An Eddy Viscosity Model for Friction in Gas—Solids Flow

F. M. JULIAN and A. E. DUKLER
University of Houston, Houston, Texas

The equations normally used to predict velocity distribution and eddy viscosity in single-phase flow systems can be adapted to fit gas-solids flow systems by including a term to account for the quantity of solid matter carried by the gas stream. This solids loading is expressed as pounds of solid per pound of gas. These modified equations can then be used to calculate the pressure drop of a solids-laden gas stream flowing in a pipe by means of a two-phase friction factor. Comparison of this approach with available pressure drop data is used to check its validity and to evaluate the constants in the equations.

One of the problems that has challenged engineers for many years is that of transporting solid, particulate materials. Since such materials cannot be easily pumped or blown like fluids, complicated mechanical equipment has usually been required to move them. At the turn of the century, engineers began to experiment with pneumatic conveying of finely ground solids. This method had so many distinct advantages of capacity, simplicity, reliability, and containment of the particles, that it soon found widespread acceptance throughout industry. One of the principal uses in the early days, as now, was the loading and unloading of grain from ships and railroad cars.

Despite the widespread use of pneumatic conveying, there has never been any sound theoretical basis for the design of the equipment required. Originally, design was strictly by trial and error. The theoretical aspects were largely neglected until 1924, when Gasterstadt published the first scientific paper on the subject (*8*). As other investigators entered the field, they concentrated the bulk of their studies on the subject of pressure drop measurement in gas-solids flow systems. They have usually expressed their findings in one of two ways.

Several investigators have made use of a term α, defined as the ratio of the pressure drop of the two-phase system to that of air flowing in the same pipe at the same velocity. This term is equated to an empirical expression based on their experimental results. The form is usually expressed as

$$\alpha = \frac{\Delta P_T}{\Delta P_g} = 1 + F(r)$$

where r is defined as the pounds of solid flowing per pound of gas. Actually the term α has no theoretical significance but is strictly a notation of convenience.

A more logical approach has been to represent the total pressure drop as a sum of several separate and distinct pressure drops, each of which has a definite assignable cause.

$$\Delta P_T = \Delta P_{ag} + \Delta P_{as} + \Delta P_{fg} + \Delta P_{fs} + \Delta P_x$$

Pressure drop due to acceleration of the gas ΔP_{ag} is so small that it may usually be neglected, but acceleration of the solids to flow velocity ΔP_{as} consumes a great deal of energy and must be considered. Figure 1 shows the approximate portion of the total pressure drop at various points, which is due to solids acceleration. This is based on the data of Hinkle (*12*), who covered the effect of particle acceleration on pressure drop very thoroughly.

F. M. Julian is with E. I. duPont de Nemours, Inc., LaPorte, Texas.

ΔP_x, which represents the presure drop due to the support of an inventory of the two-phase mixture in a vertical pipe, may be calculated from the expression

$$\Delta P_x = G_s\, L/u_s$$

The core of the pressure drop problem is basically to evaluate the two terms for frictional drop: ΔP_{fg} and ΔP_{fs}.

Many authors assume that the pressure drop of the gas ΔP_{fg} is the same as if the gas were flowing alone in the pipe, and include the observed increased pressure drop in the solids friction term, ΔP_{fs}. Methods have been proposed for correlating the *excess pressure drop*. This is frequently done by defining a solids friction factor f_s in terms of the excess pressure drop, some characteristic velocity and the particle, and/or channel dimensions. Attempts are then made to define a Reynolds number that will correlate f_s in a manner similar to that for single-phase flow. Examples of this attack can be found in references *2, 4, 6, 10, 11*, and *17*. It should be noted that this approach attempts to evolve a correlation without considering the local conditions of flow.

Other studies have been directed toward evolving a detailed understanding of single particle and group particle behavior in fluid-particle flow situations. A detailed

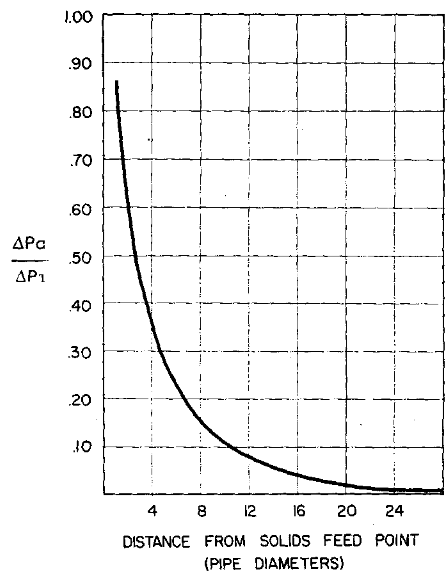

Fig. 1. Contribution of particle acceleration to total pressure drop.

