# Moving-Bed Solids Flow in an Inclined Pipe Leading Into a Fluidized Bed:

## Part I. Gas Leakage and Pressure Drop.

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For moving-bed solids flow in an inclined pipe leading into a fluidized bed, the pressure drop across the pipe is a retarding force while the gas leakage rate alters the flow properties of the solids. The pressure drop has been related to the pipe exit pressure, which, in turn, can be calculated by integrating the solids bulk density over the bed depth. A correlation for the rate of gas leakage in terms of the geometric features and operating conditions of the system is developed. The correlation agrees with experimental data to within ±25%. Part II of this paper will deal with solids flow.

## SCOPE

Solids transfer in and out of a fluidized bed is a common industrial practice (e.g., Zenz and Othmer 1960, Kunii and Levenspiel 1969). The purpose of the transfer may be for feeding solids into the bed or for circulating solids from one vessel to another, (from a reactor to a regenerator and vice versa). The transfer can be accomplished by one or more of the following three flow regimes: pneumatic conveying, fluidized bed, and moving bed. The advantages of transferring solids as a moving bed over pneumatic conveying are that no injection of gas is required in the transfer line and that solids attrition can be reduced. Further, if solids are likely to agglomerate in a fluidized bed reactor, the transfer of solids as a moving bed is preferable, because pneumatic conveying is not suitable for carrying coarse solids. Recently, fluidized bed processes in which solids are transferred between reactors as a moving bed have been proposed (Bailie 1974, Kunii et al. 1974, Engler et al. 1975) and have been successfully operated (Kunii et al. 1974, 1977).

Despite the successful applications of the moving-bed transfer of solids between fluidized bed, published literature on the fundamental aspects of this subject is scarce. Experiments by Trees (1962) appear to be the first among such works. He examined the effects of the pipe diameter, length, and angle of inclination, and the back pressure on the rate of solids flow. The solids flow rate was empirically correlated with these parameters. In addition, he studied the effects of the fluidizing gas flow rate and pipe-exit type on the rate of solids flow. He found that increasing the air rate to the upper bed (from which the solids flowed) had little effect on the solids flow rate. But, for the range studied, increasing the air flow rate to the lower bed (into which the solids flowed) resulted in higher solids flow rates.

The type of pipe exit he studied was characterized by the projected area of the exit onto a horizontal plane. According to Trees, the solids flow rate was roughly proportional to the projected area up to a certain value, beyond which very little increase in flow occurred. Based on the argument that the gas flow rate in the transfer line increases as the projected area increases, Trees qualitatively explained the effect of the pipe-exit type on the solids flow rate, in terms of the gas flow rate in the transfer line. However, in his report, specific values of the rates of gas flow were not cited.

The purpose of this part of our work is to present methods for predicting the back pressure and the rate of gas flow in the transfer line connecting a solids storage bin and a fluidized bed. The predictions are given in terms of the design and operating parameters of the system, namely, the transfer line length, particle size, fluidizing gas velocity, fluidized bed height, feed position, and free board pressure.

## CONCLUSIONS AND SIGNIFICANCE

The pressure drop and the rate of gas leakage, which are two important factors influencing the solids flow, were studied in this paper. It was found that the pressure drop across the pipe is closely related to the pressure at the pipe exit in the fluidized bed. As for the pressure in the fluidized bed, data show that the pressure variation in the radial direction of a fluidized bed is insignificant. On the other hand, the pressure in the axial direction varies with the bed depth in a manner similar to the static head in a fluid. Therefore, existing correlations for axial solids density distribution in a fluidized bed have been used to predict the pressure variation with bed depth; the predictions agree well with data.

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The rate of gas leakage was correlated with the variables studied in this work with the aid of dimensional analysis and the non-linear parameter estimation technique. Correlations show that the gas leakage rate is essentially proportional to the fluidizing gas velocity. It increases as the elevation of the transfer line exit increases, but decreases with an increase in the length of the transfer line. Smaller particles give a smaller rate of gas leakage. The free board pressure is an insignificant parameter. In addition to the correlated parameters, the design of the gas distributor also shows a significant effect on the rate of gas leakage. These results can provide guidelines for the design and operation of moving bed flow systems.

Solids transfer in and out of a fluidized bed is a common industrial practice. Recently, fluidized bed processes in which solids are transferred between reactors as a moving bed have been proposed and successfully operated (Kunii et al. 1974, 1977). Despite the successful applications, published literature on this topic is scarce. Experiments by Trees (1962) appear to be the first among such works. He examined the effects of pipe diameter, length, and angle of inclination, and back pressure on the rate of solids flow. In addition, he qualitatively explained the effect of the pipe-exit type on the solids flow rate, in terms of the rates of gas flow in the transfer line. In his report, however, specific values of the rates of gas flow were not cited.

