CIRCULATION OF LARGE PARTICLES IN AN AGGREGATIVELY FLUIDIZED BED

by

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To my parents, whose constant and everlasting words of encouragement paved the way to a successful completion of this work.

ABSTRACT

A new graphical method for predicting the minimum fluidization velocity was developed. This method gives an instantaneous measurement of minimum fluidization velocity.

The mixing behavior of cylindrical particles in an aggregatively fluidized bed, 9.72 inches in diameter, was studied as a function of particle size and density, gas flow rate, bed material and induced circulation. The cylindrical particles studied were 1/2", 1" and 2" in. diameter (= height) and the bed of solids was composed of 50 mesh sand. Some runs were also carried out with 80 mesh glass beads.

There was a density range (approximately 83-89 lb/ft^3) over which particles mixed in the bed (bed bulk density = 95.8 $\frac{lb}{ft^3}$ at minimum fluidization velocity). Above this range, particles sank to the distributor, and below this range they floated. The mixing range was influenced moderately by the gas velocity and greatly by induced circulation (using a plexiglass partition with auxiliary air flow on one side of the partition.

ACKNOWLEDGEMENT

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NOMENCLATURE

B	Generalized shape factor calculated with sieve diameter, dimensionless
d	Sieve diameter, ft
$D_p = d_p$	Particle diameter, ft
f ( )	Function of ( )
f_m	Modified friction factor, dimensionless
g	Gravitational acceleration, ft/sec^2
G	Mass flowrate, $lb/(hr)(ft^2)$
g_c	Conversion factor, 32.2 lb $ft/(lb_f)(sec^2)$
$G_{mf} = G_o$	Minimum fluidization mass flowrate, $lb/(hr)(ft^2)$
h	Height of pressure tap from bottom of bed
K	Coefficient of permeability, $\frac{lb}{lb_f} \cdot \frac{ft}{hr}$
L	Bed height, ft
N_{Ga}	Galileo number $= \frac{d_p^3 \rho_f (\rho_s - \rho_f) g}{\mu^2}$, dimensionless
N_{Re}	Particle Reynolds number $= \frac{d_p V \rho_f}{\mu}$, dimensionless
$N_{Re_{mf}}$	Particle Reynolds number at onset of fluidization, dimensionless

$N_{Re_{mf}}$	Particle Reynolds number at terminal conditions, dimensionless
$(-\Delta P)$	Pressure drop across the bed, lb_f/ft^2
$(-\Delta P_o)$	Pressure drop across the bed, corrected for bed support pressure drop, psi
$(-\Delta P_1)$	Pressure drop across the bed at pressure tap # 1, psi
ΔP_{Wo}	Weight of solids divided by the cross sectional area of the bed, psi
ΔP_{W1}	Weight of solids above pressure tap # 1 per unit area, psi
Q	Fluidization gas flowrate, ft^3/min
q	Auxiliary flow flowrate, ft^3/min
S	Distance below surface of bed traveled by large particle
Save	Average value of S
t_R	Residence time, sec
V	Superficial velocity of fluid, ft/sec
V_o	Terminal falling velocity, ft/sec
W	Superficial mass solid circulation rate in gas fluidized beds, $lb/(hr)(ft^2)$
Z	Height of sand in the region 1 to 2 (Figure 4)

GREEK LETTERS

ϵ	Bed voidage, dimensionless

ϵ_{mf}	Minimum fluidization voidage, dimensionless
ρ_b	Bulk density of bed, lb/ft^3
ρ_{bm}	Bed density at maximum porosity, lb/ft^3
ρ_f	Density of fluidizing medium, lb/ft^3
ρ_s	Density of solid particle, lb/ft^3
μ	Viscosity of fluid, lb/hr ft
Φ	Shape factor, dimensionless
φ	Axial position at the surface of the bed

INTRODUCTION

The stimulus for the widespread study of fluidized beds came about as a technical revolution in the petroleum industry in the early 1940's. Pioneering operations on a commercial scale were those of Winkler gas generators (1) in Germany in 1921 and those in the United States that pertained to catalytic cracking of oil vapors at about the same period.

The most ancient applications that resembled the methods of fluidization were the processes of purifying ores and municipal water supplies. The water, together with its suspended coagulated dirt, was allowed to run through a graded sand by gravity. The graded sand, therefore, served as a filter. As the dirt and solids accumulated on the sand filters, the permeability deteriorated and eventually completely blocked the passage of flow. The phenomenom of backwashing and expanded sand bed while being backwashed are examples of fluidization with a liquid medium.

However, it appears that the first patent taken on fludization was by Phillips and Bulteel (2) in 1910. The invention was a process for contacting a gas with a fine catalyst. The catalyst was suspended in the gas and carried into a reaction chamber. The reaction occurred and simultaneously the spent catalyst was carried by the products into a recovery vessel and recycled back to mix with the fresh catalyst feed lines. This was finally developed into a fluid catalytic process.

Recent applications of fluidization include: fluid catalytic reforming, one of the most frequently used means of upgrading naphthas;

fluid coking, used in producing higher yields of distillate oils; fluid catalytic oxidation of ethylene; iron ore reductions and more recently, it has been used for drying.

The reasons for the rapid and tremendous increase in the utilization of fluidization techniques are attributed to:

a) the presence of a greater surface area available to the gas which effect better transfer of heat and mass,

b) the rapid agitation of the particles by the incoming fluid. Turbulence in the fluidized bed helps to bring the mass to an isothermal condition,

c) the ease with which the particles can be transported.

Conventionally, two terms are used to describe the two forms of fluidization. "Particulate fluidization" refers to a liquid-fluidized bed in which the particles are evenly spaced so that the liquid passes smoothly through the interstices without the formation of bubbles. "Aggregative fluidization" refers to gas-solid systems where some of the gas pass through the bed as bubbles, and these can be seen to burst when they reach the top surface. The bubbles agitate the bed and consequently its height fluctuates.

Although fluidized beds have been extensively studied, no information is available regarding the mixing behavior of a large particle in a bed of small particles.

The present study was undertaken to observe

(I) the general behavior of an air-sand system in terms of the pressure drop across the bed as a function of gas velocity in the fixed bed and fluidized bed regions,

(II) the mixing behavior of cylindrical particles in air-sand and air-glass beads systems reported as a function of gas velocity,

(III) the recirculation of large particles in the sand bed promoted by recirculation of the sand using internals and auxiliary flow.

LITERATURE SURVEY

The fluidized technique, as it is now known, was initiated by the pioneering work of the Standard Oil Development Co., The M. W. Kellog Co., and Standard Oil of Indiana in their efforts to find a better catalytic-cracking process than the fixed-bed method which was introduced commercially in 1937. However, scattered references to early observations of what is known today as fluidization can be found in the published literature as far back as 1878 (3).

The early studies closely related and pertaining to fluidization were those of sedimentation and fixed beds. The flow of fluid in porous media can be viewed at one limit as flow past separate particles in a continuous fluid medium and at the other limit as flow through a solid body containing channels or pores. The classical expression for flow through porous media was contributed through the works of D'Arcy (4). The expression is:

$$G = K(-\Delta P/L) \qquad (1)$$

where G is the average rate of flow through the porous structure and K is the coefficient of permeability.

Perhaps the first paper to appear which dealt with the mechanics of fluidization as we know it today was that of Daniels (5) who discussed the mechanics of flow in a fluid catalytic cracking unit.

Based on the Carman-Kozeny (6) relationship for pressure drop in packed beds for laminar flow,

$$(-\Delta P) = \frac{200\, G\, \mu\, L\, \Phi^2}{\rho\, D_p^2\, g_c} \frac{(1-\epsilon)^2}{\epsilon^3} \qquad (2)$$

and on the Parent et al (7) result for the pressure drop at incipient fluidization,

$$(-\Delta P) = L\,(1-\epsilon)\,(\rho_s - \rho_f) \cdot \frac{g}{g_c} \qquad (3)$$

Leva (8) developed the most commonly used correlation in fluidization,

$$G_{mf} = 0.005 \frac{D_p^2\, g_c\, \rho_f\, (\rho_s - \rho_f)\, \epsilon_{mf}^3}{\mu\,(1-\epsilon_{mf})\, \Phi^2} \qquad (4)$$

which predicts the mass velocity required to initiate bed expansion.

The previous work by Leva et al (8) indicated that the groupings involving ϵ_{mf} and Φ were unique functions of the average diameter of the particles. Since void fraction and particle shape data are usually not available, the development of an equation omitting these terms would be helpful. Leva et al (8) correlated the data of several investigators using an equation of the form,

$$G_{mf} = k \frac{D_p^a\, [\rho_s - \rho_f]^b\, \rho_f^c}{\mu^d} \qquad (5)$$

The resulting equation from the evaluation data of ten sources was,

$$G_{mf} = 1.40 \times 10^5 \frac{D_p^{1.8235} [\rho_f (\rho_s - \rho_f)]^{0.9412}}{\mu^{0.88235}} \quad (6)$$

Many attempts have been made to improve Leva's equation. Miller and Logwinuk (9) defined and related a "critical mass velocity" by the following equation:

$$C.M.V. = 0.00125 \frac{D_p^2 (\rho_P - \rho_f)^{0.9} \rho_f^{1.1} g}{\mu} \quad (7)$$

The C.M.V. (which is Leva's G_{mf}) was graphically defined as the point where the laminar flow line intersected the turbulent flow line and was found to be independent of weight or bed height for a given material.

From the measurements of closely sized samples and mixtures of different sizes of powdered solids, using air, argon, carbon dioxide, nitrogen-hydrogen mixtures, town gas and methane as fluidizing gas, Heerden et al (10) arrived at the following C.M.V. equation:

$$Re_o = \frac{0.00123}{B} \frac{\rho_f \rho_{bm} g d^3}{\mu^2} \quad (8)$$

where Re_o $[= \frac{G_o d}{\mu}]$ is the particle Reynolds number, B is the generalized shape factor and ρ_{bm} is the bed density at maximum porosity.

More recently Joseph Frantz (11) tried to improve Leva's correlation by arguing that the assumption of pressure drop being equal to the weight of bed per unit area was frequently in error. Frantz recommended a somewhat less complicated equation,

$$G_{mf} = 4.45 \times 10^5 \frac{D_p^2 \rho_f (\rho_s - \rho_f)}{\mu} \tag{9}$$

Frantz also found that experimental values of G_{mf} increased with bed height up to heights of 1 ft.

G. Narishman (12) was the first to derive a generalized expression for the minimum fluidization velocity by extending the correlation proposed by Leva et al (8) into intermediate and turbulent flow regions. Narishman considered 3 cases in his derivation:

1) For a bed of uniform spheres,

$$G_{mf} = \frac{100188}{D_p} \mu \left[(1 + 0.00057\, N_{Re_t})^{1/2} - 1 \right] \tag{10}$$

where the subscript t refers to terminal conditions.

2) For a bed of irregular (nonspherical) particles,

$$G_{mf} = \frac{154440}{D_p} \mu \;(0.231 \log D_p + 1.417) \left[(1 + 2.12 \times 10^{-5} D_p^{-0.55} N_{Re_t})^{1/2} - 1 \right] \tag{11}$$

3) For irregular particles larger than 0.02 in.,

$$G_{mf} = \frac{154440}{D_p} \cdot f_o(\Phi_s)\,[(1 + 0.0056\, f_i(\Phi_s) \cdot N_{Re_t})^{1/2} - 1] \qquad (12)$$

where,

$$f_o(\Phi_s) = \frac{(0.232 + 0.42\,\Phi_s)}{\Phi_s} \qquad (13)$$

$$f_i(\Phi_s) = \frac{(0.768 - 0.42\,\Phi_s)^3}{(0.232 + 0.42\,\Phi_s)^2} \cdot \Phi_s^3 \qquad (14)$$

An overall standard deviation of 46% and an average deviation of ± 34% are obtained from Narishiman's correlations based on 267 data points tested.

Based on pressure drop considerations, Wen and Yu (13) developed a simpler and more accurate generalized correlation for the prediction of the minimum fluidization velocity. The proposed equation,

$$N_{Re_{mf}} = [(33.7)^2 + 0.0408\, N_{G_a}]^{1/2} - 33.7 \qquad (15)$$

is a correlation of 284 experimental values in a Reynolds number range of 0.001 to 4,000 with an overall standard deviation of 34% and an average deviation of ± 25%.