treatment of single particle behavior was reported by Gauvin (*18*). Kada and Hanratty (*14*) explored the effect of the presence of the solid particles on the local turbulence in the fluid. A continuing study by Soo has reported on studies of local particle diffusion, particle-particle collisions, wall-particle interactions, local concentrations, and fluid-particle slip. A partial selection of reports of direct interest to this work appears in references *19* through *23*. While these studies have contributed much to our understanding of local particle behavior, the theories are not yet complete enough to permit the prediction of pressure drop in conduit flow. Such an extension will require a generalized method for predicting local drag on each particle, radial concentration distribution of particles, radial distribution of particle velocity, and information on the effect of particle concentration on the local turbulence intensity.

What appears to be required at this stage of our knowledge is a model that uses local flow conditions as its starting point (in contrast to the excess pressure drop approach), but that resolves unknown conditions of local particle behavior by use of phenomenological methods. Such methods have been successfully applied to simple problems of turbulent flow in conduits. As this study will show, they appear to work with good success in fluid-particle systems as well.

A THEORY FOR FRICTIONAL LOSS

Most previous analyses have assumed that the gas phase friction is unaffected by the presence of the solid, while the apparent increased dissipation of energy is due to form drag around the particle or to the resistance caused by particle to wall or particle to particle collisions. Contrary to this approach, it is suggested here that under conditions where saltation or choking do not occur (homogeneous dilute phase transport), the solids make their influence felt primarily by modifying the local turbulence in the gas phase, increasing the turbulent fluctuations, mixing length, eddy viscosity, and thus the frictional pressure drop. Such a model would seem quite reasonable in view of the fact that in dilute phase transport the pipe volume occupied by the solid phase is very small. For example, when transporting a solid with a density of 100 lb./cu. ft., using low pressure air at a mass ratio of twenty pounds of solid per pound of gas, the volume fraction occupied by the solid is less than 5%. The study of Kada and Hanratty (*14*) provides clear experimental evidence that the presence of solids increases the local eddy viscosity whenever an appreciable slip velocity is present.

A model of this type yields an expression for the frictional contribution to pressure drop that would be the same for vertical and horizontal flow. Total measured pressure drop in vertical flow would, of course, differ from

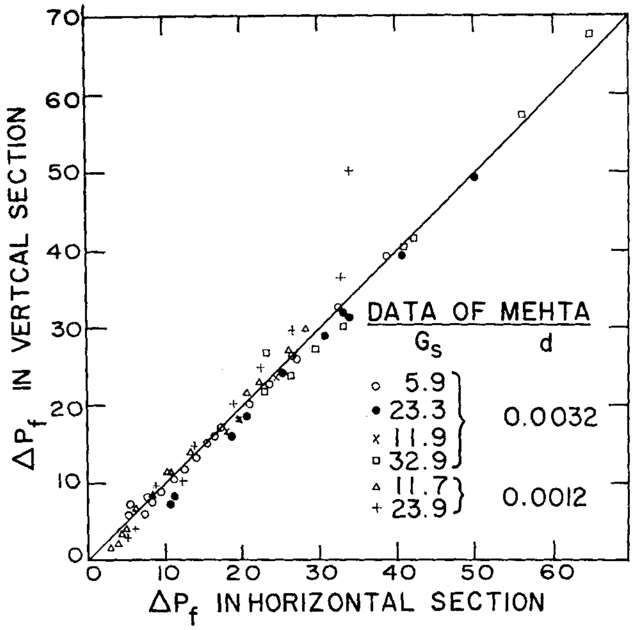

Fig. 2. Comparison of friction in vertical and horizontal flow. Data of Mehta (*17*).

the horizontal by the amount of static head due to holdup. That frictional loss is the same for vertical or horizontal dilute phase transport can be seen by the data of Mehta (*17*). Mehta designed a system where his two-phase stream flowed through a horizontal and a vertical conduit in series. Thus, under identical conditions of flow, he measured pressure drop in both vertical and horizontal sections, along with the holdup in the vertical run. The friction in Mehta's vertical flow (total measured pressure drop less the measured holdup) is compared with horizontal friction in Figure 2. It is quite clear that the friction is the same under these conditions of dilute phase transport (see Table 1).