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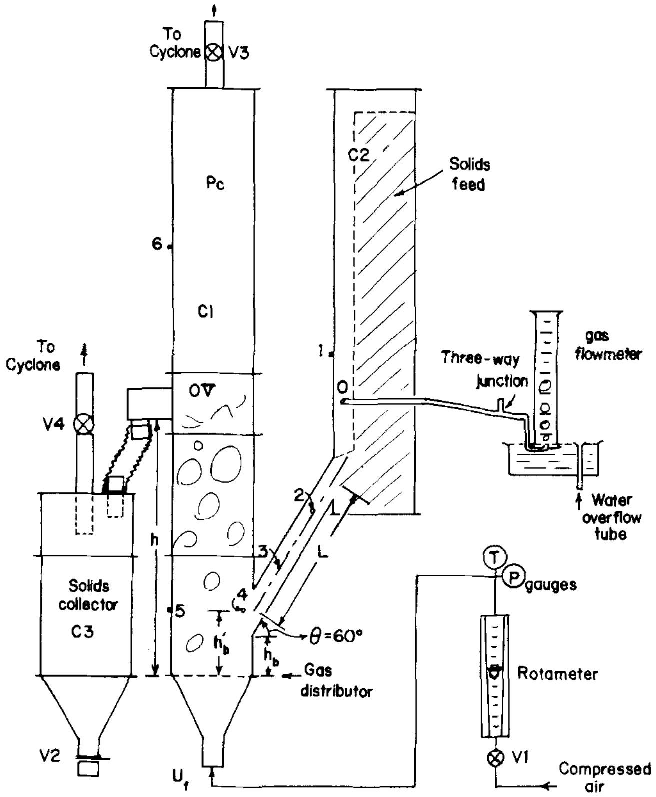


Figure 1. A schematic diagram of the experimental setup.

## SETUP

The columns and transfer lines were fabricated from plexiglas to permit observation. Figure 1 is a schematic diagram of the experimental setup. The center column, C1, is a fluidized bed (0.203 m I.D.) with an overflow section, OV. The outlet of the overflow is located at a height *h*, relative to the distributor plate. Air is supplied to the fluidized bed through a rotameter calibrated at standard conditions. Air temperature and pressure were measured to permit correction for deviations from standard conditions. Solids (sand) were fed into the fluidized bed through a transfer line which made a 60° angle to the horizontal. The transfer line has an I.D. of 0.051 m. The

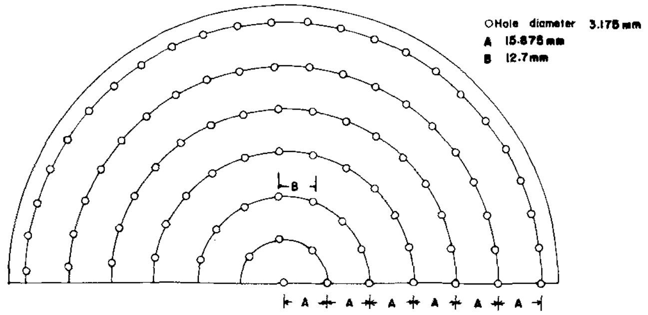


Figure 2a. Hole layout for distributor 10.

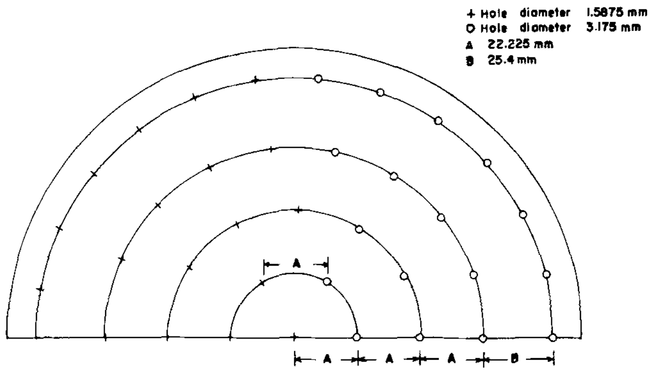


Figure 2b. Hole layout for distributor 7.

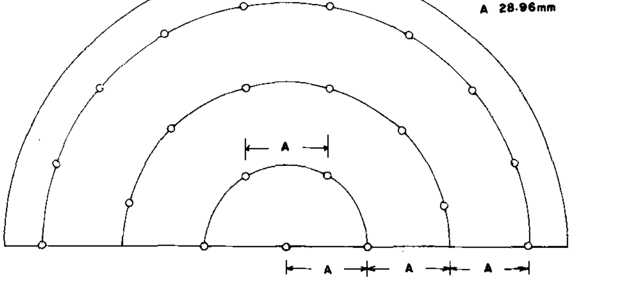


Figure 2c. Hole layout for distributor 1.

