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The Applicability of the Homogeneous Flow Model to Pressure Drop in Straight Pipe and Across Area Changes

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1.0 Introduction

The prediction of two-phase (vapor-liquid) pressure drops in straight-pipes has received considerable attention. It is recognized that in most instances the assumption of a homogeneous flow model does not lead to an accurate estimation of frictional pressure drop. However, under some conditions the homogeneous model does seem to agree with the data. Meyer and Wallis⁽¹⁾ showed that the homogeneous model provided a good representation of the pressure drop data in mist flow. Fitzsimmons⁽²⁾ found that his high pressure, high velocity data were reasonably well predicted by homogeneous flow theory.

In the previous topical report⁽³⁾ issued by this project, the data for pressure drop across abrupt area changes were examined. It was found that the homogeneous model provided a good approximation of the pressure drop across abrupt contractions but that use of a slip model was required for the abrupt expansion data.

The objective of this study was to evaluate systematically the conditions under which the homogeneous model may be applied to prediction of frictional two-phase pressure drop in straight pipes and abrupt area changes. The approach taken was first to examine the available observations on flow pattern transitions. On the basis of the observations in the literature, as well as additional observations made with the Freon-Freon vapor system, a simple criterion for the existence of homogeneous flow was established. The homogeneous model was then compared to the available pressure drop data in the region where it could be considered applicable.

2.0 Flow Pattern Studies

2.1 Flow Patterns

Previous observations of cocurrent vapor-liquid flow in transparent pipe lines have all indicated that the flow occurs in a number of different configurations. At the lowest gas and liquid flow rates, the liquid is observed to flow along the bottom of the tube and the gas above it. At higher flow rates, large gas bubbles, or slugs, are formed. These travel preferentially along the upper section of the pipe. At high gas flows and void fractions, the vapor flows in a central core. At very high mass flows, the large liquid bubbles are dispersed and a homogeneous froth is formed. At high void fractions and mass flows the flow appears as a largely homogeneous mist with only a very thin liquid film on the walls.

These observations are in accord with the flow pattern classification suggested by Choe and Weisman⁽⁴⁾. Their classification is a modification of that proposed by Hubbard and Dukler⁽⁵⁾. Choe and Weisman suggest that the flow patterns observed in cocurrent vapor-liquid flow be classified as:

homogeneous	(includes froth & mist)
intermittent	(includes bubble, slug and plug)
annular	
separated	(includes stratified and wavy)

To determine where a homogeneous model may be applied, a criterion for the transition from homogeneous to non-homogeneous flow (intermittent or annular) is required.

2.2 Experimental Observations

In order to augment the flow pattern observations available in the literature, additional flow pattern observations were made using the Freon-Freon vapor system. Freon 113 was used in the University of Cincinnati Boiling Freon Loop. This loop, described in the previous topical report in this

series⁽³⁾ produces a vapor-liquid mixture by electrical heat addition to a circulating Freon stream.

To allow visual observation, a 4 ft. long glass tube having a nominal diameter of one in. was mounted horizontally in the Boiling Freon Loop. The glass pipe was preceeded by about 4 ft. of steel pipe. However, the flange joining the steel and glass pipes created some mixing action. Flow patterns were observed visually at the end of the glass test section.

At the lowest void fractions ($\alpha \approx 7\%$) and flow rate ($G \approx 1 \times 10^6$ lbs/lb ft²) most of the gas travels in large bubbles or plugs at the top of the tube (see Fig. 1a). As the flow is increased to $G \approx 1.6 \times 10^6$ with α remaining constant, much of the vapor is dispersed throughout the liquid although some of the vapor still travels along the top of the tube (see Fig. 1b).

At a void fraction of about 20% and a mass velocities of about 1×10^6 lbs/hr ft², much of the vapor is travelling over a wavy upper surface of the liquid but a portion of the vapor appears as a froth below the liquid surface (see Fig. 2a). As the flow is increased, the amount of froth increases. Where $G \approx 1.6 \times 10^6$ lbs/hr ft² (see Fig. 2b), essentially all of the vapor is dispersed in a nearly homogeneous froth.