The circulation of solids in gas fluidized beds is known to be responsible for their unique characterisitics. Yet, very few quantitative results have been published in the literature. Talmor and Benenati (14) correlated the solids circulation rate, W, with the excess of superficial air flow rate over the minimum required for fluidization and the mean particle size. The resulting equation was,

$$W = .654 \ [(G - G_{mf}) \ e^{-2022 \ D_p}] \qquad (16)$$

J. F. Davidson (15) studied the differences between large and small fuidized beds, and observed that Gulf Stream circulation, that is, violent circulation current induced by bubbles was likely to be more vigorous in large beds. His apparatus consisted of a central draught tube concentric with a 0.3 m. diameter tube containing the bed. With the aid of a small sphere containing a transmitter, the radio pill, Davidson observed that the circulation path was related to the air flows through the draught tube and the annulus.

Prior to Talmor and Benenati (14) and Davidson (15), Leva (16) studied the effect which stirring of fluidized solids has on the pressure drop. He observed the effect of stirrer height, stirrer position, stirrer speed and power requirements.

Although fluidization has been extensively studied, no information is available regarding the mixing behavior of large particles in a bed of small solid particles.

THEORY

Minimum Fluidization Velocity

Two important design variables for fluidized beds are minimum fluidization velocity and pressure drop. Minimum fluidization velocity not only sets a lower limit on gas rate to the fluidized bed, but also is useful for prediction of bed expansion, for calculating heat transfer rates, in calculating pressure drop, etc. Pressure drop across the fluidized bed influences the sizing of the blower or compressor supplying the gas to the fluid bed and is useful as an index of fluidized bed solids inventory.

There are several methods of determining the minimum fluidization velocity and more often than not the application of the different methods to a particular system yield different results. One of the most commonly used method is Parent et al (7) result, which is based on the fact that when the pressure drop equals the weight of the bed per unit area corrected for buoyancy, theoretically the bed becomes fluidized, i. e.

$$(-\Delta P) = L\,(1-\epsilon)\,(\rho_s - \rho_f)\cdot\frac{g}{g_c} \tag{17}$$

or $(-\Delta P) = L\rho_B$, if the fluid is a gas (i. e. $\rho_s \gg \rho_f$)

where,

ρ_s = density of solid particles

ρ_f = density of fluid

ϵ = fraction of voids

L = height of packing

ρ_B = bulk density of bed = $\rho_s (1 - \epsilon)$

Another non-graphical method consists in observing the point at which the bubbles first appear.

These non-graphical methods appear to be indeed theoretically sound, but actually, the bed may bubble at lower gas velocities, owing to bypassing around stagnant areas or may even remain stationary at rates above the theoretical minimum, owing to the extra force required to overcome electrostatic forces.

The most commonly used graphical method consists in measuring the pressure drop through the bed of solids. Some authors have even observed the change of heat transfer coefficent with gas velocity as a criteria for the minimum fluidization velocity.

These graphical methods are quite useful if the behavior of the system under study is of the form shown in Figure 1 and 2, i. e. with definite changes.

However, if the behavior of the system is of the form shown in Figure 3, i. e. a smooth curve, the location of a particular discontinuity or inflection point becomes guesswork. Some authors have used the method of intersecting lines (see Figure 3).

A simple but new approach was used in this work to calculate the minimum fluidization velocity of the system under study. In order

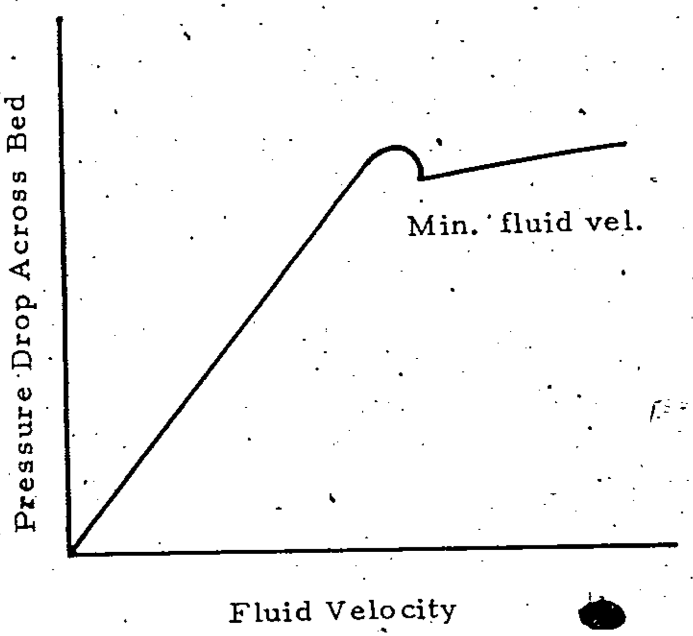

Figure 1: Typical of well packed Beds

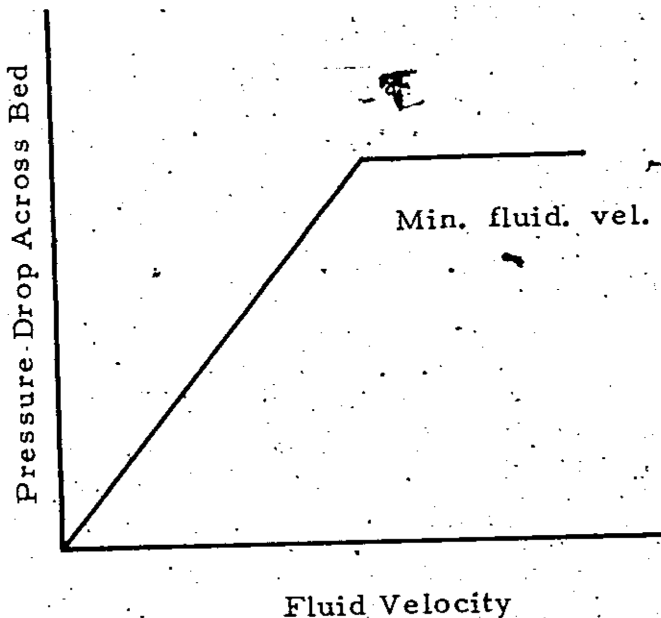

Figure 2: Typical of the fluidization of fine powders

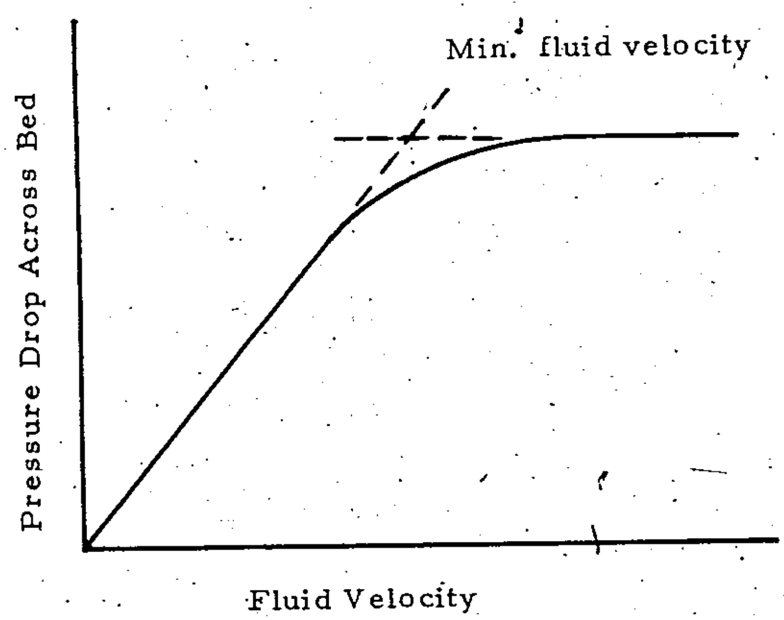

Figure 3: Typical of most beds with a low bulk density

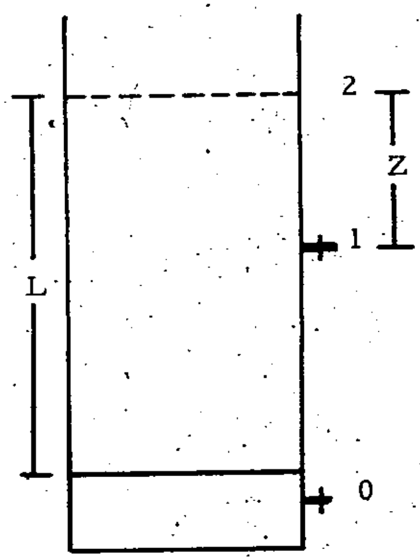

Figure 4: Fluidization Column

to show the theory behind this new approach, consider a bed packed with particles up to a height L, with two pressure taps, one measuring the total pressure drop in the bed and one located a distance Z from the top of the bed which measures the pressure drop between points 1 and 2 as shown in Figure 4.

In the fixed bed region, assuming ε (or ρ_B) to be constant throughout the bed, then

$$\frac{(-\Delta P_1)}{Z} = \frac{(-\Delta P_o)}{L} \tag{18}$$

and

$$\frac{(-\Delta P_1)}{\rho_B Z} = \frac{(-\Delta P_o)}{\rho_B L} \tag{19}$$

Therefore,

$$\frac{(-\Delta P_1)}{\Delta P_{W1}} = \frac{(-\Delta P_o)}{\Delta P_{Wo}} \tag{20}$$

where:

ΔP_{Wo} = weight of solids per unit area = constant

ΔP_{W1} = weight of solids per unit area in the region from 1 to 2 in Figure 4

Now, beyond the point of incipient fluidization,

$$\frac{(-\Delta P_1)}{\Delta P_{W1}} > \frac{(-\Delta P_o)}{\Delta P_{Wo}},$$ because the amount of solids in the 1-2 region increases due to bed expansion while the amount of solids in the 0-2 region remains constant.

Therefore, the point beyond the straight line region (fixed bed) at which equation (20) does not hold is known as the minimum fluidization velocity.

APPARATUS AND EXPERIMENTAL PROCEDURE

Apparatus

A simple apparatus, a schematic diagram of which is shown in Figure 5, was used to obtain the required experimental data. Originally, the system consisted of a fluidizing column, a rotameter manufactured by Brooks Rotameter Co. (Lansdale, Pennsylvania; Type 1110, Serial #39179 and reported maximum reading of 50 ft^3/min. at 70°F and 40 psig.), two manometers (specific gravity of oils, 0.827 and 1.75) and a pressure gauge manufactured by Binks Mfg. Co. (Chicago, U.S.A.).

The fluidizing column, 9.72 inches in internal diameter and 53 inches in length, was made of plexiglass to allow for visual observation of the fluidized solids. The bed of solids was supported by a porous stainless steel plate (nominal 200 mesh pores) eleven inches in diameter and 0.35 inches thick, which in turn was supported by plexiglass flanges. Seven pressure taps were located along the column, one just below the porous plate and the others at intervals of six inches starting at 3 inches above the plate. There were two air inlets at the base of the column, 180° apart, to allow for better distribution of the air.

Air was used as the fluidizing medium and was maintained at 40 psig by a standard pressure regulator in the measurement section.

The bed of solids was composed of silica sand varying in size from larger than 70 Mesh (0.0083 inches) to a maximum of 25 Mesh

(0.0278 inches) and with a density of 165.345 lb/ft^3. Glass beads (-60 to + 85 Canadian Mesh) were also used.

Once part I was concluded the column was shortened to 36 inches for convenience in part II. The apparatus was further modified in part III by adding:

(1) a vertical plexiglass divider two inches above the porous plate, eight inches high and nine inches wide, which separated the bed into two columns, unequal in area. Both columns having a rounded side and a common flat side.

(2) an auxiliary air flow, two inches above the porous plate which entered the smaller column. The auxiliary flow rotameter was manufactured by Brooks Instrument of Canada Limited (Scarborough, Ontario; Type 8-1110, serial #625-1058-2).