TABLE 1. RANGES OF THE PARAMETERS

| Parameters | Variations |
|---|---|
| $U_f$, m/sec | 0.350-1.00 |
| $h_b$, m | 0.045 and 0.197 |
| $h$, m | 0.38 and 0.68 |
| $L$, m | 0.44 and 0.62 |
| $P_c$, KPa(gauge) | 0-6.22 |
| $\overline{d}_p$, m | 0.00051 and 0.00029 |

TABLE 2. PHYSICAL PROPERTIES OF SAND PARTICLES

| | No. I particles | No. II particles |
|---|---|---|
| $\overline{d}_p$, m | $5.1 \times 10^{-4}$ | $2.9 \times 10^{-4}$ |
| $U_{mf}$, m/s | 0.198 | 0.059 |
| $\rho_b$, Kg/m$^3$ | 1,560 | 1,520 |
| $\rho_s$, Kg/m$^3$ | 2,635 | 2,615 |
| $\psi$ | 0.85 | 0.77 |
| angle of repose | 48° | 48° |

upper end of the transfer line is connected to the solids storage column, C2, which also has an I.D. of 0.203 m and contained an inner column with an O.D. of 0.178 m. Solid particles were stored in the inner column (maximum height 1 m). The outer column was sealed so that gas flowing into the solids storage column could only vent into the atmosphere through the tubing at position 0. The three-way junction in the line connecting column C2 and the gas flowmeter permits the measurement of the gas leakage rate in the transfer line. Details of the gas flowmeter are described later. Solids flowing out of the fluidized bed were collected in collector C3 with valve V2 closed. Valves V3 and V4 were used for controlling the pressure, $P_c$, in the free board section. Manometers are connected to positions 1-6 for monitoring the pressures at these positions. Small pieces of cloth were used to cover the openings of pressure taps to prevent sand from entering the manometers.

Perforated aluminum plates of 0.00158 m (1/16″) thickness were used as gas distributors. Three types of distributors were tested. Their geometric features are shown in Figures 2a, b, c. During operation, the #7 distributor was oriented so that the side with the smaller hole size was adjacent to the transfer line.

## PROCEDURE

Based on the results of a preliminary study (Chen et al. 1978), the following operating and design parameters were selected for study (see Figure 1):

1) $U_f$, the superficial fluidizing gas velocity,
2) $h_b$, height of the bottom of the transfer-line relative to the distributor plate,
3) $h$, height of the overflow relative to the distributor,
4) $L$, effective length of the transfer line,
5) $P_c$, pressure in the free board,
6) $\overline{d}_p$, the harmonic mean particle diameter.

The range of these parameters are listed in Table 1.

In the conduct of an experiment, the air flow was regulated by valve V1 and the pressure, $P_c$, was set by adjusting valves V3 and V4. The pressure at position 5 and the bed height were observed to ensure that steady state was reached. Valve V2 at the bottom of solids collector was opened to clear any solids inside the collector and then shut. At the moment valve V2 closed, a stop watch began to measure the solids overflow rate. After the timing started but before all the solids emptied from column C2, the rate of gas leakage through the transfer line and the pressures at positions 1-6 were measured. In addition, we took readings of the rotameter, the temperature gauge, and the pressure gauge. Finally, when the solids level in C2 was about to reach the upper end of the transfer line, the air flow was suddenly terminated by shutting valve V1, and simultaneously we stopped the watch. The solids in C3 were emptied from the collector and weighed.

Since C2 is sealed, except for the transfer line and the outlet tubing, O, the gas flow through the transfer line could be accounted for in two ways. One is the volume of the gas that replaced the space originally occupied by solids in C2. The other is the volume of the gas vented through tubing O. The average rate at which solids volume in column C2 is replaced by gas was calculated from the average solids overflow rate. The average flow rate venting into C2 was measured by the gas flowmeter depicted in Figure 1. The gas flowmeter consists of an inverted graduated cylinder (capacity $5.0 \times 10^{-4}$ m$^3$) and a water container with an overflow tube. It was connected to tubing O through a three-way junction normally vented to the atmosphere. Before the measurement, the graduated cylinder filled with water, with the gas flowing out of column C2 venting to the atmosphere. Measurement started by blocking the vent branch of the three-way junction, so that the gas flowing out of column C2 is fed into the cylinder.

Simultaneously, a stop watch was started. The overflow tube in the container kept the water level essentially constant when water was displaced from the cylinder. When a predetermined amount of water is displaced, one branch of the three-way junction vents to the atmosphere and, simultaneously, the timing is stopped. Three such measurements were taken for each experiment.

As mentioned earlier, the pressure taps at positions 1 through 6 are connected to manometers. Because of the fluctuating nature of the fluidized bed, the water levels in the manometers oscillates. Therefore, the range of the pressure fluctuation was recorded, instead of point values. Positions 2 and 3 measure the pressure drop in the transfer line. In some of the earlier runs, they were connected to two manometers and read separately. Later, they were connected to two arms of the same manometer, and only the pressure difference was read, to minimize the error in reading.

Sand was used as the test material. The pertinent properties are summarized in Table 2.

## DATA ANALYSIS AND CORRELATION

The method of data analysis is described here. Some theoretical aspects of the pressure distribution in a fluidized bed are presented, and a correlation for estimating the rate of gas leakage in the transfer line is developed.