In single-phase flow, once an expression for the eddy viscosity is available, it is possible to write the equation for the local shear stress in terms of the velocity gradient.

$$\frac{\tau}{\tau_o} = \frac{1}{g_c}\left(1 + \frac{\epsilon\rho}{\mu}\right)\frac{du^+}{dy^+} \qquad (1)$$

By using a momentum balance to obtain an independent equation for the shear stress distribution τ/τ_o Equation (1) can be integrated to give an equation for the local velocity distribution. This can then be integrated across the tube for volumetric flow rate to give an equation relating the friction factor and Reynolds number.

$$\frac{1}{\sqrt{f/2}} = \frac{2}{(y_m^+)^2}\int_o^{y^+_m} (y_m^+ - y^+)\, u^+ dy^+ \qquad (2)$$

TABLE 1. RANGE OF DATA USED FOR CORRELATION EVALUATION

Author	Particle diameter, in.	Particle density, lb./cu. ft.	Pipe diameter, in.	N_{Re}	$1 + r$
Hinkle (*12*)	0.09 to 0.25	69.7 to 113.0	2, 3	178,000	1.5 to 5.5
DePew (*6*)	0.001 to 0.008	161.0	0.71	13,500 to 27,400	1.5 to 4.5
Clark (*4*)	0.04	73.0	1.0	39,000 to 75,000	3.2 to 8.4
Helander (*11*)	0.017 to 0.037	83.0 to 157.0	1½, 2	33,000 to 59,000	1.4 to 13.1
Hariu & Molstad (*10*)	0.008 to 0.02	165.0	¼, ½	3,700 to 11,000	1.5 to 37.0
Mitlin (*18*)	0.005 to 0.06	75.0 to 525.0	1.0	43,000 to 78,000	2.6 to 17.5
Mehta (*17*)	0.0014 and 0.0038	158	½	9,700 to 62,500	1.4 to 5.8
Welschof (*27*)	0.15	165	2.4	42,000 to 113,000	1.3 to 28.4

where

$$y_m^+ = N_{Re}\sqrt{f/8} \quad (2a)$$

and

$$f = \frac{\tau_0}{\frac{\rho \bar{u}^2}{2g_c}} = \frac{(\Delta P/\Delta L)_f D}{\frac{2\rho \bar{u}^2}{g_c}} \quad (3)$$

Numerous expressions have been proposed for the eddy viscosity coefficient, all based on phenomenological reasoning. All depend on experimental data for evaluation of constants and, when used in Equation (2), almost all result in a friction factor-Reynolds number equation for single-phase flow that agrees with experimental data to within 10%.

If, as is assumed here, the presence of a solid increases friction only insofar as it increases local dissipation, then the friction factor for gas-solids flow can be derived from Equations (1) and (2), once an expression for eddy viscosity can be derived suitably modified for the presence of the solid.

A MODIFIED EDDY VISCOSITY

The starting point for this modification is the relationship proposed by Gill and Sher (9) for single-phase flow through a tube:

$$\epsilon = k^2 y^2 \left[1 - \exp\left(-\frac{\phi y}{D/2}\right) \right]^2 \frac{du}{dy} \quad (4)$$

where k is the Von Karman constant that has the value of 0.36 for single-phase flow.

This equation has the advantage of describing the eddy viscosity over the entire flow field in contrast to earlier expressions of Deissler (5), Von Karman (26), and others who proposed equations each valid only in the wall region or in the region of fully developed turbulent flow. Equation (4) can be modified to give

$$\epsilon = \frac{k^2 y^2}{\rho_g} \left[1 - \exp\left(-\frac{\phi y}{D/2}\right) \right]^2 \frac{dG_g}{dy} \quad (5)$$

where G_g is the *local* mass velocity of the gas $\rho_g u_g$. Equation (5) suggests that the local eddy viscosity depends on the gradient of the local mass velocity of the gas. Now it is hypothesized that *for gas-solids flow* the eddy viscosity depends on the gradient of the *total* local mass velocity, *that of gas plus solid,* or

$$\epsilon_{g-s} = \frac{k^2 y^2}{\rho_g} \left[1 - \exp\left(-\frac{\phi y}{D/2}\right) \right]^2 \frac{d}{dy}(G_g + G_s) \quad (6)$$