Key to Numbers Figure 5

1	Air inlet valve
2	Pressure gauge and filter unit
3	Main rotameter
4	Auxiliary flow rotameter
5	Main valve leading to the fluidizing column
6	Auxiliary flow valve
7	Manometer #1, specific gravity of oil = 1.75
8	Manometer #2, specific gravity of oil = 0.827
9	Pressure taps
10	Fluidizing column
11	Fluidizing particles

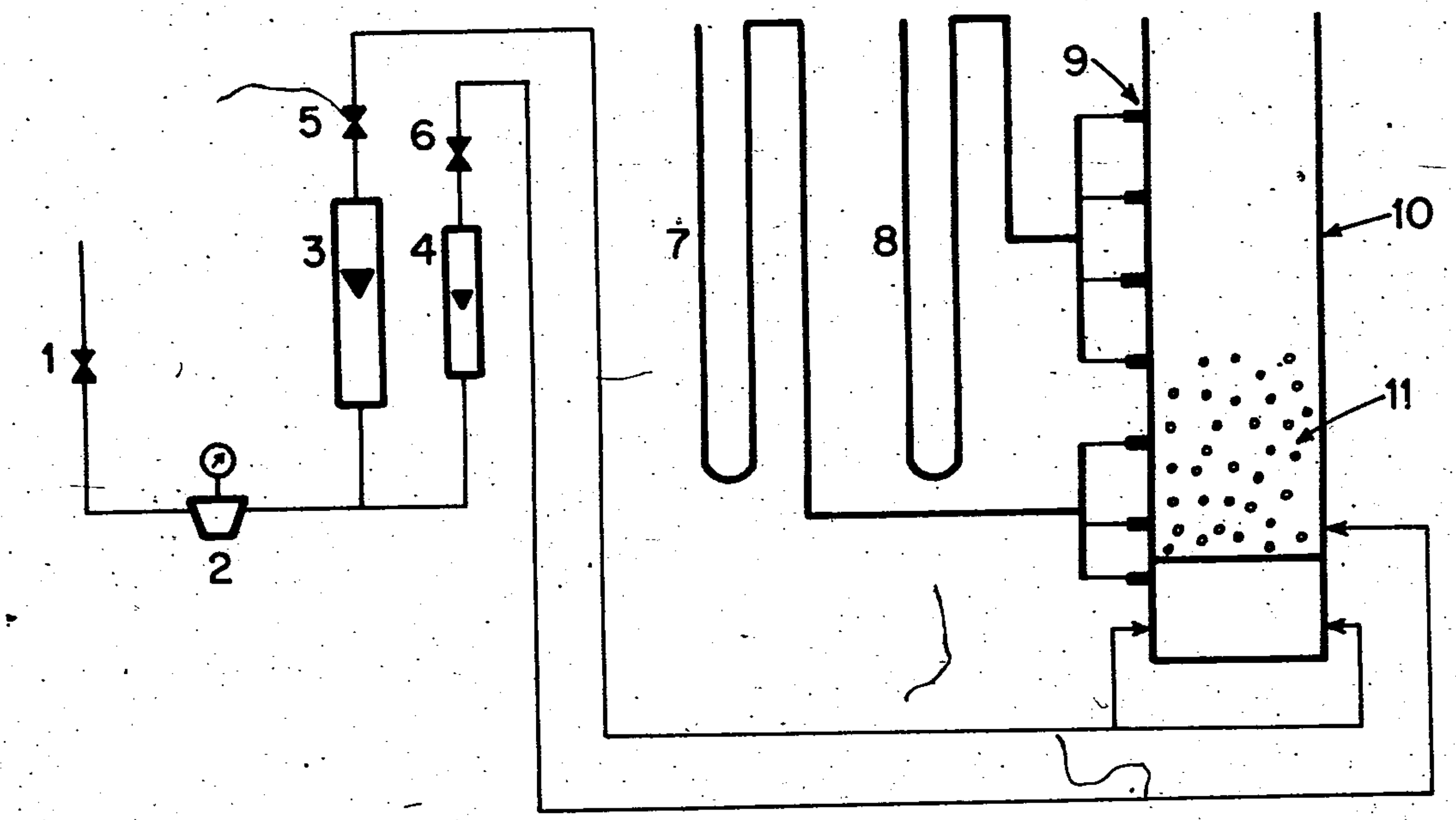

FIGURE 5: <u>SCHEMATIC DIAGRAM OF EXPERIMENTAL SET-UP</u>

PROCEDURE

Part I

Compressed air at a constant 100 lb/in^2 gauge pressure was obtained from the building compressor. It was first passed through a filter and pressure regulator device so as to removed the entrained oil and water droplets carried along from the compressor and reduce the pressure in the measurement section to 40 psig. The dried air was allowed to flow through the bed of solids and readings of pressure drop versus gas flowrate were recorded. Four different weights corresponding to multiples of 25.5 lb up to 102 lb were studied.

Part II

A known weight of sand was poured into the fluidization column, and the bed was fluidized. The gas flowrate was decreased slowly and the particles allowed to settle freely. The resulting bed height was taken as the fixed-bed height. A one-inch cylindrical particle (one inch in diameter by one inch in height) was placed with its flat end down on the surface of the fixed bed. Five different positions were chosen for measurement. They were position (φ) and the four radial positions 90° apart near the wall of the bed (A, B, C, D).

After the particle was placed, the gas flowrate was adjusted to a particular setting, and the behavior of the particle was observed. This was repeated for the range of gas flowrates of interest. The depth to which the particle sank was observed on a scale on the side of the particle. A thin thread was attached in order to determine its position when below the surface of the bed.

The sand bed was composed of 51.0 lb of silica sand (-25 to +70 Mesh) and had an initial bulk density of 96.6 lb/ft^3 which corresponded to an initial height of 12.3 in. Twenty-five different one-inch cylindrical particles varying in density from 38 lb/ft^3 to 169 lb/ft^3 were studied. The above procedure was repeated using 2 in. and 1/2 in. cylindrical particles. All particles were made of aluminum. Similar studies were made with an air-glass beads (-60 to +85 Canadian Mesh) system using one-inch cylindrical particles.

Part III

The mixing behavior of the one-inch cylindrical particles was further studied by recirculating the sand together with the cylindrical particles. The residence time of the cylindrical particles as well as an approximation of the sand mass flowrate were recorded. The amount of sand in the bed was 37.9 lb.

RESULTS AND DISCUSSION

Pressure Drop as a Function of Gas Flowrate and Bulk Density*

Data consisting of pressure drop across a sand bed as a function of gas flow rate were obtained for an air-sand fluidization system. The data was then plotted with bulk density as parameter as shown in Figures 6 through 9 and exhibited the following classical behavior:

1) Forward Flow:

a) Fixed bed region:

At very low velocities the gas percolated through without agitating the individual particles. The pressure drop increased linearly with velocity and was less than the weight of the bed per unit area.

b). Fluidized bed region:

The accepted theoretical explanation (7) for the transition from the fixed bed region to the fluidized bed region is that when the velocity is increased sufficiently, the pressure drop should be equal or slightly in excess of the weight of solids (per unit area) present. At this point the pressure drop is slightly more than enough to support the weight of particles, owing to the consolidation of the bed. A slight increase in flow above that point frees the particles ((-ΔP) reduces) and the pressure drop becomes just enough to support, the weight of bed,

* All data in this thesis are for a 9. 72 inch (inside diameter) bed.