### Pressure

All the recorded pressures fluctuated constantly. Therefore, the pressure data used in the analysis are the midpoints of the range of the recorded pressures.

The term fluidized bed implies that it behaves like a liquid. Therefore, the variation of pressure in the radial direction of the bed is insignificant, and that in the axial direction of the bed is essentially identical to the static head. The pressure at $h_b'$ is:

$$P = \int_{h_b'}^{h} \rho g \, dh + P_c \quad (1)$$

If we assume that the bulk density does not vary axially in the bed we have $\rho = \rho_b$ = constant and

$$P = \rho_b g (h - h_b') + P_c \quad (2)$$

While both $\rho_b$ and $h$ in this equation vary with the change in operating conditions, one of them can be eliminated because of the following material balance:

$$\rho_b A_f h = \rho_{mf} A_f h_{mf} \quad (3)$$

By eliminating $\rho_b$ from Equation (2) by means of Equation (3) we have

$$P = \rho_{mf} g h_{mf} \frac{h - h_b'}{h} + P_c \quad (4)$$

Note that both $\rho_{mf}$ and $h_{mf}$ are independent of the operating conditions.

If the assumption of a constant bulk solids density is not valid, the solids density distribution along the height of the fluidized bed must be employed in Equation (1). The axial solids density distribution in gas-solids fluidized beds was studied by Bakker and Heertjes (1960) and by Fan et al. (1962). According to their works, the density profile may be divided into a lower-density zone and a falling-density zone. Fan et al. (1962) have established the following equation for the density profile in the falling-density zone:

$$\frac{\rho}{\rho_{mf}} = \exp\left[ A + B \left( \frac{h}{h_{mf}} - 1 \right)^2 \right] \quad (5)$$

where $A$ and $B$ are given by:

$$A = -1.396 \times 10^{-5} \left( \frac{\overline{d}_p U_f \rho_g}{\mu_g} \right)^{0.978} \left( \frac{\overline{d}_p}{D_f} \right)^{-1.265} \left( \frac{h_{mf}}{D_f} \right)^{0.347} \quad (6)$$

$$B = 5.492 \times 10^{9} \left( \frac{\overline{d}_p U_f \rho_g}{\mu_g} \right)^{-2.094} \left( \frac{\overline{d}_p}{D_f} \right)^{3.045} \left( \frac{h_{mf}}{D_f} \right)^{0.347} \quad (7)$$

Equation (5) applies to a bed height higher than $h_{mf}$. For a bed height lower than $h_{mf}$, or for the lower-density zone, the density remains fairly constant and is given by

$$\frac{\rho}{\rho_{mf}} = \exp(A) \quad (8)$$

**Gas Leakage**

We did not succeed in our attempt to correlate the present experimental data according to the modified Ergun equation (Yoon and Kunii 1970), which uses the slip velocity in place of the superficial gas velocity in the original Ergun equation. This indicates that the gas flow in the transfer line was highly non-uniform over a cross sectional area. Here, a correlation between the gas leakage rate and the independent variables of the system is developed with the aid of dimensional analysis.

In this treatment, the gas leakage rate is expressed as a superficial velocity based on the cross-sectional area of the transfer line. All the independent variables of the system are listed in Table 3. In the dimensional analysis of the operating parameters of fluidized beds, Broadhurst

TABLE 3. INDEPENDENT VARIABLES

| | |
|---|---|
| Gas properties: | $\rho_g$: density<br>$\mu_g$: viscosity |
| Solid properties: | $\rho_s$: density<br>$\overline{d}_p$: harmonic mean size<br>$\psi$: shape factor |
| Flow system variables: | $D_f$: inside diameter of the fluidized bed<br>$D_t$: inside diameter of the transfer line<br>$h_b'$: defined in Fig. 1<br>$L$: length of the transfer line<br>$\theta$: angle of inclination of the transfer line |
| Operational parameters: | $h$: mean bed height<br>$U_f$: superficial air velocity in the fluidized bed<br>$P_c$: pressure above the fluidized bed |
| Other: | $g$: gravitational acceleration |

and Becker (1973, 1975) recommended the characteristic parameter $\gamma = g(\rho_s - \rho_g)$ to account for the gravitational effect. The present analysis uses this parameter in place of g listed in Table 3. Thus, we can write:

$$U_g = \phi \{\rho_g, \mu_g, \rho_s, \overline{d}_p, \psi, D_f, D_t, h_b', L, \theta, h, U_f, P_c, \gamma\} \quad (9)$$

If we eliminate three variables namely, $\rho_g$, $\overline{d}_p$, and $\gamma$ which are dimensionally independent, Equation (9) can be expressed in dimensionless form as:

$$\frac{\rho_g U_g d_p}{\mu_g} = \phi_1 \left\{ \frac{\rho_g U_f d_p}{\mu_g}, \frac{h_b'}{d_p}, \frac{L}{d_p}, \frac{h}{d_p}, \frac{\rho_g \gamma d_p^3}{\mu_g^2}, \frac{P_c}{\gamma d_p}, \frac{\rho_s}{\rho_g}, \frac{D_t}{d_p}, \frac{D_f}{d_p}, \theta, \psi \right\} \quad (10)$$