Fig. 3a illustrates the behavior at $\alpha = 42\%$ and $G \approx 1 \times 10^6$ lbs/hr ft². The behavior is similar to that seen in Fig. 2a but the vapor is more widely and evenly distributed in the liquid. As the vapor velocity is increased, the froth is distributed around the circumference of the tube. Fig. 3b shows the appearance at $G \approx 1.6 \times 10^6$ and $\alpha \approx 55\%$. Here the froth is distributed around the circumference of the tube. Direct visual observations indicate there is a small annular gas core but this is not apparent in the photograph. Further increases in velocity lead to a disappearance of the annular core and froth flow throughout the tube.

A series of flow pattern observations were undertaken in the region of the intermittent-homogeneous (or intermittent-froth) transition. A number of these observations were made simultaneously with the recording of the abrupt area change pressure drop data reported in the preceding topical report⁽³⁾. The observations were initially plotted using the flow map coordinates suggested by Baker⁽⁶⁾. These coordinates are G_g/λ and $L\lambda\psi/G_g$ where

G_g = mass velocity of gas phase, $\text{lb}_m/\text{hr ft}^2$

L = mass velocity of liquid phase, $\text{lb}_m/\text{hr ft}^2$

λ = $[(\rho_g/.075)(\rho_L/62.3)]^{1/2}$

ψ = $(73/\sigma)[\mu_1(62.3/\rho_1)^2]^{1/2}$

ρ_g = density of gas phase, lb_m/ft^3

ρ_L = density of liquid phase, lb_m/ft^3

σ = liquid surface tension, dynes/cm

μ_1 = viscosity of liquid, centipoise

The data, however, were displaced somewhat from Baker's proposed transition line derived from air-water data. An improved agreement is obtained if " ψ " is eliminated from the dimensionless group used for the abscissa. The data are shown in this modified Baker plot in Fig. 4.

2.3 Empirical Correlation Describing Homogeneous-Non-Homogeneous Transition

As the first step in development of a general correlation for the homogeneous-non-homogeneous transition, the data of Hoogendorn^(7,8) for various liquid-water systems were plotted using Baker's coordinates. As may be seen in Fig. 5, there is very poor agreement between Baker's proposed transition line and some of the experimental observations in the literature. The Baker plot agrees with the air water system which furnished the original

data on which the map was based. However, it agrees poorly with data for other vapor-liquid systems.

An alternative flow pattern mapping procedure which has been widely used is that suggested by Kosterin⁽⁹⁾. In this mapping procedure, the average mixture velocity, V_m is plotted against the void fraction, α , Fig. 6 shows the experimental data plotted in this fashion. Again, there is very poor agreement between the various systems.

Various other flow pattern mapping schemes were examined. The best results were obtained using a modified version of Baker's map similar to that previously described. The ordinate remains G_g/λ but the abscissa is changed to $G_t \lambda / G_g$ where $G_t = G_g + L$. The results of this approach are seen in Fig. 7. Reasonable agreement between the various systems is now obtained. The data showing the greatest disagreement is that of Alves⁽¹⁰⁾ for the air water system. Other air-water data, however, are in agreement with the remaining systems. The deviation of Alves' data from the remaining systems may be due to the fact that Alves had a trombone-like arrangement of pipes connected by return bends in contrast to the single long pipes of other investigators. It is interesting to note that the line drawn through the mean of the adiabatic data may be extrapolated through the boiling data of Bergles and Suo⁽¹¹⁾.

The correlation of Fig. 7 can be represented by the solid line shown which has the equation

$$G_g/\lambda = 2 \times 10^6 (G_t \lambda / G_g)^{-1} \quad (1)$$

Upon rearrangement, we obtain the result that the homogeneous-non-homogeneous flow transition occurs when

$$G_T = 2 \times 10^6 \text{ lbs/hr ft}^2 \quad (2)$$

Fig. 1a

$$\alpha = .08$$
$$G = 1.1 \times 10^6 \text{ lbs/hr ft}^2$$

Fig. 1b

$$\alpha = .08$$
$$G = 1.6 \times 10^6 \text{ lbs/hr ft}^2$$

Figure 1 - Behavior at Low Void Fractions

Fig. 2a

$$\alpha \approx 0.2$$
$$G = 1.1 \times 10^6 \text{ lbs/hr ft}^2$$

$$\alpha \approx 0.2$$
$$G = 1.6 \times 10^6 \text{ lbs/hr ft}^2$$

Fig. 2 - Behavior at $\alpha = .2$

Fig. 3a

$$\alpha \approx .4$$
$$G = 1.1 \times 10^6 \text{ lbs/hr ft}^2$$

Fig. 3b

$$\alpha \approx .55$$
$$G = 1.1 \times 10^6 \text{ lbs/hr ft}^2$$

Fig. 3 - Behavior at Higher Void Fractions

Fig. 4. Comparison of Present Investigation's Experimental Results with Modified Baker's Plot for Slug-Froth Transition.