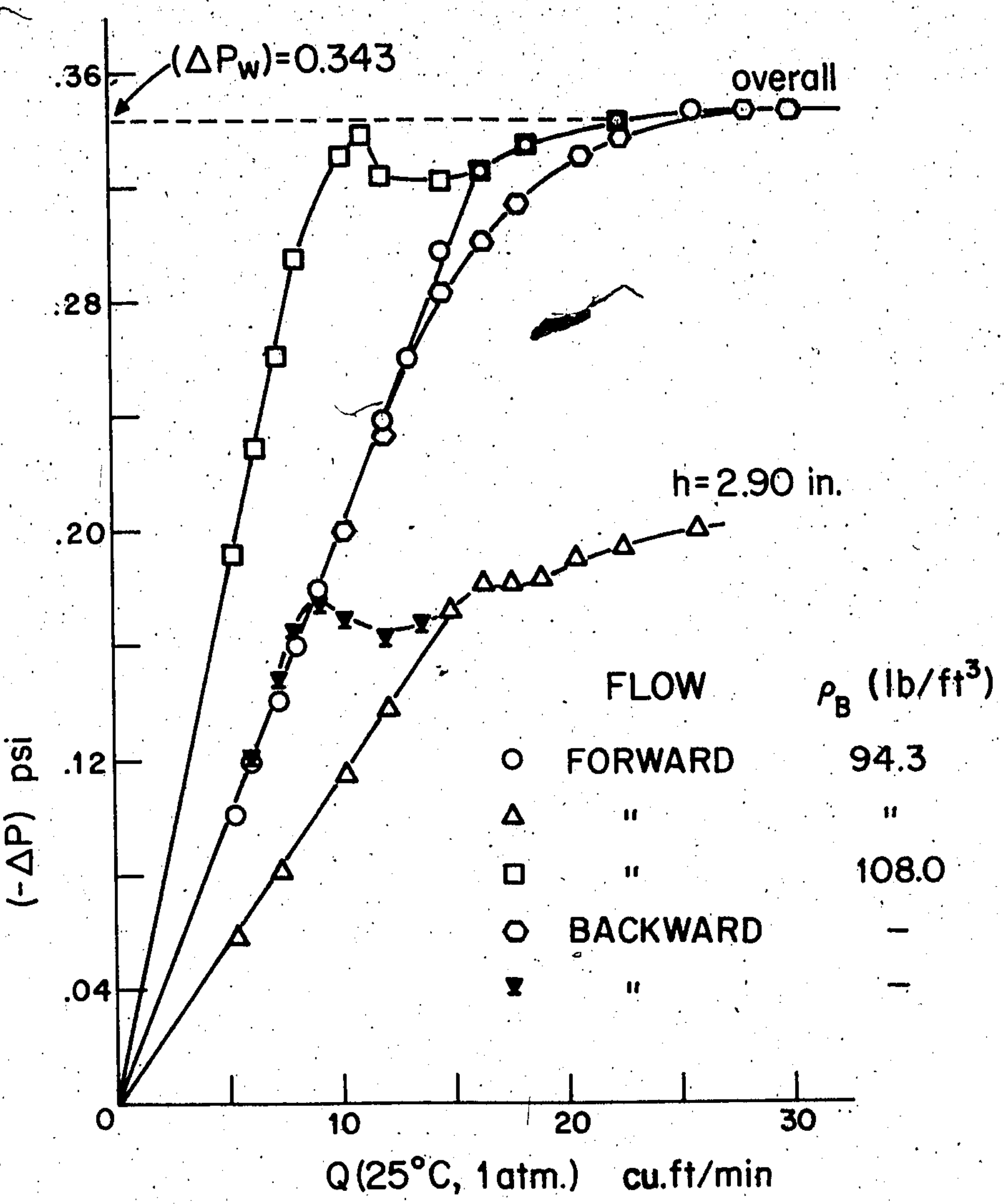

FIGURE 6: PRESSURE DROP ACROSS 25.5 LB OF SAND

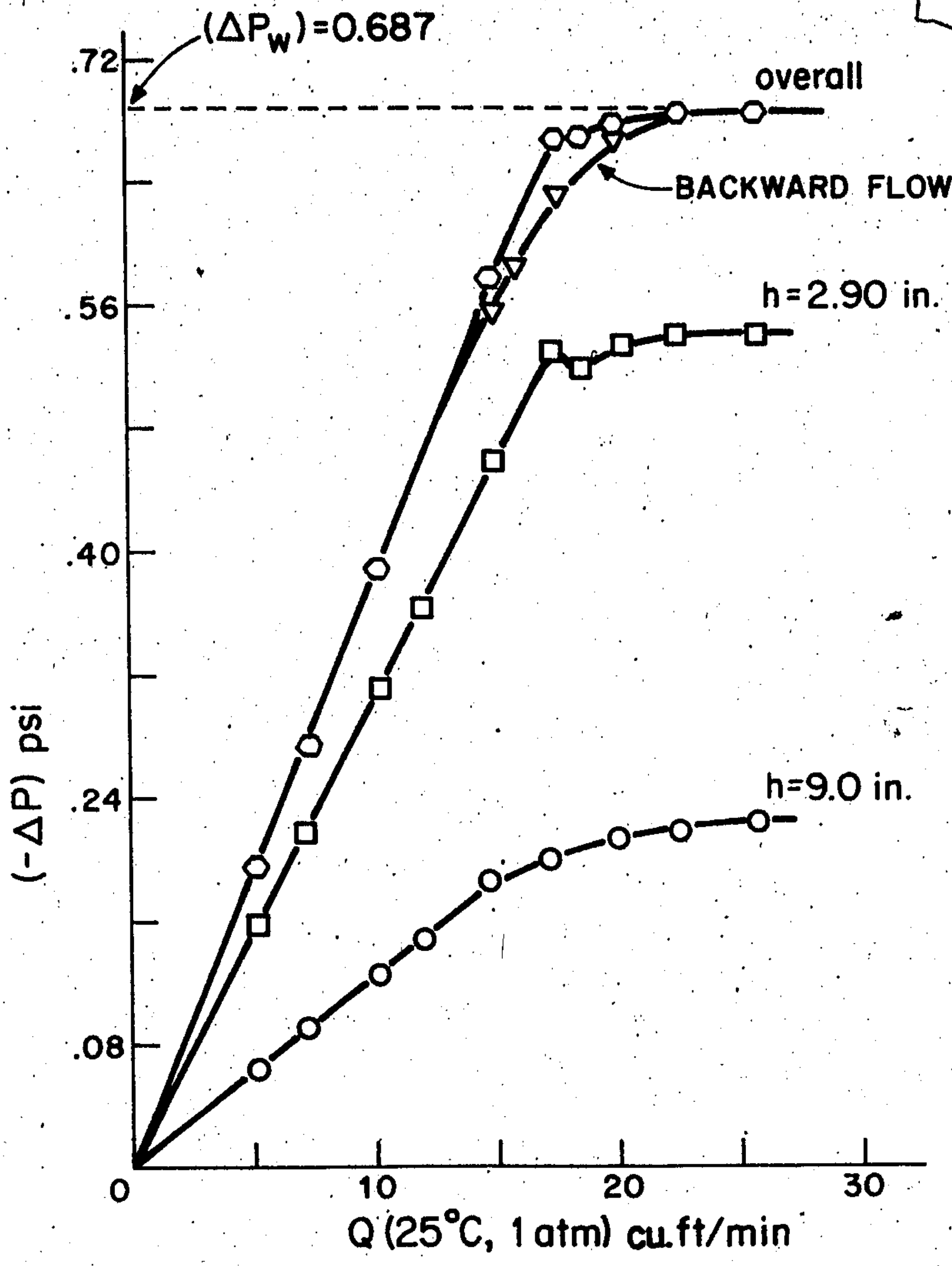

FIGURE 7: PRESSURE DROP ACROSS 51.0 LB OF SAND

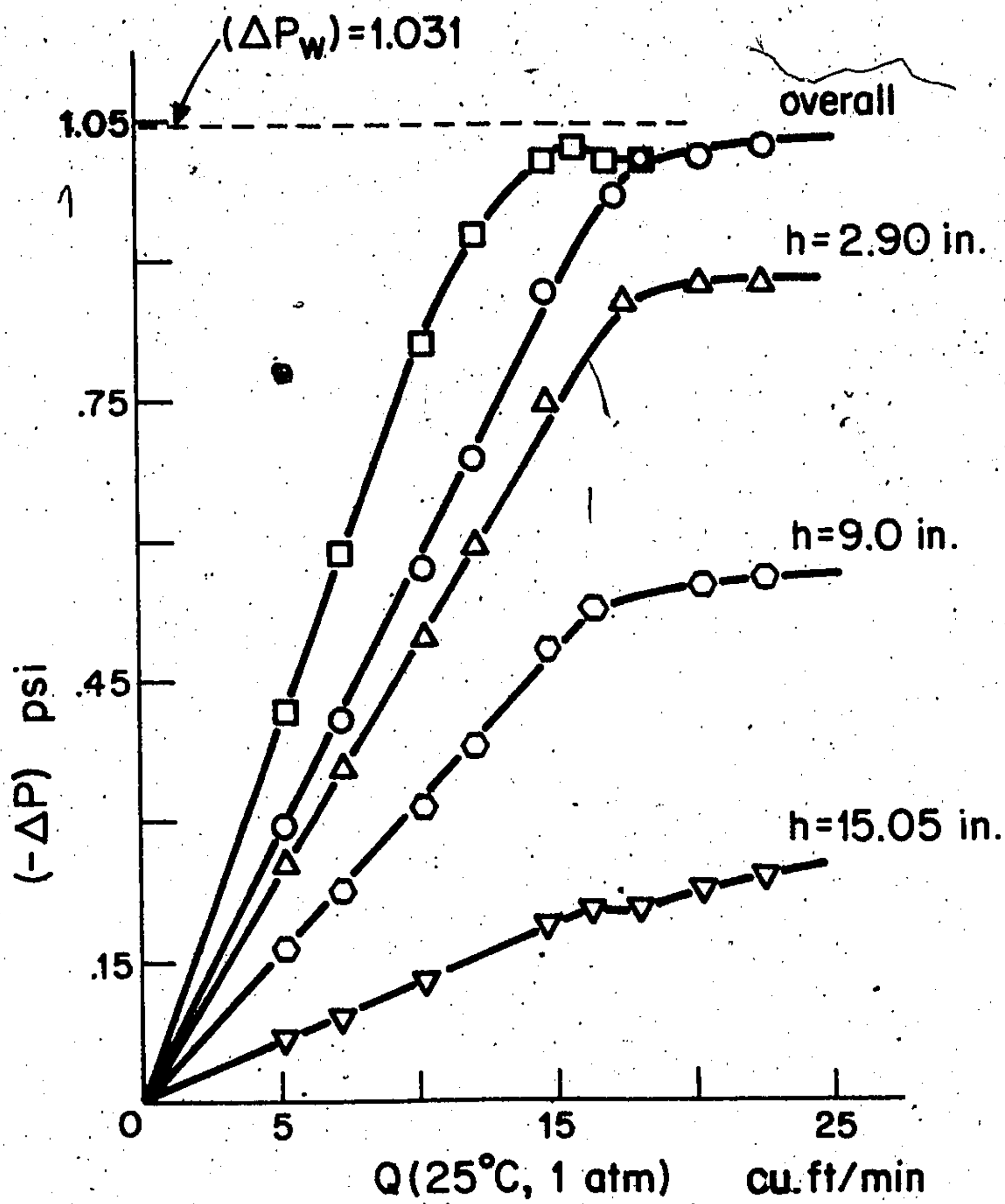

FIGURE 8: PRESSURE DROP ACROSS 76.5 LB OF SAND

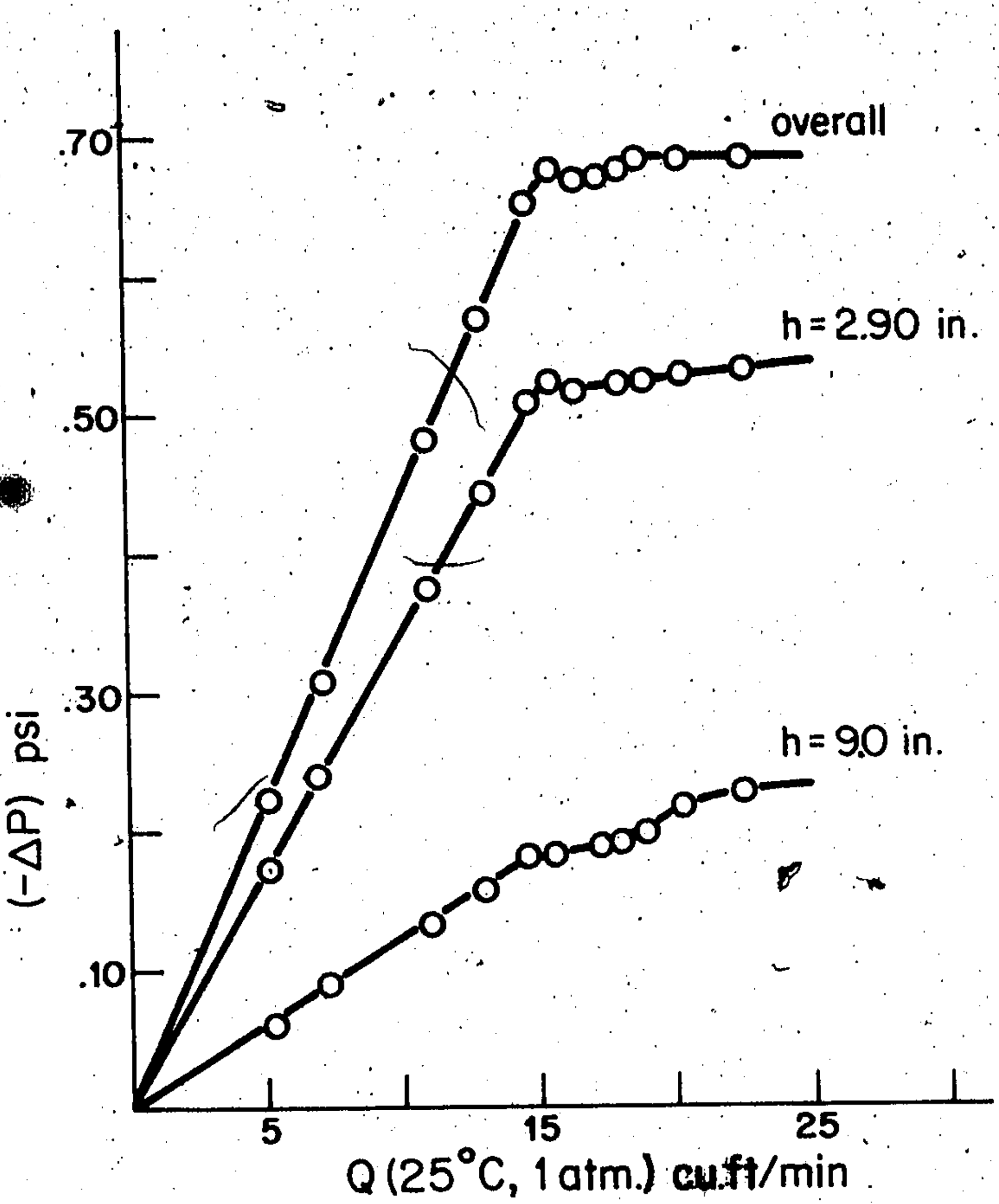

FIGURE 9: PRESSURE DROP ACROSS 51.0 LB OF SAND

and this point is usually defined as the minimum fluidization velocity.

Although some authors (17) agree that the above explanation is frequently in error and have reported deviations of as much of 22%, the majority of the cases have proved the contrary. In this work deviation of less than 1% to a maximum of 6% were observed.