This equation can be rewritten as

$$\epsilon_{g-s} = k^2 y^2 \left[1 - \exp\left(-\frac{\phi y}{D/2}\right) \right]^2 \frac{d}{dy}[u_g(1 + r)] \quad (7)$$

where r is the local solids to gas ratio. Assuming this *ratio* will remain essentially constant, even if the local gas and solids mass velocities vary along the radius, gives

$$\epsilon_{g-s} = k^2 (1 + r) y^2 \left[1 - \exp\left(-\frac{\phi y}{D/2}\right) \right]^2 \frac{du_g}{dy} \quad (8)$$

Equation (8) implies that the gas and solids mass velocities contribute equally to the local eddy viscosity and the resulting turbulent shear stress. A more general relationship that would provide for unequal effect would be

$$\epsilon_{g-s} = k^2 (1 + r)^m y^2 \left[1 - \exp\left(-\frac{\phi y}{D/2}\right) \right]^2 \frac{du_g}{dy} \quad (9)$$

where the size of the exponent m would be a measure of the influence of the solid phase.

Before proceeding with the development of the velocity distribution and frictional loss relations, the assumptions underlying the model are summarized as follows:

(1) In dilute phase gas-solids flow, the primary mechanism for the observed higher friction is increased local dissipation in the gas phase owing to the presence of the solid. Under these conditions losses due to particle-particle collisions and particle-wall collision are assumed small by comparison.

(2) The particle makes its influence felt through the slip that takes place between particle and fluid. While the solids distribution radially need not be a constant, the ratio of local solids to vapor rates (and thus the slip) is assumed independent of radial position.

The theory applies, of course, only to conditions where saltation and choking do not occur.

VELOCITY DISTRIBUTION

Equation (9) substituted in Equation (1) and combined with the usual shear stress distribution equation for flow in a tube gives the following equation for velocity distribution in dimensionless coordinates.

$$c\left(\frac{du^+}{dy^+}\right)^2 + \frac{du^+}{dy^+} = d \quad (10)$$

where

$$c = (Ky^+)^2 \left[1 - \exp\left(-\frac{\phi y^+}{y_m^+}\right) \right]^2 \quad (11)$$

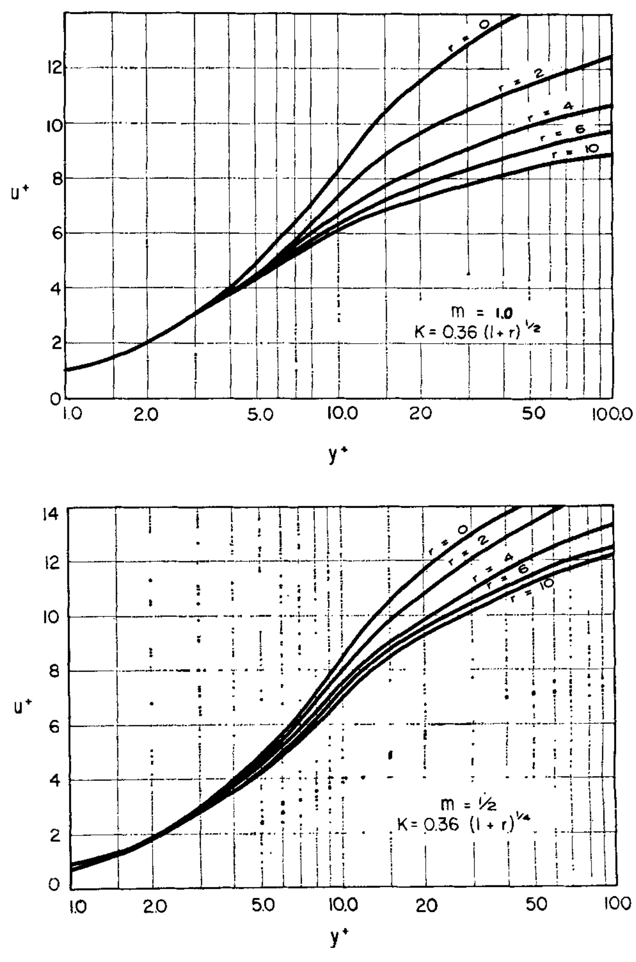

Fig. 3. Velocity distributions predicted by the model.