Note that the bar over $d_p$, indicating the harmonic mean, is dropped for convenience. The gas leakage was correlated in terms of the first six terms on the right-hand side of Equation (10) because the last five terms are essentially constant in all experiments. The correlation is assumed to have the form:

$$\frac{\rho_g U_g d_p}{\mu_g} = a_0 \left( \frac{\rho_g U_f d_p}{\mu_g} \right) a^1 \left( \frac{h_b'}{d_p} \right) a^2 \left( \frac{L}{d_p} \right) a^3 \left( \frac{h}{d_p} \right) a^4 \left( \frac{\rho_g \gamma d_p^3}{\mu_g^2} \right) a^5 \left( \frac{P_c}{\gamma d_p} \right) a^6 \quad (11)$$

The coefficients, $a_0$ through $a_6$, can be determined by means of a nonlinear parameter estimation technique (Bard 1967).

## RESULTS AND DISCUSSION

Figure 3 compares the pressure drop per unit length of the transfer line between positions 2 and 3 and that between the two ends of the transfer line (i.e., positions 1 and 4). As can be seen, the two measurements yield essentially the same value, considering the fact that the pressure fluctuation about the median varied from ±5% to ± 33%. The greater scatter (double in percentage deviation) at higher values of the pressure drop can be attributed to the doubled error induced in the earlier runs, when the pressures at positions 2 and 3 were measured separately.