Fig. 5. Baker Plot for Intermittent-Homogeneous Transition

Fig. 6. Kosterin Plot for Homogeneous-Non Homogeneous Flow Transition

Fig. 7. Modified Baker Plot for Non-Homogeneous-Homogeneous Transition

One would then expect homogeneous flow whenever $G_T \geq 2 \times 10^6 \text{ lbs/hr ft}^2$.

It should be noted that the data upon which this correlation is based were obtained by observing the behavior at the end of long pipe lines ($L/D > 20$). Entrance effects are not considered. Homogeneous, or nearly homogeneous flow, can be created temporarily by disturbances in the line. Such flow will gradually revert to its original flow pattern.

2.4 Analysis of Transition to Homogeneous Froth

One may consider the creation of a homogeneous froth of bubbles in a liquid in terms which are analogous to the suspension of solid particles. Solid particles are suspended when the turbulent velocity fluctuations provide sufficient lift forces to overcome the tendency of the particles to settle. From theoretical considerations, Weisman and Efferding⁽¹²⁾ showed that, for dilute suspensions, the settling and suspending forces are in equilibrium when

$$P_p = \frac{g V_p (\Delta \rho) u_o}{4g_c} \quad (3)$$

and

P_p = vertical component of power input to single particle, $\text{ft lb}_f/\text{sec}$.

V_p = volume of particle, ft^3

g_c = gravitational conversion factor, $\text{lb}_m \text{ft}/\text{lb}_f \text{sec}^2$

u_o = free settling velocity of particles in infinitely dilute suspension, ft/sec .

$\Delta \rho$ = difference in density between particle & liquid, lb/ft^3

Weisman and Efferding⁽¹²⁾ applied this concept to the suspension of solid particles by mechanical mixers. They reasoned that the power input to an individual particle in a given system would be proportional to the total power input per unit volume, P/V . They further concluded that the

actual power input to the particle would depend on the ratio of tank diameter to particle diameter (D/D_p) the ratio of solid volume to liquid volume $[(1-\epsilon)/\epsilon]$, and the distance above the impeller. They were able to correlate suspension power requirements by this approach.

Weisman⁽¹³⁾ used an extension of this idea to predict the minimum power requirements for suspension of slurries in horizontal conduits. Weisman concluded that the dimensionless group $(\frac{g_c P}{gV(\Delta\rho)u_o})$ should be a function of $(1-\epsilon)/\epsilon$ and (D/D_p) . He showed that the data in the literature could be well correlated by

$$\frac{g_c P}{gV(\Delta\rho)u_o} = 0.4 \left(\frac{1-\epsilon}{\epsilon} \right)^{1/2} \left(\frac{D}{D_p} \right)^{1/2} \quad (4)$$

The power dissipation per unit volume (P/V) in this case is determined from the pressure drop per unit length.

If a homogeneous suspension of small bubbles in a liquid is now considered, it is clear that such a suspension can only be maintained if the turbulent velocity fluctuations are sufficient to overcome the bubble rise velocity. Hence Equation (4) may be applicable if D_p represents the diameter of bubble obtained at given conditions and u_o represents the rise velocity of such a bubble.

The validity of this hypothesis may be examined by rearranging Equation (4). Since the power dissipated per unit volume, P/V , is given by $[(\Delta p/\Delta L)G/\rho]$, the gas velocity at which the required power dissipation is attained may be calculated when $(\Delta p/\Delta L)$ is written in terms of G , ρ and D . We then have

$$G_g = x \left[\frac{0.8 \rho_l D}{\bar{R} f_l} \rho_{av} g (\Delta \rho) u_o \left(\frac{\alpha}{1-\alpha} \right)^{1/2} \left(\frac{D}{D_p} \right)^{1/2} \right]^{1/3} \quad (5)$$