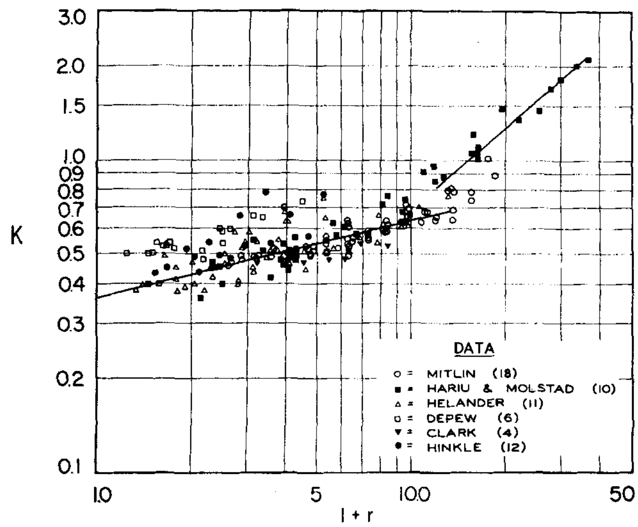

Fig. 4. The modified Von Karman constant as a function of solids loading.

$$K = k(1 + r)^{m/2} \tag{12}$$

$$d = 1 - \frac{y^+}{y_m^+} \tag{13}$$

$$\phi = \frac{y_m^+ - 60}{22} \tag{14}$$

The constant 60 in the numerator of Equation (14) was suggested by Gill and Sher to force the laminar-turbulent transition at a Reynolds number of 1,800. The value of 22 *for the denominator of ϕ was selected to give agreement* with experimental velocity distribution data for single-phase flow. Solving the quadratic for the gradient and integrating gives

$$u^+ = \int_0^{y+} \frac{-1 + (1 + 4cd)^{1/2}}{2c} dy^+ \tag{15}$$

This equation clarifies the effect of the solids loading on the velocity distribution. The entire effect of the solids is isolated in the quantity K, which can be considered a modified Von Karman turbulence constant. Figure 3 shows the results of numerical integrations of Equation (15) for various values of the solids-gas loading and for two possible values of m, the weighting exponent. While there are few data available to test these theoretically derived velocity distributions, the downward displacement of the curves is consistent with the observed increase in pressure drop.

METHOD OF USING PRESSURE DROP DATA TO TEST THE MODEL

In the absence of adequate velocity distribution data, which would permit the most direct test of the model, it is necessary to use frictional pressure drop measurements. Combining Equations (2) and (15) gives

$$\frac{1}{\sqrt{f/2}} = \frac{2}{(y_m^+)^2} \int_0^{y+m} (y_m^+ - y^+) \int_0^{y+} \frac{-1 + (1 + 4cd)^{1/2}}{2c} dy^+ dy^+ \tag{16}$$

An experimental measurement makes it possible to calculate all terms in Equation (16) except c. The measurement of frictional pressure drop along with flow rate in a conduit of known size makes it possible to calculate f from Equation (3), d from (13), y_m^+ from (2*a*), and ϕ from (14). Equation (16) can then be solved, in principle, for c, and a value of the modified Von Karman constant K calculated from Equation (11). If the model is correct, this K calculated from experiment should depend only on $(1 + r)$ as dictated by Equation (12), and K plotted against $(1 + r)$ should be linear on logaritmic coordinates.

RESULTS

Original source data on pressure drop were collected and examined for internal consistency. Data were not used when: insufficient information was available to determine the contributions to measured pressure drop by acceleration or gravity, and when a single or group of measurements showed wide discrepancies relative to data taken by the same author at near conditions that were following systematic trends. A detailed discussion of the data available and an evaluation of them are given in reference *13*. Some idea of the wide range of data used to evaluate the theory can be obtained from Table 1.

For each experimental point the frictional pressure drop was calculated by subtracting acceleration and static terms from the total measured value. All quantities necessary to integration of Equation (16) were then calculated. Equation (16) was then integrated numerically for a series of values of K until one was found that would satisfy the equation. The values of K thus obtained are plotted against $(1 + r)$ in Figure 4. As required by the model, a straight line described the trend of the data well up to a value of r of about 12. The intercept of the best line through the data suggests a value for K of 0.36 at r equal to zero (which is equivalent to the case of single-phase flow). This is the accepted value of the Von Karman constant for single-phase flow. The scatter of the data is such that approximately 90% of the points fall within ± 20% of the line shown. In view of the experimental difficulties involved in measuring the pressure drops, solids concentrations, etc., in two-phase flow systems, it is not surprising that this degree of variation in the data should exist. The fact that the data of eight independent investigators who used widely different apparatus and measuring techniques are coordinated this closely appears to support the proposed correlation.