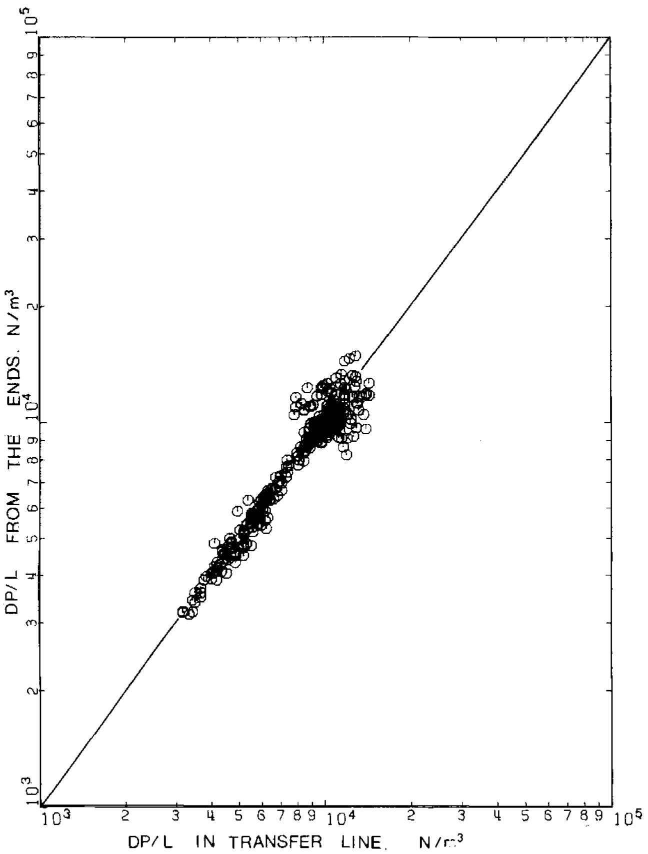


**Figure 3. Comparison between pressure drop measured in the transfer line and from the two ends of the transfer line.**

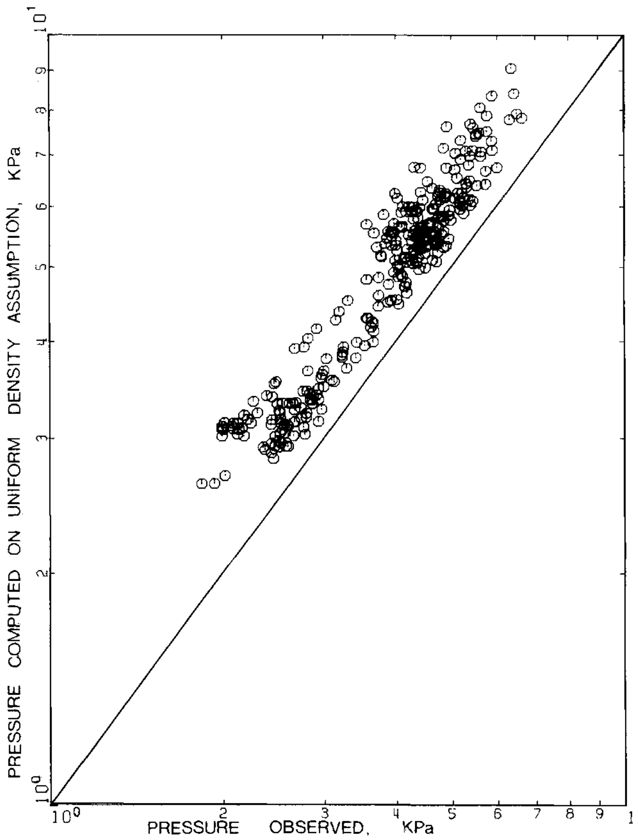


**Figure 4. Pressure calculated based on uniform density assumption.**

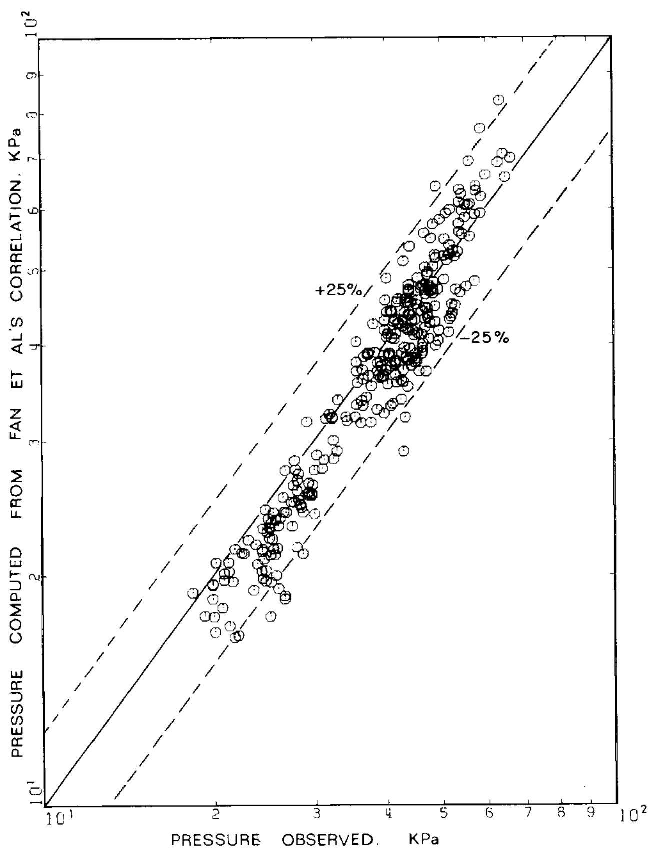


**Figure 5. Pressure calculated by assuming an axial solids density distribution.**

The recorded static pressure data at positions 4 and 5 indicate that the difference between them is negligible. This confirms that the pressure in a fluidized bed does not vary radially over a cross section.

Figure 4 compares the static pressures in the fluidized bed (viz., the pressures at position 4) with those estimated from Equation (4), which is based on the assumption of constant bulk solids density. Obviously, Equation (4) overestimates the pressures. This may arise from the fact that the variation of the bulk density in the axial direction of the fluidized bed was neglected in deriving this equation.

Equation (1), with the density distribution given by Equation (5), was integrated numerically, using Simpson's rule. The calculated results are compared with the experimental data in Figure 5, which shows a satisfactory agreement.

Combining the results in Figures 3 and 5 reveals that the pressure drop in the transfer line can eventually be expressed in terms of the design and operating parameters of the fluidized bed.

Table 4 presents the values of coefficients $a_0$ through $a_6$ in Equation (11), estimated from the gas leakage data of distributor 10. Figure 6 indicates that the data can be

TABLE 4. COEFFICIENTS OF THE CORRELATION

| | |
|---|---|
| $a_0$ | $0.8121 \times 10^{-3}$ |
| $a_1$ | 0.9824 |
| $a_2$ | 0.4364 |
| $a_3$ | −0.3375 |
| $a_4$ | −0.2204 |
| $a_5$ | 0.6022 |
| $a_6$ | −0.0150 |

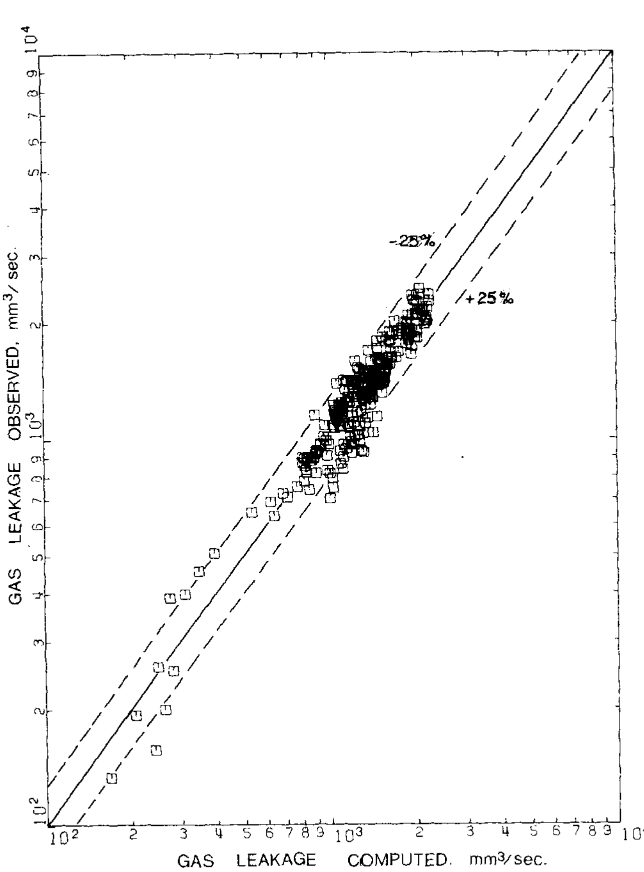


Figure 6. The gas leakage correlation.

