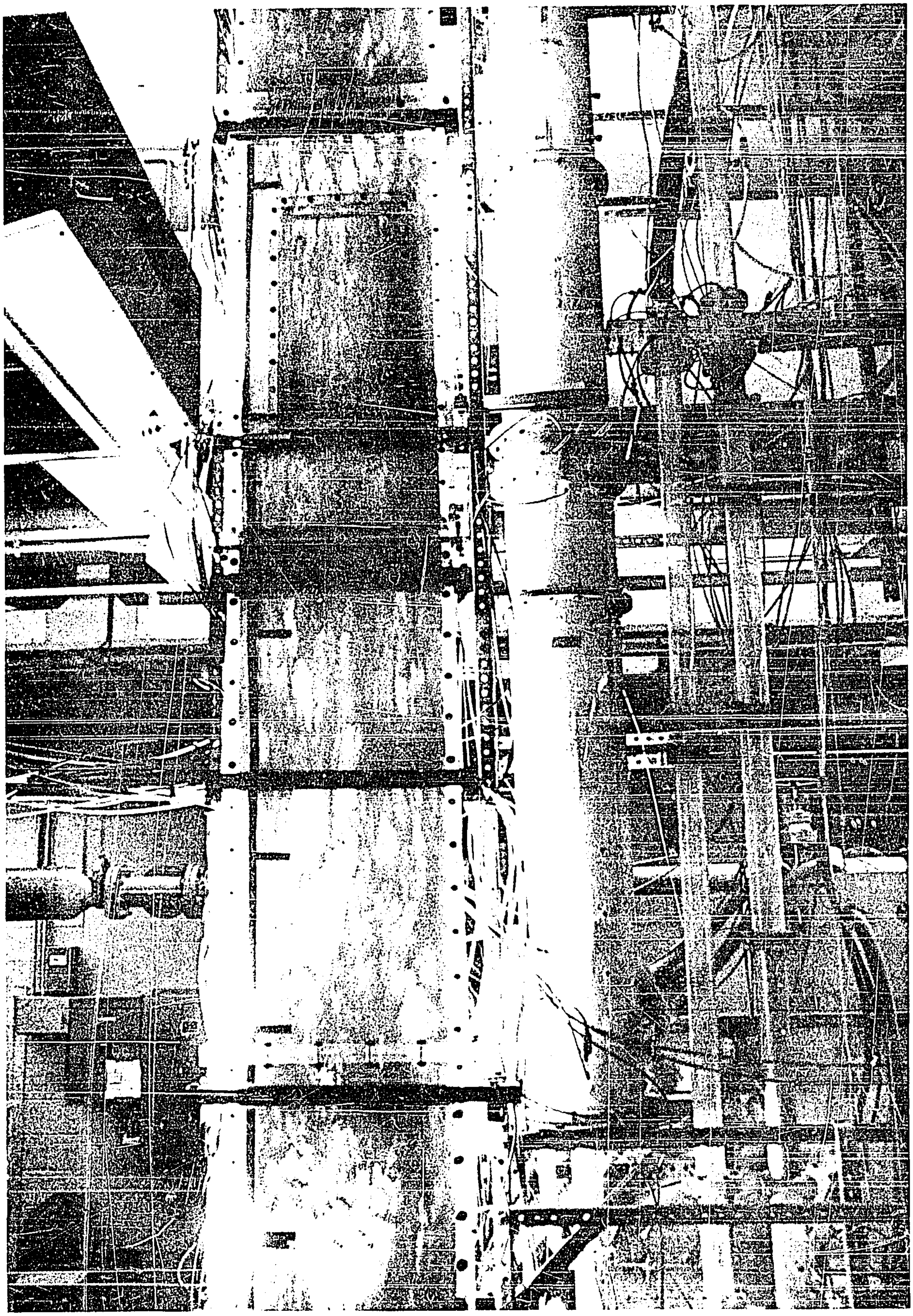


Exhibit 8
Fast 2-D Bed at a Superficial Gas Velocity of 11.5 ft/sec.
Solid is HFZ-20 (Solid C).