The slope as measured from this figure is approximately 0.25. Thus for r from 0 to 12 the following equation applies.

$$K = 0.36\ (1 + r)^{0.25} \tag{17}$$

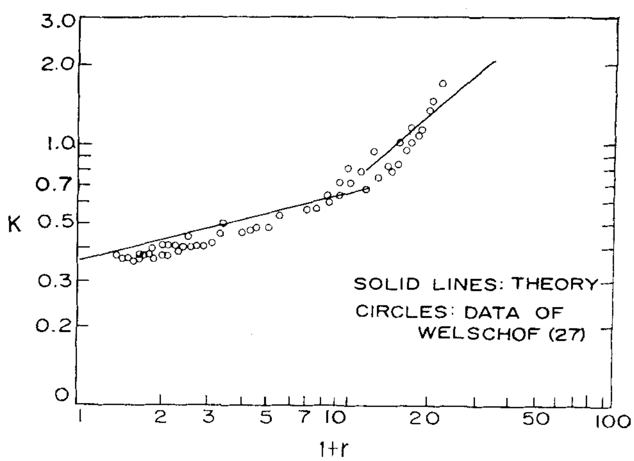

Fig. 5. Comparison of theory with data of Welschof (27).

The data confirm the assumption that the eddy viscosity in a gas-solids stream is a function only of the gas properties and the solids loading. This is demonstrated to be true by the fact that particles with a two hundred fiftyfold range of diameters, a sevenfold range of densities, a twelvefold range of pipe diameters, and a fortyfold range of Reynolds numbers (see Table 1) appear on the plot; yet all distribute about the same straight line. It is also of interest that in the earliest study of pneumatic transport, Gasterstadt (*8*) concluded that pressure drop was a function only of the gas properties and the solids loading. This appears to have been a sound experimental observation on his part. There now exists a theoretical framework that fits the observed relationships and the results can be used with confidence.

Conspicuous by its absence in the correlation is any reference to solid properties such as density or particle size. Although some authors (*3, 11, 20*) have felt the particle diameter or the diameter ratio d/D to be important variables in the pressure drop correlation, the majority (*1, 2, 4, 7, 12, 16, 17*) find no evidence to support this. Attempts to separate the different particle sizes on a plot such as Figure 4 were completely unsuccessful. Plots of individual author's data on graphs of ΔP vs. d likewise showed a random arrangement of particle sizes. Particle density also appeared to have no effect, since points for particles of widely varying densities fell on the same line in Figure 4. This is in agreement with most of the recent work in the field.

Mitlin has thought particle shape to be an important factor in pressure drop correlations. Once again, all the particle shapes found in the data used for this correlation fell on a single line in Figure 4. However, most of them were roughly spherical or cubical, and none had seriously irregular shapes. At least for the range of particles studied here, shape had no effect.

Above an r of 12 an abrupt change takes place in the K vs. $(1 + r)$ dependence as shown in Figure 4. At higher values of $(1 + r)$, the value of K (and therefore of pressure drop) rises much more rapidly as $(1 + r)$ increases. Based on the rather limited data in this high range, the following relationship appears to exist:

$$K = 0.11\,(1 + r)^{0.8} \qquad (18)$$

This break point in the curve may correspond to a sudden change in the mechanism of solids transport. The most likely suggestion is that, as the value of $(1 + r)$ is raised, the mixture ceases to behave as a *pseudogas* and the solids are transported in large agglomerated masses. This kind of dense-phase solids transport has been investigated by Korn (*15*) and Simons (*1, 3*). In this region the solids contribute a proportionately larger share of the total pressure drop.

The data of Welschof (*27*) were discovered after the correlation described above had been developed and the constants evaluated. His data were then used as an independent check on the initial correlation. Values of K calculated from Welschof's data are plotted against $(1 + r)$ in Figure 5. The agreement of these data with the lines representing Equations (17) and (18) is seen to be excellent, the only difference being that Welschof's data appear to show a smooth rather than an abrupt transition to dense-phase flow at $(1 + r) = 13$. Thus the use of this correlation predicts the frictional pressure drop data of Welschof closely.

DESIGN CURVES

With the use of the experimental exponent as determined from Figure 4, Equation (17) was used to determine K and the numerical integration of Equation (16) performed in order to relate Reynolds number and friction factor. The results of these computer calculations are plotted in Figure 6. Calculation of the frictional pressure drop through any pipe for gas-solids loadings up to 12 can be accomplished from these curves in a manner fully analogous to that for single-phase flow.