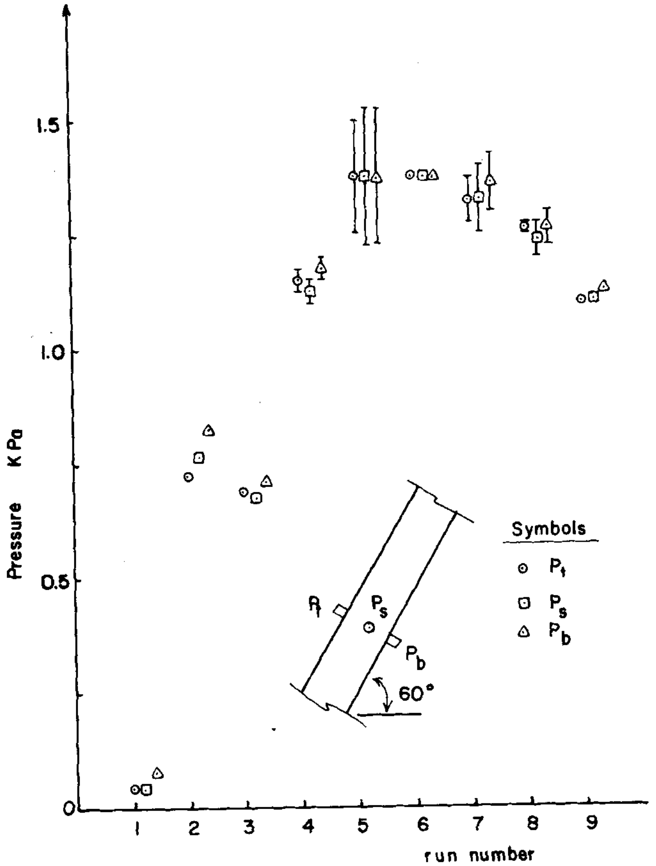


Figure 7. Comparison of the pressures measured at different locations on the transfer line.

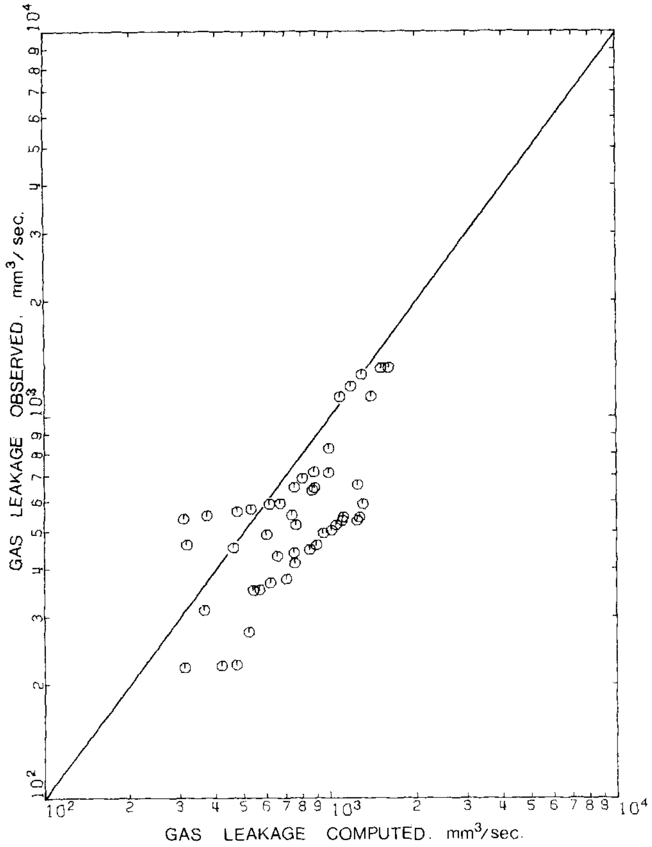


Figure 8. Results from distributor #1.