2) Backward Flow:

Ideally the backward flow curve should coincide with the forward flow curve. As shown in Figure 6 through 9 deviations may occur in certain regions.

a) Fluidized bed region:

In the highly fluidized region both curves will obviously coincide since the particles are in random motion and all hindering forces such as bed consolidation and electrostatic effects have been overcome.

b) Transition region:

The backward flow curve will always lie below the forward flow curve in the transition region. Since there are no forces to be overcome (particles are suspended in the gas) the gas is exposed to a lower resistance as it percolates through the bed of solids.

c) Fixed bed region:

Whether the backward flow curve lies below or

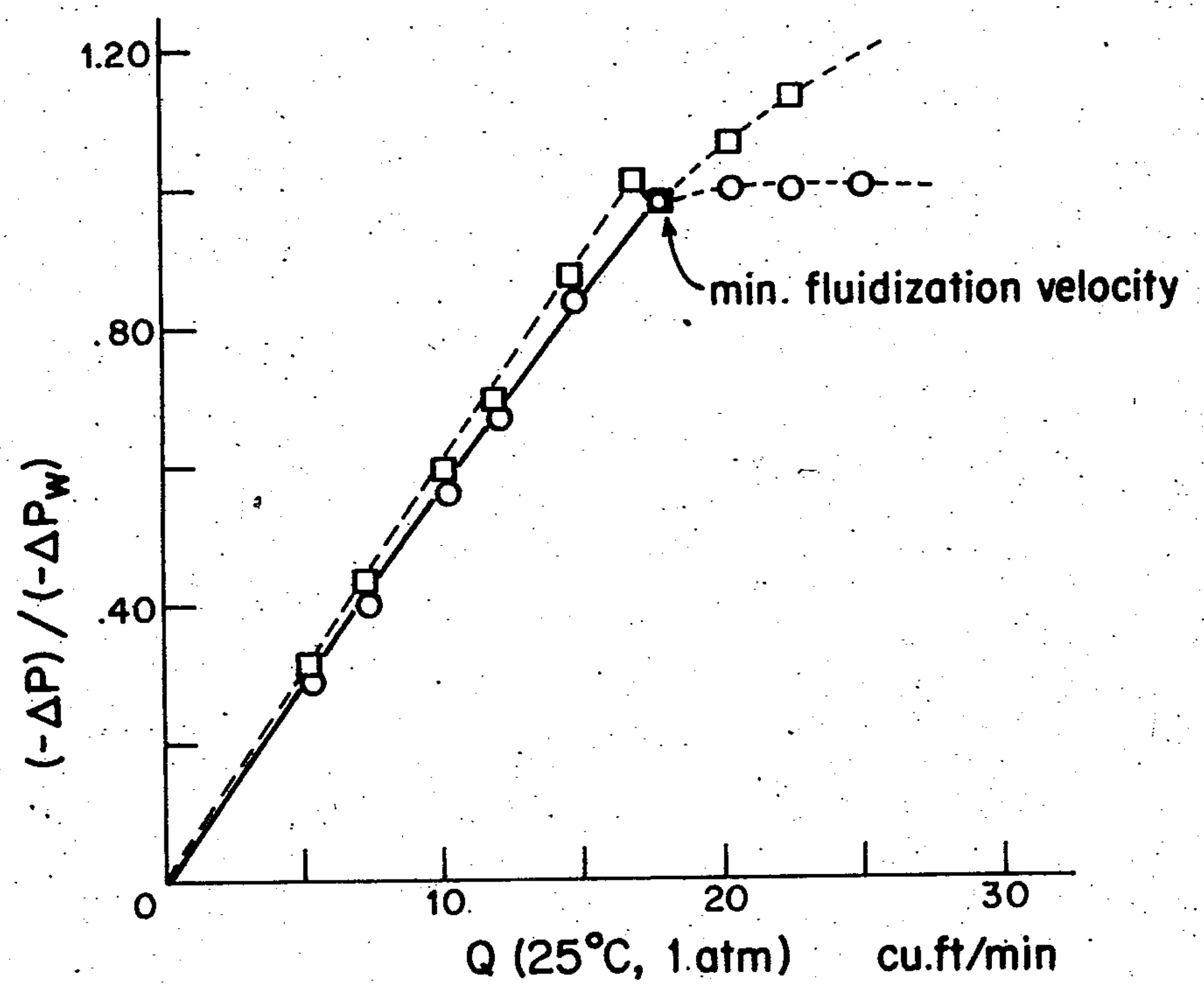

FIGURE 10: MINIMIUM FLUIDIZATION VELOCITY DETERMINATION

above the forward flow curve in the fixed bed region depends upon the final packing arrangement. If the settled bulk density is less than the initial bulk density of the bed then the backward flow curve lies below. It lies above if the opposite is true.

The minimum fluidization velocity was found by the graphical method as shown in Figure 10. It is clear from this figure that this method gives an instantaneous reading of the minimum fluidization velocity. The results were compared with existing methods and tabulated in Table 1.

The non-graphical methods (#2 and 3), although theoretically sound, are frequently in error since the bed may bubble at lower gas velocities, owing to bypassings around stagnant areas or the pressure drop at incipient fluidization may not equal the weight of the bed per unit area corrected for buoyancy. These two behaviours were clearly evident in this work. The method used in this work (#4) is in excellent agreement with the most commonly used method (#1).

As expected, Figure 11 shows that for a particular system, the normalized pressure drop, $\frac{(-\Delta P)}{(-\Delta P_W)}$, is a function of bulk density and gas flow rate in the fixed bed region and a function of gas flow rate alone in the fluidized bed region.

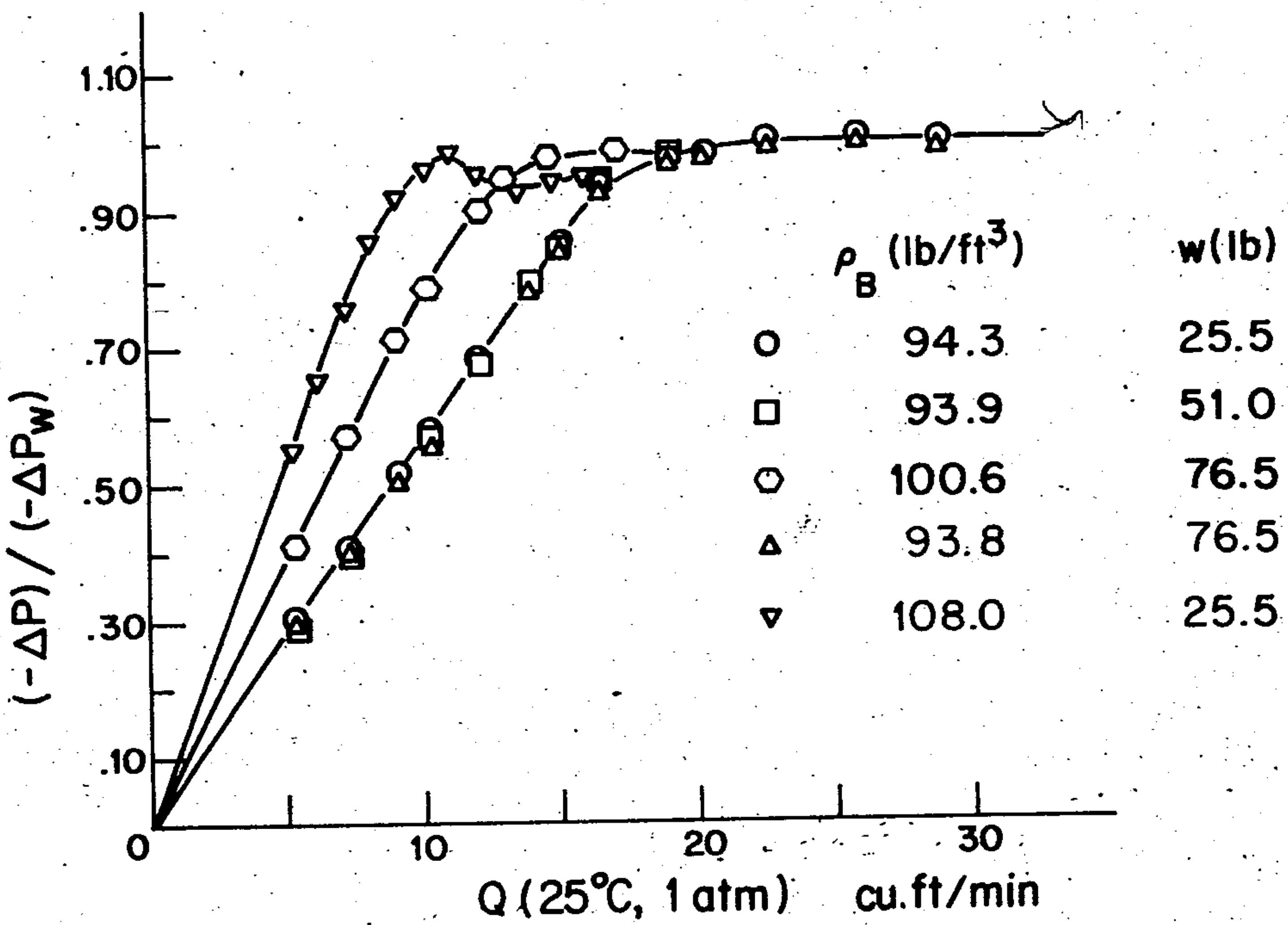

FIGURE 11: NORMALIZED PRESSURE DROP

The evenness with which a bed fluidizes was found to decrease with height because of bubbling coalescence. Bubbles tend to interact with each other and coalesce so that bubble size increases with height while the number decreases. When no further increase in bubble size is possible, then we have the so called slugging effect.

Mixing Behavior of Cylindrical Particles

Essentially the data obtained consisted of the amount of solid in the unit, average superficial gas velocity, pressure drop, bed height and the mixing behavior of cylindrical particles. Two-inch, one-inch and half-inch cylindrical particles were studied in the air-sand system and one-inch particles were studied in an air-glass beads system. All observations are recorded in appendices E, F, G, H, and I.

a) Air-Sand System

(a-1) Bubbles were observed around the particles at readings as low as half the minimum fluidization velocity.

(a-2) As the particle sank, sand was displaced upwards with the aid of the air bubbles. When bubbling around the particle stopped, the particle stopped sinking.

(a-3) The size of the bubbles around the particles was proportional to the size of the cylindrical particle but it was not affected by density.

TABLE 1

Weight of bed (lb)	ρ_{Bo} (lb/ft^3)	Minimum Fluidization Flowrate* (ft^3/min.) #1 Graphical Method	#2 Bubbles first appear	#3 Weight of bed per unit area	#4 This work
25.5	94.3	16.8	14.6	22.5	17.5
25.5	108.0	12.4	—	11.5	11.5
51.0	93.9	17.2	15.0	22.5	17.2
76.5	93.8	17.2	15.1	22.5	17.2
76.5	100.6	—	12.0	—	15.0

* All data in this thesis are for a 9.72 inch (inside diameter) bed.

(a-4) The one-inch and two-inch cylindrical particles, irrespective of density that would eventually sink or float half exposed were completely covered at a gas flowrate of 13.8 ft^3/min. (and not before nor after) which was below the minimum fluidization velocity of 15.5 ft^3/min. Half-inch particles began sinking above 13.8 ft^3/min. and were completely covered at 14.6 ft^3/min.

(a-5) It was observed that if any of the particles used were to sink, they would do so at 16 to 17 ft^3/min, just slightly above the minimum fluidization velocity of 15.5 ft^3/min.

(a-6) The ideal particle density for mixing, irrespective of size, was found to be between 83 to 89 lb/ft^3. The bed density at incipient fluidization was 95.8 lb/ft^3 and decreased with gas flow rate as shown in table H-2.

(a-7) It was also observed that the bubbles play an important role in the mixing behavior of the cylindrical particles. Although the density of the bed may be lower than the density of a particular particle, the particle may float due to upcoming bubbles.

b) Air-Glass Beads System

The mixing behavior of one-inch cylindrical particles in the air-glass beads system was, not surprisingly, similar to the results obtained with the air-sand system. The ideal particle density was also found to be 83 to 89 lb/ft^3.

Recirculation of Large Particles

The somewhat restrictive density range of cylindrical particles that will effectively mix with the bed of solids and the lack of control of the residence time of these particles led to the study of the mixing behavior of one-inch cylindrical particles in a recirculating fluidized sand bed. Recirculation was accomplished by an internal plexiglass divider and an auxiliary air flow as described in the apparatus section.

i) As shown in Appendix J, only a small auxiliary flow was required to induce recirculation at fluidizing flowrates above the minimum fludization velocity. The higher the flow of the fluidizing gas the lower the required auxiliary flow. Nevertheless, a minimum auxiliary flowrate of about 2.0 to 2.4 ft^3/min. was required for effective recirculation, relative to the flowrate of 15.5 ft^3/min. at incipient fluidization. Circulation of sand in the order of 0.7 to 2 tons/hr was observed (Table J-7).

ii) The density range of cylindrical particles that will recirculate with the bed of solids is proportional to the fluidizing and auxiliary flowrates. The residence time of the recirculating particles is inversely proportional to both flows and does not exhibit a particular dependency on particle density.

iii) It was possible to see a definite boundary between the moving (recirculating) sand and the stagnant portion of solids. This boundary was maximum at the top of the bed and decreased diagonally to about one inch from the divider at the bottom of the bed.

iv) The closer to the divider the faster the particle recirculated.

v) Particles that sink to the bottom do so by following the downward recirculation stream and settling in the less dense region.

vi) Recirculation decreased bubbling.

CONCLUSIONS

1) A new method for determining the minimum fludization velocity was developed. It consisted in observing the graphical variation of the normalized pressure drop as a function of gas flowrate with pressure tap location as parameter.