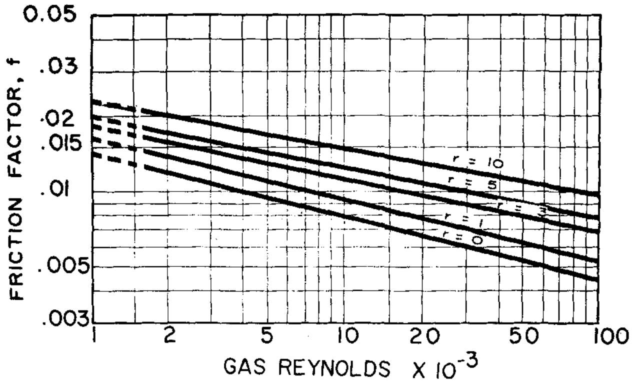

Fig. 6. Friction factors for two-phase gas-solids flow.

NOTATION

D = pipe diameter, ft.
d = particle diameter, ft.
f = friction factor based on gas velocity, defined by Equation (3)
f_s = friction factor based on solids flow
G_g = gas mass velocity, lb./sec./sq. ft.
G_s = solids mass velocity, lb./sec./sq. ft.
g_c = dimensional constant, lb.$_m$ ft./lb.$_f$ sec.^{2}
k = Von Karman constant for single-phase flow
K = Von Karman constant modified for two-phase flow
L = pipe length, ft.
m = power to which $(1 + r)$ is raised in two-phase flow equations
N_{Re} = $D\bar{u}\rho_g/\mu_g$
r = ratio of solids flowing to fluid flowing, lb./lb.
u = local gas velocity in single-phase flow, ft./sec.
$\bar{u}$ = area average gas velocity, ft./sec.
u_g = local gas velocity during gas-solids flow, ft./sec.
u_s = local solids velocity, ft./sec.
u^+ = universal velocity parameter, $= u_g/(\tau_o g_c/\rho)^{1/2}$
y = radial position in the pipe, measured from the wall, ft.
y^+ = universal distance parameter, $= y\rho(\tau_o g_c/\rho)^{1/2}/\mu$

Greek Letters

α = ratio of pressure drop of the two-phase mixture to gas flow pressure drop
ϵ = eddy viscosity
ϵ_{g-s} = eddy viscosity for gas-solids flow
ΔP = total pressure drop between two points, lb./sq. ft./ft.
ΔP_{ag} = pressure drop due to acceleration of the gas, lb./sq. ft./ft.
ΔP_{as} = pressure drop due to acceleration of the solids, lb./sq. ft./ft.
ΔP_f = pressure drop due to friction only, lb./sq. ft./ft.
ΔP_{fg} = pressure drop due to friction of the gas, lb./sq. ft./ft.
ΔP_{fs} = pressure drop due to friction of the solid, lb./sq. ft./ft.
ΔP_T = total pressure drop, lb./sq. ft./ft.
ΔP_x = pressure drop due to static head, lb./sq. ft./ft.
ρ_g = density of the gas, lb./cu. ft.
ρ_s = density of the individual solid particles, lb./cu. ft.

τ = shear stress at a point in the pipe, lb./sq. ft.
τ_o = shear stress at the pipe wall, lb./sq. ft.
μ = gas viscosity, lb./ft.-sec.