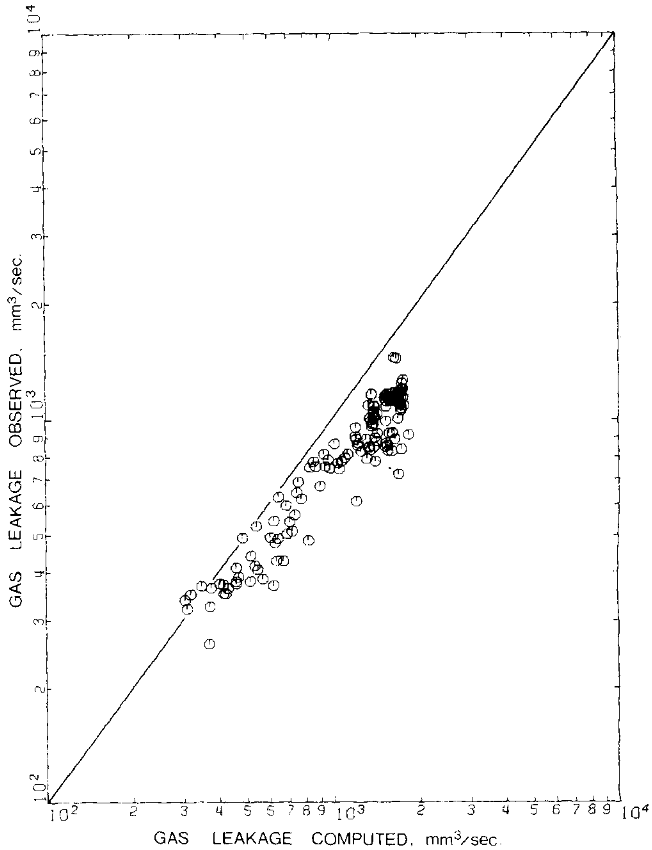


Figure 9. Results from distributor #7.

correlated reasonably well by Equation (11), because more than 90% of the data deviate less than ± 25% from the correlation.

The value of $a_1$ in Table 4 reveals that the rate of gas leakage, is essentially proportional to the fluidizing gas velocity, $U_f$. And, the value of $a_2$ indicates that the rate of gas leakage increases as the elevation of the transfer line exit, $h_b'$ goes up. Unlike these variables, an increase in the fluidized bed height, $h$, reduces the gas leakage rate, as shown by the negative value of $a_4$ in the table. For the range studied, the pressure in the free board, $P_c$, is an insignificant parameter, shown by the value of $a_6$. Besides the influence of the three variables pertaining to the fluidized bed, those from the length of the transfer line and from the particle size can be observed from the values of the parameters given in the table. The rate of gas leakage is inversely proportional to roughly the one-third power (i.e., $a_3 = -0.3375$) of the length of the transfer line. For the two sizes of particles studied, the smaller one results in a smaller rate of gas leakage ($a_5$ is positive).

Since the rate of gas flow in the transfer line was highly non-uniform, we investigated the possibility of a non-uniform pressure distribution over the cross-section. This was done by measuring the pressure at three locations of the same cross section of the transfer line for sets of different operating conditions (shown in Figure 7). The results here show that, although the gas flow is non-uniform in the transfer line, the pressure did not vary significantly in the radial direction.

The rates of gas leakage obtained with distributors 1 and 7 are compared respectively in Figures 8 and 9, with the predictions based on Equation (11). These figures reveal that the correlation overpredicts the gas leakage rate. Distributors 1 and 7 have approximately the same hole area, but the area is considerably lower than that for distributor 10. The hole area ratio (10:7 or 1) is about 4.6. Distributors 1 and 7 are further distinguished by the fact that the hole area is distributed non-uniformly in the latter. A definite trend in the results presented in Figures 8 and 9 is not evident. However, both cases show a significant deviation from Figure 6. These results suggest that distributor design (at least in terms of hole area) is an influential feature not to be neglected.


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## NOTATION

$A$ = parameter defined by Equation (6)
$A_f$ = cross-sectional area of the fluidized bed
$B$ = parameter defined by Equation (7)
$D_f$ = inside diameter of the fluidized bed
$D_t$ = inside diameter of the transfer line
$d_p$, $\overline{d}_p$ = harmonic mean particle diameter
$g$ = acceleration of gravity
$h$ = fluidized-bed height
$h_b$, $h_b'$ = height of transfer-line exit relative to the distributor (see Figure 1)
$h_{mf}$ = fluidized-bed height at minimum fluidization
$L$ = effective length of the transfer line
$P$ = pressure
$P_c$ = pressure in the free board
$U_f$ = superficial fluidizing gas velocity
$U_g$ = superficial gas velocity in the transfer line
$U_{mf}$ = superficial velocity of minimum fluidization

### Greek Letters

$\gamma$ = $g(\rho_s - \rho_g)$
$\theta$ = angle of inclination from the horizontal
$\mu_g$ = gas viscosity
$\rho$ = density
$\rho_b$ = solids bulk density
$\rho_g$ = gas density
$\rho_s$ = solids density
$\rho_{mf}$ = solids bulk density at minimum fluidization condition
$\phi$, $\phi_1$ = function
$\psi$ = shape factor or the quotient of the area of a sphere equivalent to the volume of the particle divided by the actual surface area of the particle

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