2) There is a range of particle densities over which particles mix in the bed. This range is 83 to 89 lb/ft^3 for one-inch cylinders in the sand bed. Below this range the particles float, above they sink.

3) The overall bulk density of the bed was higher than the ideal mixing density range of the cylindrical particles at all flowrates studied. As a matter of fact, particles with densities lower than that of the bed density (but above the ideal value of 89 lb/ft^3) sank to the bottom of the bed. This could be due to a lower localized density caused by the presence of the cylindrical particles and the upcoming gas.

4) Two-inch and one-inch cylindrical particles can be considered as large particles relative to the bed, whereas half-inch particles cannot.

5) The density range of cylindrical particles that will effectively mix with the bed can be greatly increased by a relatively simple recirculating system. For example, the range was increased from 83 to 89 lb/ft^3 to 65 to 112 lb/ft^3 by using an auxiliary flow rate of 2.91 ft^3/min., which is less than 19% of the minimum fluidization flowrate.

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APPENDIX A

Pressure Drop Across Plate and Rotameter Calibration Data

TABLE A-1

PRESSURE DROP ACROSS STAINLESS STEEL POROUS PLATE

ROT RDG	(-ΔP) (in. of oil*)	(-ΔP) (psi)	ROT RDG	(-ΔP) (in. of oil*)	(-ΔP) (psi)	ROT RDG	(-ΔP) (in. of oil*)	(-ΔP) (psi)
5.0	0.21	0.0063	16.0	0.65	0.0194	27.0	1.09	0.0326
6.0	0.25	0.0075	17.0	0.69	0.0206	28.0	1.13	0.0338
7.0	0.29	0.0087	18.0	0.73	0.0218	29.0	1.17	0.0350
8.0	0.33	0.0099	19.0	0.77	0.0230	30.0	1.21	0.0362
9.0	0.37	0.0111	20.0	0.81	0.0242	34.0	1.37	0.0409
10.0	0.41	0.0122	21.0	0.85	0.0254	38.0	1.55	0.0463
11.0	0.45	0.0134	22.0	0.89	0.0266	42.0	1.74	0.0520
12.0	0.49	0.0146	23.0	0.93	0.0278	46.0	1.91	0.0571
13.0	0.53	0.0158	24.0	0.97	0.0290	50.0	2.07	0.0618
14.0	0.57	0.0170	25.0	1.01	0.0302			
15.0	0.61	[illegible]	26.0	1.05	0.0314			

* Specific Gravity of oil = 0.827

TABLE A-2

ROTAMETER CALIBRATION*

Rotameter at 40 psig and Volumetric Flow Rate at 25°C and 1 atm.

MAIN ROTAMETER		Auxiliary Flow Rotameter	
ROT RDG	Q (ft^3/min)	ROT RDG	q (ft^3/min)
5.0	5.30	0.05	1.01
7.0	7.28	0.07	1.24
10.0	10.17	0.10	1.60
12.0	11.96	0.12	1.86
15.0	14.60	0.15	2.25
17.0	16.39	0.17	2.53
20.0	18.86	0.20	2.91
22.0	20.31	0.22	3.17
25.0	22.51	0.25	3.66
30.0	25.75	0.27	3.88
		0.30	4.31

* For description of rotameters see "Apparatus" section

APPENDIX B

Pressure Drop Data

Part I

Key to Numbers in Remark Section

MEANING	NUMBER
First sign of bubbling appears in small area	(1)
Slugging effects; manometer difficult to read due to up and down movement of oil	(2)
Bed evenly fluidized	(3)
Bubbling disappears	(4)

L = a number, specifies the height of the sand bed (in inches) at a particular gas flow rate.

$(-\Delta P_o)$ = Pressure drop across bed of solids, psi (corrected for distributor pressure drop)

$(-\Delta P_1)$ = Pressure drop at pressure tap # 1, psi

All data are for a 9.72 inch (inside diameter) bed

TABLE B-1

WEIGHT OF SAND BED: 25.5 lb

BULK DENSITY IN THE FIXED BED REGION: 94.3 lb/ft^3

FORWARD FLOW DATA (See Figure 6)

MAIN ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ psi	REMARKS
5.0	0.101	0.058	L = 6.30
6.0	0.122	0.071	
7.0	0.141	0.081	
8.0	0.157	0.091	
9.0	0.179	0.104	
10.0	0.200	0.115	
11.0	0.219	0.127	
12.0	0.239	0.137	
13.0	0.258	0.149	
14.0			
15.0	0.298	0.172	(1)
16.0	0.315	0.181	
16.5	0.323	0.180	
17.0	0.326	0.181	L = 6.35
17.5	0.327	0.178	

(Cont'd)

MAIN ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_l)$ psi	REMARKS
17.0	0.326	0.181	L = 6.35
17.5	0.327	0.178	
18.0	0.329	0.179	L = 6.40
19.0	0.331	0.179	
20.0	0.334	0.183	L = 6.42
21.0	0.337	0.186	
22.0	0.339	0.190	
23.0	0.342	0.191	L = 6.65
25.0	0.343	0.194	L = 6.80
27.0	0.345	0.197	
29.0	0.346	0.199	
31.0	0.346	0.200	
34.0	0.346	0.200	
38.0	0.347	0.200	

TABLE B-2

WEIGHT OF SAND BED: 25.5 lb

BACKWARD FLOW DATA (See Figure 6)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	REMARKS
38.0	0.347	0.200	
34.0	0.346	0.200	
33.0	0.346	0.200	
29.0	0.345	0.197	
27.0	0.342	0.196	
25.0	0.336	0.192	L = 6.80
23.0	0.331	0.189	
21.0	0.324	0.187	
19.0	0.314	0.181	L = 6.45
18.0	0.314	0.183	
17.0	0.308	0.182	L = 6.40
16.0	0.299	0.177	(3)
15.0	0.283	0.166	
14.0	0.272	0.162	

(Cont'd)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	REMARKS
13.0	0.253	0.152	
12.0	0.233	0.140	
11.0	0.219	0.130	L = 6.20
10.0	0.200	0.121	
9.0	0.179	0.114	(4)
7.0	0.143	0.087	
5.0	0.105	0.064	L = 6.17

TABLE B-3

WEIGHT OF SAND BED: 25.5 lb

BULK DENSITY IN THE FIXED BED REGION: 108.0 $\frac{lb}{ft^3}$

FORWARD FLOW DATA (See Figure 6)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	REMARKS
5.0	0.190	0.107	L = 5.50
6.0	0.228	0.129	
7.0	0.261	0.148	
8.0	0.295	0.165	
9.0	0.318	0.175	
10.0	0.331	0.170	
11.0	0.338	0.167	L = 5.60
11.5	0.341	0.167	(1)
12.0	0.327	0.164	
12.5	0.325	0.168	
13.5	0.321	0.169	
15.0	0.322	0.173	
16.0	0.325		
17.0	0.326	0.175	

(Cont'd)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	REMARKS
18.0	0.329	0.175	
19.0	0.330	0.176	
20.0	0.333	0.180	
21.0	0.335	0.182	
23.0	0.339	0.184	

TABLE B-4

WEIGHT OF SAND BED: 51.0 lb

BULK DENSITY IN THE FIXED BED REGION: 93.9 lb/ft^3

FORWARD FLOW DATA (See Figure # 7)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	$(-\Delta P_2)$ (psi)	REMARKS
5.0	0.197	0.156	0.065	L = 12.65
6.0	0.239	0.189	0.079	
7.0	0.274	0.217	0.090	
8.0	0.312	0.246	0.102	
9.0	0.350	0.277	0.114	
10.0	0.390	0.309	0.124	
11.0	0.424	0.336	0.139	
12.0	0.459	0.363	0.151	
13.0	0.500	0.395	0.163	
14.0	0.546	0.433	0.179	
15.0	0.579	0.459	0.189	(1)
16.0	0.616	0.487	0.197	
16.5	0.630	0.499	0.197	
17.0	0.645	0.510	0.199	(3)

(Cont'd)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	$(-\Delta P_2)$ (psi)	REMARKS
17.5	0.657	0.518	0.198	L = 12.70
18.0	0.668	0.525	0.198	L = 12.80
19.0	0.668			
20.0	0.672	0.529	0.214	L = 13.0
22.0	0.682	0.534	0.215	L = 13.10
24.0	0.684	0.539	0.221	L = 13.40
27.0	0.687	0.542	0.229	L = 13.70
30.0	0.687	0.544	0.236	L = 14.00
36.0	0.688			

TABLE B-5

WEIGHT OF SAND BED: 51.0 lb

BACKWARD FLOW DATA (See Figure # 7)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	$(-\Delta P_2)$ (psi)	REMARKS
30.0	0.687	0.544	0.236	
27.0	0.687	0.542	0.229	
24.0	0.684	0.539	0.221	L = 13.2
22.0	0.682	0.534	0.215	
20.0	0.668	0.530	0.212	L = 13.0
18.0	0.631	0.506	0.199	
16.0	0.586	0.471	0.192	L = 12.75
15.0	0.563	0.456	0.192	L = 12.70
12.0	0.466	0.379	0.162	L = 12.60
10.0	0.394	0.318	0.138	
9.0	0.355	0.288	0.125	
7.0	0.281	0.228	0.100	L = 12.55
5.0	0.204	0.165	0.072	

TABLE B-6

WEIGHT OF SAND BED: 76.5 lb

BULK DENSITY IN THE FIXED BED REGION: 93.8 $\frac{lb}{ft^3}$

FORWARD FLOW DATA (See Figure # 8)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	$(-\Delta P_2)$ (psi)	$(-\Delta P_3)$ (psi)	REMARKS
5.0	0.293	0.253	0.162	0.065	L = 19.0
6.0	0.356	0.307	0.206	0.079	
7.0	0.410	0.354	0.227	0.091	
8.0	0.459	0.398	0.254	0.102	
9.0	0.519	0.449	0.285	0.115	
10.0	0.571	0.492	0.315	0.127	
11.0	0.633	0.548	0.350	0.140	
12.0	0.687	0.593	0.379	0.152	
14.0	0.805	0.698	0.446	0.177	
15.0	0.869	0.488	0.488	0.192	
15.5	0.896	0.775	0.498	0.202	(1)
16.0	0.918	0.796	0.509	0.202	
17.5	0.978	0.844	0.529	0.205	
18.0	1.000	0.857	0.536	0.207	
19.0	1.000	0.863	0.541	0.210	L = 19.4
20.0				0.217	

(Cont'd)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	$(-\Delta P_2)$ (psi)	$(-\Delta P_3)$ (psi)	REMARKS
22.0	1.0102	0.869	0.553	0.230	
25.0	1.022	0.876	0.559	0.244	L = 20.1

TABLE B-7

WEIGHT OF SAND BED: 76.5 lb

BULK DENSITY IN THE FIXED BED REGION: 100.6 lb/ft^3

FORWARD FLOW DATA (See Figure # 8)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_2)$ (psi)	REMARKS
5.0	0.418	0.237	L = 17.70
6.0	0.511	0.290	
7.0	0.586	0.332	
8.0	0.655	0.372	
9.0	0.734	0.416	
10.0	0.807	0.455	
11.0	0.874	0.488	
12.0	0.926	0.507	L = 18.0, (1)
13.0	0.971	0.517	L = 18.15
13.5	0.984	0.519	
14.0	0.998	0.523	
15.0	1.005	0.521	
15.50	1.012	0.516	L = 18.50
17.50	1.008	0.514	

TABLE B-8

WEIGHT OF SAND BED: 102 lb

BULK DENSITY IN THE FIXED BED REGION: 95.4 $\frac{lb}{ft^3}$

FORWARD FLOW DATA

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_2)$ (psi)	$(-\Delta P_4)$ (psi)	REMARKS
5.0	0.422	0.283	0.070	L = 24.90
7.0	0.590	0.390	0.095	
9.0	0.753	0.498	0.122	
11.0	0.911	0.602	0.146	
13.0	1.081	0.721	0.176	
15.0	1.225	0.819	0.200	
16.0	1.299	0.854	0.202	
17.0	1.343	0.873	0.214	
18.0	1.343	0.882	0.226	

APPENDIX C

Minimum Fluidization Velocity Data

Part I.