LITERATURE CITED

1. Albright, C. W., J. H. Holden, H. P. Simons, and L. D. Schmidt, *Ind. Eng. Chem.*, 43, 1837–1840 (1951).
2. Belden, D. H., and L. S. Kassel, *ibid.*, 41, 1174–1181 (1949).
3. Chin-Yung, Wen, and H. P. Simons, *A.I.Ch.E. J.*, 5, 263 (1959).
4. Clark, R. H., D. E. Charles, J. F. Richardson, and D. M. Newitt, *Trans. Inst. Chem. Eng.*, 30, 209 (1952).
5. Deissler, R. G., *Natl. Advisory Committee Aeronaut. Tech. Note No. 2138* (July, 1950).
6. Depew, C. A., Ph.D. thesis, Univ. California (1960).
7. Farber, L., *Ind. Eng. Chem.*, 41, 1184-1191 (1949).
8. Gasterstadt, H., *Z. Ver Deut. Ing.*, 265, 3-75 (1924).
9. Gill, W. N., and M. Scher, *A.I.Ch.E. J.*, 7, 61-63 (1961).
10. Hariu, O. H., and M. C. Molstad, *Ind. Eng. Chem.*, 41, 1148–1160 (1949).
11. Helander, R. E., Ph.D. thesis, Northwestern Univ., Evanston, Illinois (1956).
12. Hinkle, B. L., Ph.D. thesis, Georgia Inst. Technol., Atlanta, Georgia (1953).
13. Julian, F. M., M.S. thesis, Univ. Houston, Houston, Texas (1964).
14. Kada, H., and T. J. Hanratty, *A.I.Ch.E. J.*, 6, 624 (1960).
15. Koble, R. A., Ph.D. thesis, Univ. West Virginia, Morgantown, West Virginia (1952).
16. Korn, A. H., *Chem. Eng.*, 57, No. 3, 108-111 (1950).
17. Mehta, N. C., Ph.D. thesis, Purdue Univ., Lafayette, Indiana (1955).
18. Mitlin, L., Ph.D. thesis, Univ. London, London, England (1954).
19. Soo, S. L., *Ind. Eng. Chem. Fundamentals*, 1, 33 (1962).
20. ———, *Chem. Eng. Sci.*, 9, 57 (1956).
21. ———, H. K. Ihrig, and A. F. El Kouh, *J. Basic Eng.*, 82, 609 (1960).
22. Soo, S. L., and C. L. Tien, *J. Appl. Mech.*, 27E, 5 (1960).
23. Soo, S. L., G. J. Treyek, R. C. Dimick, and G. F. Hohnstreiter, *Ind. Eng. Chem. Fundamentals*, 3, 98 (1964).
24. Torobin, L. B., and W. H. Gauvin, *Can. J. Chem. Eng.*, Parts I-III, 37 (1959); Parts IV, V, 38 (1960); Part VI, 39 (1961).
25. Vogt, E. G., and R. R. White, *Ind. Eng. Chem.*, 40, 1731–1738 (1948).
26. Von Karman, T., *J. Aeronaut. Sci.*, 1, 1-20 (1934).
27. Welschof, G., *VDI Forschungsheft*, 492, 5 (1962).

Manuscript received November 20, 1964; revision received April 15, 1965; paper accepted April 29, 1965. Paper presented at A.I.Ch.E. San Francisco meeting.

Suction Nucleate Boiling of Water

P. C. WAYNER, JR., and A. S. KESTEN

United Aircraft Corporation Research Laboratories, East Hartford, Connecticut

Suction nucleate boiling (consisting of saturated pool boiling on a porous heat source with the generated vapor exhausting through the pores) was investigated experimentally and theoretically. The experimental results demonstrated that: (1) interfacial free energy can be used to direct the flow of liquid and vapor in a desired direction and to separate vapor from liquid at the point of vapor generation; (2) the heat transfer coefficient for suction nucleate boiling is higher than that associated with normal boiling; and (3) a porous heat exchanger can be designed to give stable transition from nucleate to film boiling. The theoretical analysis, which was based on experimental observations, indicated that extremely high heat fluxes and heat transfer coefficients are possible with small pores. Comparison of the experimental and theoretical results demonstrated that the full potential of suction nucleate boiling was not attained in the experiments and indicated some of the experimental refinements needed to attain this potential.

The feasibility and limitations of many engineering devices depend on normal boiling heat transfer. The results of extensive study in this field have demonstrated that normal boiling is an extremely complex phenomenon with many undesirable characteristics, such as random nucleation and fluid flow; an area of hydrodynamic instability; a heat transfer coefficient dependent on many variables including heat flux, vapor quality, and gravity; a low average heat flux and heat transfer coefficient in the film boiling regime or when a high quality product is desired; and difficulty in obtaining 100% quality vapor because of liquid entrainment. In an effort to eliminate some of these undesirable characteristics, a modified mode of heat transfer consisting of film boiling on a porous heat source with the generated vapor exhausting through the heat source has been proposed (*1*) and studied experimentally (*2*, *3*). The results of these film boiling studies demonstrated the feasibility and advantages of this concept. Since the heat transfer coefficient and heat flux are higher in nucleate boiling than in film boiling, a logical extension of the film boiling work is the study of nucleate boiling on a porous heat source with the generated vapor exhausting through the heat source. This mode of heat transfer, herein designated suction nucleate boiling (S.N.B.), has not been previously studied. The basic difference between suction nucleate boiling and film boiling on a porous heating ele-

P. C. Wayner, Jr., is with the Rensselaer Polytechnic Institute, Troy, New York.