(See Figure # 10)

TABLE C-1

WEIGHT OF SAND BED: 25.5 lb

BULK DENSITY IN THE FIXED BED REGION: 94.3 $\frac{lb}{ft^3}$

ROT RDG	$\frac{(-\Delta P_o)}{0.343}$	$\frac{(-\Delta P_1)}{0.185}$	ROT RDG	$\frac{(-\Delta P_o)}{0.343}$	$\frac{(-\Delta P_1)}{0.185}$
5.0	0.294	0.314	17.5	0.951	0.961
6.0	0.354	0.380	18.0	0.959	0.966
7.0	0.410	0.437	18.5	0.961	0.967
8.0	0.459	0.493	19.0	0.963	0.967
9.0	0.522	0.559	20.0	0.973	0.985
10.0	0.583	0.623	21.0	0.981	1.004
11.0	0.639	0.683	22.0	0.987	1.025
12.0	0.695	0.740	23.0	0.996	1.033
13.0	0.752	0.774	25.0	1.000	1.044
14.0			27.0	1.003	1.062
15.0	0.869	0.927	29.0	1.007	1.075
16.0	0.918	0.974	31.0	1.007	1.077
17.0	0.949	0.975	34.0	1.009	1.079

TABLE C-2

WEIGHT OF SAND BED: 25.5 lb

BULK DENSITY IN THE FIXED BED REGION: 108.0 lb/ft^3

ROT RDG	$\frac{(-\Delta P_o)}{0.343}$	ROT RDG	$\frac{(-\Delta P_o)}{0.344}$
5.0	0.552	19.0	0.959
6.0	0.664	20.0	0.968
7.0	0.761	21.0	0.974
8.0	0.857	23.0	0.985
9.0	0.926		
10.0	0.962		
11.0	0.984		
12.0	0.950		
13.0			
13.5	0.935		
15.0	0.938		
16.0	0.947		
17.0	0.949		
18.0	0.957		

TABLE C-3

WEIGHT OF SAND BED: 51.0 lb.

BULK DENSITY IN THE FIXED BED REGION: 93.9 lb/ft^3

ROT RDG	$\frac{(-\Delta P_o)}{0.687}$	$\frac{(-\Delta P_1)}{0.530}$	$\frac{(-\Delta P_2)}{0.198}$	ROT RDG	$\frac{(-\Delta P_o)}{0.687}$	$\frac{(-\Delta P_1)}{0.523}$	$\frac{(-\Delta P_2)}{0.198}$
5.0	0.287	0.294	0.327	18.0	0.973	0.991	1.001
6.0	0.348	0.358	0.396	19.0	0.973	0.998	1.008
7.0	0.399	0.410	0.455	22.0	0.992	1.008	1.085
8.0	0.454	0.465	0.514	24.0	0.996	1.018	1.115
9.0	0.509	0.522	0.577	27.0	0.999	1.024	1.153
10.0	0.567	0.582	0.627	30.0	1.000	1.027	1.190
11.0	0.617	0.634	0.703	36.0	1.016		
12.0	0.668	0.686	0.761				
13.0	0.727	0.746	0.822				
14.0	0.794	0.817	0.904				
15.0	0.843	0.866	0.953				
16.0	0.897	0.919	0.991				
17.0	0.938	0.962	1.005				
17.5	0.956	0.979	1.001				

TABLE C-4

WEIGHT OF BED SAND: 76.5 lb

BULK DENSITY IN THE FIXED BED REGION: 93.8 lb/ft^3

ROT RDG	$(-\Delta P_o)$ 1.031	$(-\Delta P_1)$ 0.874	$(-\Delta P_2)$ 0.543	$(-\Delta P_3)$ 0.214
5.0	0.285	0.289	0.298	0.304
6.0	0.345	0.351	0.380	0.368
7.0	0.397	0.405	0.418	0.425
8.0	0.445	0.455	0.468	0.474
9.0	0.503	0.514	0.526	0.537
10.0	0.554	0.564	0.584	0.591
11.0	0.614	0.627	0.644	0.654
12.0	0.667	0.679	0.698	0.708
14.0	0.781	0.799	0.823	0.827
15.0	0.843	0.854	0.899	0.895
16.0	0.890	0.911	0.938	0.943
17.5	0.949	0.966	0.975	0.958
18.0	0.970	0.981	0.987	0.966
19.0	0.972	0.988	0.997	0.978
22.0	0.980	0.995	1.019	1.073
25.0	0.992	1.002	1.030	1.136

TABLE C-5

WEIGHT OF SAND BED: 76.5 lb

BULK DENSITY IN THE FIXED BED REGION: 100.6 lb/ft^3

ROT RDG	$\frac{(-\Delta P_o)}{1.031}$	ROT RDG	$\frac{(-\Delta P_o)}{1.031}$
5.0	0.406	15.5	0.982
6.0	0.496	17.0	0.972
7.0	0.567	20	0.982
8.0	0.635		
9.0	0.712		
10.0	0.783		
11.0	0.847		
12.0	0.898		
12.5	0.919		
13.0	0.942		
13.5	0.954		
14.0	0.968		
14.5	0.963		
15.0	0.974		

APPENDIX D

Cylindrical Particles

Part II

TABLE D-1

PARTICLE DENSITY OF ONE-INCH CYLINDRICAL PARTICLES

Particle #	Density (lb/ft^3)	Particle #	Density (lb/ft^3)
1	38.1	14	91.4
2	53.8	15	99.7
3	60.0	16	101.7
4	61.4	17	103.7
5	62.8	18	107.8
6	66.6	19	108.2
7	67.9	20	121.3
8	74.7	21	136.1
9	79.4	22	149.1
10	81.7	23	154.9
11	83.2	24	163.3
12	86.4	25	169.1
13	89.8		

TABLE D-2

PARTICLE DENSITY OF TWO-INCH AND HALF-INCH CYLINDRICAL PARTICLES

Two-inch Cylindrincal Particles		Half-inch Cylindrical Particles	
Particle #	ρ_B (lb/ft^3)	Particle #	ρ_B (lb/ft^3)
A	169.3	E	82.6
B	83.0	F	70.4
C	78.3	G	164.3
D	88.9	H	73.6

APPENDIX E

Mixing Behavior of One-inch Cylindrical Particles in an Air-Sand System

TABLE E-1

Particle #1 (ρ_p = 38.1 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save* (inches)	REMARKS
13.0	12.30	0.0		Bubbling observed around the particles but not sinking occured
14.0		φ: 0.30, 0.35		
		A: 0.30, 0.30		
		B: 0.30, 0.30	0.30	
		C: 0.30, 0.30		
		D: 0.30, 0.30		
14.6	Sinks to a maximum of 0.5 inches and then turns over horizontally and floats on top the bed (about 1/4 submerged).			

At higher flow rates, the behavior is similar to 14.6 ft^3/min.

* Save = average value of S

TABLE E-2

Particle # 9 (ρ_p = 79.4 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
11.0	12.30			Bubbling observed around the particle but no sinking occured.
12.0		φ: 0.2 , 0.2 A: 0.2 , 0.15 B: 0.1 , 0.15 C: 0.2 , 0.15 D: 0.25, 0.15	0.19	Sinks straight down
13.0		φ: 0.7 , 0.8 A: 0.65, 0.8 B: 0.45, 0.6 C: 0.7 , 0.5 D: 0.6 , 0.8	0.66	Sinks straight down

14.0 Sinks straight down until covered and then turns over and floats horizontally (about 3/5 submerged)

At higher flow rates, similar behavior as 14.0 ft^3/min.

TABLE E-3

Particle # 13 ($\rho_p = 89.8$ lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	S_{ave} (inches)	REMARKS
11.0	12.30			Bubbling observed around the particle but no sinking occured.
12.0		φ: 0.3 , 0.3 A: 0.3 , 0.2 B: 0.2 , 0.2 C: 0.25, 0.15 D: 0.2 , 0.1	0.22	Sinks straight down
13.0		φ: 0.8 , 0.8 A: 0.8 , 0.85 B: 0.65, 0.8 C: 0.8 , 0.8 D: 0.7 , 0.6	0.76	Sinks straight down

14.0 Sinks below the surface of the bed ($S_{ave} \approx 1.2$ inches) and tends to float under at a slight inclination.

14.6-16.4 Difficult to obtain a steady reading due to oscillation of the particle. Sinks until covered, then adopts a slight horizontal position and sinks with up and down movements.

17.2 After oscillating in the upper half of the bed it sinks to about 9.5 inches below the surface of the bed and remains there.

(Cont'd)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
18				Slowly sinks to bottom
19				and higher flow rates: slowly sinks to bottom.

TABLE E-4

Particle # 19 (ρ_p = 108.2 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
11.0	12.30			Bubbling observed around particle but no sinking
12.0		φ: 0.25, 0.20		
		A: 0.30, 0.30		
		B: 0.35, 0.25	0.24	Sinks straight down
		C: 0.30, 0.25		
		D: 0.15, 0.15		
13.0		φ: 0.7 , 0.9		
		A: 0.9 , 1.0		
		B: 0.9 , 0.8	0.8	Sinks straight down
		C: 0.7 , 0.6		
		D: 0.8 , 0.7		
14.0		φ: 1.2 , 1.3		
		A: 1.3 , 1.3		
		B: 1.6 , 1.4	1.4	
		C: 1.3 , 1.3		
		D: 1.7 , 1.5		
14.6		φ: 3.0 , 2.5		
		A: 2.0 , 2.0		
		B: 2.0 , 1.8	2.1	

(Cont'd)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
		C: 1.9 , 1.8		
		D: 1.8 , 2.0		
15.5		φ: 6.0 , 7.5		
		A: 5.5 , 6.0		
		B: 4.5 , 4.3	5.2	
		C: 3.5 , 4.0		
		D: 6.0 , 4.7		
17.2		Particle sank to bottom		

TABLE E-5

Particle # 20 (ρ_B = 121.3 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
11.0	12.30			Bubbling observed but no sinking occured
12.0		φ : 0.4 , 0.4		
		A : 0.6 , 0.4		
		B : 0.3 , 0.3	0.3	
		C : 0.3 , 0.3		
		D : 0.1 , 0.1		
13.0		φ : 0.9 , 0.8		
		A : 0.9 , 0.9		
		B : 0.9 , 0.9	0.9	
		C : 0.9 , 0.8		
		D : 0.8 , 0.9		
14.0		φ : 1.5 , 1.3		
		A : 1.5 , 1.4		
		B : 1.3 , 1.5	1.4	
		C : 1.3 , 1.4		
		D : 1.3 , 1.4		
14.6		φ : 2.0 , 2.5		
		A : 2.4 , 2.2		
		B : 1.8 , 1.9	2.0	
		C : 1.9 , 1.8		
		D : 1.8 , 2.0		

(Cont'd)

Q (ft^3/min)	L (inches)	S (inches)	S_{ave} (inches)	REMARKS
15.5	12.4	φ: 4.3 , 3.0		
		A: 3.5 , 4.5		
		B: 3.0 , 4.0	3.7	
		C: 3.0 , 4.3		
		D: 3.7 , 4.5		
17.2	Particle sank to bottom			

TABLE E-6

Particles # 11 (ρ_B = 83.2 lb/ft^3) and 12 (86.4 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
Very difficult to distinguish between these two particles				
14.0				Sinks until covered; slight tendency to adopt a horizontal position.
14.6				Sinks to about 1.2 inches after adopting a horizontal position.
18.0				Both particles move up and down in the upper half of the bed.
20.0				Both particles mix well with the bed. It is difficult to pinpoint their location at a particular time. When particles rise to the bed usually aided by upcoming bubbles.
22.2				There is a slight difference in the mixing behavior of the particles. Particle # 12 tends to sink to the bottom and is not brought to the top by bubbles as easily as before. Whereas particle # 11 prefers the upper half of the bed.

At higher readings, they are both nicely mixed with the bed.

APPENDIX F

Mixing Behavior of Two-inch Cylindrical Particles in an Air-Sand System

TABLE F-1

Particle # A ($\rho_B \doteq 169.3$ lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
8.0	12.40			Bubbling observed but no sinking occurred
9.0		φ : 0.20, 0.20		
		A : 0.20, 0.20		
		B : 0.20, 0.20	0.20	
		C : 0.20, 0.20		
		D : 0.20, 0.20		
10.0		φ : 0.40, 0.50		
		A : 0.50, 0.40		
		B : 0.50, 0.50	0.50	
		C : 0.40, 0.50		
		D : 0.50, 0.50		
11.0		φ : 0.70, 0.60		
		A : 0.70, 0.70		
		B : 0.70, 0.70	0.70	
		C : 0.70, 0.70		
		D : 0.60, 0.70		
12.0		φ : 0.80, 1.00		
		A : 0.90, 1.00		
		B : 1.00, 1.00	1.00	
		C : 1.00, 1.00		
		D : 1.00, 1.00		

(Cont'd)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
13.0		φ: 1.2 , 1.5		
		A: 1.4 , 1.4		
		B: 1.4 , 1.3	1.4	
		C: 1.4 , 1.5		
		D: 1.4 , 1.4		
14.0		φ: 2.0 , 2.5		
		A: 2.5 , 2.5		
		B: 3.0 , 2.5	2.5	
		C: 2.5 , 2.5		
		D: 2.0 , 2.5		
14.6		φ: 3.5 , 3.7		
		A: 4.0 , 4.1		
		B: 4.0 , 3.7	3.7	
		C: 3.5 , 3.5		
		D: 4.0 , 3.5		
15.5		φ: 5.0 , 5.5		
		A: 6.5 , 6.0		
		B: 7.0 , 6.5	6.0	
		C: 6.0 , 6.5		
		D: 5.5 , 6.0		
16.4			9.5	
17.2	12.8	Sank to bottom		

TABLE F-2

Particle # B ($\rho_B \doteq 83.0\ lb/ft^3$)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
9.0	12.4			Bubbling observed around particle but no bubbling occurred
10.0		φ: 0.10, 0.10		
		A: 0.10, 0.10		
		B: 0.10, 0.10	0.10	
		C: 0.10, 0.10		
		D: 0.10, 0.10		
11.0		φ: 0.40, 0.50		
		A: 0.60, 0.60		
		B: 0.50, 0.50	0.50	
		C: 0.70, 0.60		
		D: 0.50, 0.40		
12.0		φ: 0.70, 0.80		
		A: 0.80, 0.80		
		B: 0.90, 0.80	0.80	
		C: 0.80, 0.70		
		D: 0.70, 0.80		
13.0		φ: 1.0 , 1.0		
		A: 1.2 , 1.3		
		B: 1.5 , 1.2	1.2	
		C: 1.3 , 1.2		
		D: 1.3 , 1.1		

(Cont'd)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
14		φ: 2.0 , 2.0 A: 2.0 , 2.0 B: 2.0 , 2.0 C: 2.0 , 2.0 D: 2.0 , 2.0	2.0	
14.6		Sinks until covered, then turns over and floats almost completely covered.		

TABLE F-3

Particle # C (ρ_B = 78.3 lb/ft^3)

$\dot{Q}$ (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
10	12.4		0.1	
11.0		φ: 0.40, 0.40		
		A: 0.40, 0.40		
		B: 0.30, 0.30	0.40	
		C: 0.40, 0.40		
		D: 0.40, 0.40		
12.0		φ: 0.70, 0.80		
		A: 0.80, 0.80		
		B: 0.80, 0.80	0.80	
		C: 0.90, 0.90		
		D: 0.70, 0.80		
13.0		φ: 1.0 , 1.0		
		A: 1.3 , 1.2		
		B: 1.3 , 1.1	1.20	
		C: 1.4 , 1.2		
		D: 1.2 , 1.2		
14.0	Sinks until covered and then turns over and floats.			
14.6	$\frac{3}{4}$ submerged and floats horizontally.			
15.5 - 25	Similar behavior as 14.6 lb/ft^3.			

TABLE F-4

Particle # D (ρ_B = 88.9 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
10.0	12.4		0.10	
11.0		φ: 0.50, 0.50		
		A: 0.50, 0.60		
		B: 0.50, 0.50	0.50	
		C: 0.60, 0.50		
		D: 0.40, 0.50		
12.0		φ: 0.80, 0.80		
		A: 0.80, 0.80		
		B: 0.90, 0.90	0.80	
		C: 0.70, 0.80		
		D: 0.80, 0.80		
13.0		φ: 1.1 , 1.2		
		A: 1.2 , 1.2		
		B: 1.4 , 1.3	1.3	
		C: 1.5 , 1.3		
		D: 1.2 , 1.2		
14.0		φ: 2.0 , 2.0		
		A: 2.5 , 2.0		
		B: 2.0 , 2.0	2.1	
		C: 2.3 , 2.2		
		D: 2.0 , 2.0		

(Cont'd)

Q (ft^3/min)	L (inches)	S (inches)	Save (inches)	REMARKS
14.6				Sinks very slowly (after adopting a horizontal position) to 6.0 inches.
15.5				Particle sinks until covered, then adopts a horizontal position and sinks slowly to 10.5 inches.
16.4				Sinks slowly to 12.0 inches.
17.2				Sank to bottom with occasional up and down movement.

APPENDIX G

Mixing behavior of $\frac{1}{2}$ inch Cylindrical Particles in an Air-Sand System

TABLE G-1

Particle #	Q (ft^3/min)	L (inches)	Save (inches)	REMARKS
G	up to 14			no sinking
	14.6		0.7	
	15.5		2.0	
	16.4		7.0	
	17.2		Sank to bottom	
E	up to 14			no sinking
	14.6		0.5	
	15.5		Floats completely covered	
	At higher flow rates		Similar to 15.5	
F and H	up to 14			no sinking
	14.6		0.3	
	15.5		Floats half exposed	
	At higher flow rates		Similar to 15.5	

APPENDIX H

Pressure Drop and Bulk Density Data for the Mixing Behavior of Cylindrical Particles in an Air-Sand System

TABLE H-1

WEIGHT OF SAND BED: 51.0 lb

BULK DENSITY IN THE FIXED BED REGION: 96.5 lb/ft^3

FORWARD FLOW DATA (See Figure 11)

ROT RDG	$(-\Delta P_o)$ (psi)	$(-\Delta P_1)$ (psi)	$(-\Delta P_2)$ (psi)	REMARKS
5.0	0.223	0.173	0.062	L = 12.3 inches
7.0	0.308	0.239	0.086	
9.0	0.395	0.307	0.110	
11.0	0.482	0.375	0.134	
13.0	0.569	0.442	0.159	
14.0	0.616	0.481	0.172	
15.0	0.653	0.509	0.182	light beautiful bubbling
16.0	0.675	0.522	0.184	
17.0	0.669	0.517	0.186	L = 12.65 inches
18.0	0.672	0.517	0.192	
19.0	0.676	0.521	0.195	L = 12.80 inches
20.0	0.683	0.521	0.200	
22.0	0.683	0.526	0.218	L = 13.1 inches
25.0	0.685	0.532	0.230	

TABLE H-2

BED DENSITY AS A FUNCTION OF GAS FLOW RATE

	Q (ft^3/min)	P_B (lb/ft^3)	$(-\Delta P/L)$ (lb/ft^3)
up to	14.6	96.5	94.1
	15.5	95.8	
	16.4	95.0	
	17.2	93.9	
	18.0	92.8	
	18.9	92.1	
	20.3	90.7	

APPENDIX I

Mixing Behavior of One-inch Cylindrical Particles in an Air-Glass Beads System. Glass beads density = 2.436 gm/cc, (-60 to +85 Canadian Mesh)

TABLE I-1

Particle # 25 (ρ_p = 169.1 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	REMARKS
1.60		0.10	
1.86		0.90	
2.25		2.50	
2.90		Sank to bottom	

TABLE I-2

Particle # 11 (ρ_p = 83.2 lb/ft^3) and # 12 (ρ_p = 86.4 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	REMARKS
1.60		0.10	
1.86		0.30	
2.25		1.00	
2.53		1.00	Adopt a slight horizonta position. Float on top of bed almost completel covered.
2.90 and up			Difficult to pinpoint their position. Particle # 11 mixes in the upper half of the bed while # 12 oscillates in the lower half.

TABLE I-3

Particle # 13 (ρ_p = 89.8 lb/ft^3)

Q (ft^3/min)	L (inches)	S (inches)	REMARKS
1.60		0.10	
1.86		0.80	
2.25		1.50	
2.90 and up		Slowly (very) sinks to bottom	

APPENDIX J

Recirculation of One-inch Cylindrical Particles in an Air-Sand System

Part III

TABLE J-1

$Q = 17.2\ ft^3/min$

$q = 2.91\ ft^3/min$

Particle #	t_R (sec) sides	t_R (sec) centre	REMARKS
11	27 - 32	45 - 50	Recirculates
9 and 10			Float. They will both recirculate if pushed 1/2 inch below bed's surface.
13			Recirculates. Difficult to get a stable residence time.
15			Doesn't recirculate. Sinks to bottom.
14			Recirculates. Barel makes it.

Recirculating Density Range: 82-91 lb/ft^3

TABLE J-2

$Q = 17.2 \text{ ft}^3/\text{min}$

$q = 4.31 \text{ ft}^3/\text{min}$

Particle #	t_R (sec) sides	t_R (sec) centre	REMARKS
11	12 - 14	17 - 19	
9	16 - 18	19 - 22	If placed away from recirculation stream (about 1.5 inches from divider) it will float almost completely covered.
5			Recirculates if very close to divider
4			Preference to float.
15	12 - 14	20 - 23	
18			Does not recirculate. Sinks to bottom.
16			Recirculates.
17			Recirculates if caught in main stream (centre) but not at sides.

Recirculating Density Range: 66 - 104 lb/ft^3

TABLE J-3

$Q = 18.9\ ft^3/min$

$q = 2.91\ ft^3/min$

Particle #	t_R (sec) sides	t_R (sec) centre	REMARKS
13	16 - 19	19 - 22	
4			Floats.
5			Floats at sides. Recirculates at centre if close to partition.
9		18 - 20	Floats at sides.
15 and 16			Recirculate.
17			Can be seen trying to jump over but does not quite make it.

Recirculating Density Range: $68 - 102\ lb/ft^3$

TABLE J-4

$Q = 18.9 \text{ ft}^3/\text{min}$

$q = 4.31 \text{ ft}^3/\text{min}$

Particle #	t_R (sec) sides	t_R (sec) centre	REMARKS
8		13 - 15	Recirculates at sides if close to partition.
4			Floats at sides. Recirculates at centre if close to partition.
18 and 19			Recirculate.
20			Does not recirculate. It will recirlate if placed directly in the auxiliary flow streams.

Recirculating Density Range: 65 - 112 lb/ft^3

TABLE J-5

$Q = 20.3\ ft^3/min$

$q = 2.91\ ft^3/min$

Particle #	t_R (sec) sides	t_R (sec) centre	REMARKS
8		10 - 12	Floats at centre if slightly away from recirculation stream (1.5 inches). Floats at sides.
4 and 5			Floats.
7			Recirculates at centre. Floats at sides.
18	10 - 13	10 - 13	
19			Recirculates.
20			Sinks to bottom. Recirculates if caught in the path of the auxiliary air flow.

Recirculating Density Range: 65 - 112 lb/ft^3

TABLE J-6

$Q = 20.3\ ft^3/min$

$q = 4.31\ ft^3/min$

Particle #	t_R (sec) sides	t_R (sec) centre	REMARKS
8	9 - 11	9 - 11	Floats if away from recirculation stream.
12	8 - 10	8 - 10	
3			Recirculate at centre. Floats at sides.

At these high rates it is difficult to set a specific upper limit. For example, particle # 20 will recirculate at the centre but floats at the sides. As a matter of fact particle # 22 (ρ_P = 150 lb/ft^3) could be recirculated if caught in the recirculation stream. Furthermore, the auxiliary air flow was strong enough to keep particle # 5 (ρ_P = 169 lb/ft^3) suspending, but not strong enough to push it over.

TABLE J-7

APPROXIMATE MASS FLOWRATE OF RECIRCULATING SAND

MAIN FLOWRATE (ft^3/min)	AUXILIARY FLOWRATE (ft^3/min)	SAND FLOWRATE (lb/hr)
18	2.91	1300
18	4.31	1750
20	2.91	2000
20	4.31	3400
22	2.91	2400
22	4.31	3800