where

G_g = mass velocity of gas phase, lb m/sec ft²

D = pipe diameter, ft

D_p = bubble diameter, ft

g = acceleration of gravity, ft/sec²

$\Delta \rho$ = $(\rho_f - \rho_g)$ lbs/ft³

α = void fraction

$\bar{R}$ = two phase friction factor multiplier

f_l = friction factor (based on liquid flowing at same total G)

x = flow quality,

ρ_l = liquid density, lbs /ft³

ρ_{av} = average density of mixture

Use of Equation (5) requires an estimation of the gas bubble size in a homogeneous froth. To the authors knowledge, no predictive equation for the bubble size produced under these circumstances is available. However, Hinze⁽¹⁴⁾ has considered the general process of the dispersion on one phase within another by turbulent velocity fluctuations. For the dispersion of one liquid within another, Hinze concluded that the maximum diameter, d , of the dispersed phase particles could be predicted from

$$d = .725 \left(\frac{\sigma}{\rho_f} \right)^{3/5} \left(\frac{P}{M} \right)^{-2/5} (g_c)^{1/5} \quad (6)$$

where

d = maximum size of dispersed particle, mm

σ = surface tension (lbs_f/ft)

ρ_f = density of continuous phase (liquid) lbm m/ft³

P = power dissipated, ft lb_f /sec

M = mass in which power dissipated, lbs_m

g_c = gravitational conversion factor;

$$32.17 \text{ (lbm/lbf)(ft/sec}^2\text{)}$$

In order to ascertain whether the particle diameter of Equation (6) predicted bubble diameter in homogeneous froth flow, a series of photographic observations were made in the Freon 113 system. Homogeneous flow was established and high speed photographs were taken at these mass velocities. From enlargements of these photographs, the bubble size distribution was determined. This was accomplished by choosing several different areas at random. The diameters of all bubbles within each region were measured and the number of bubbles within a series of size ranges was determined. This was repeated for each of the randomly selected regions. The mean bubble diameter was then determined from the classified data.

Fig. 8 compares the bubble sizes measured with those predicted by Equation (6). The vertical lines indicate the range of bubble sizes observed. The bubble sizes observed were the same order of magnitude as predicted by Hinze but were significantly smaller. The maximum bubble sizes were 60 to 90% of the calculated d_{95} values and the mean values observed (d_{50}) were approximately 45% of the predicted d_{95} values. (See Appendix for data tabulation.)

Based upon these bubble size observations, it was assumed that the form of Equation (6) was correct but that the predictions overestimated the average bubble size. The bubble size was therefore taken as 45% of that given by Equation (6).

The bubble sizes observed are in size range where the equation of Harmathy⁽¹⁵⁾ may be used for calculation of rise velocity. The terminal bubble rise velocity is then obtained from

$$u_o = 1.53 \left(\frac{g \Delta \rho \sigma}{\rho_f^2} \right)^{1/4} \quad (7)$$

where

u_o = terminal bubble rise velocity, cm/sec

g = gravitational acceleration, 981 cm/sec²

$\Delta \rho = (\rho_f - \rho_g)$

ρ_f, ρ_g = liquid and gas densities, gm/cc

σ = surface tension, dynes/cm

By use of Equation (5) in conjunction with the bubble rise velocity of Equation (7) and a bubble diameter based on a modified Equation (6), the predicted and observed gas mass flow rates for the homogeneous-non-homogeneous transition may be compared. This is done in Fig. 9. The agreement is quite reasonable.

The relationship between Equation (5) and the empirical transition line of Equation (1) may be elucidated by substitution of the expressions for bubble size and rise velocity into Equation (5). When this is done we obtain

$$G/\eta = \left[\frac{(1-\alpha)^{.1} \alpha^{.5}}{\bar{R}^{.8}} \right]^{1/2.4} \quad (8)$$

Fig. 8 Comparison of Bubble Size Predicted by Hinze's Equation with the Measured Bubble Size Using Freon-Freon Flow.

Fig. 9 Minimum Power Model Transition line compared with Hoogendoorn's Experimental Results for Froth Flow Transition

Fig.10: Comparison of Minimum Power Model Predictions with Experimental Homogeneous Transition Data.

where

$$\eta = \left[\frac{1.88 \rho_1^{2.45}}{f_1^{.8}} \cdot \frac{g^{1.25} D^{1.3}}{\sigma^{.05} g_c^{.05}} \right]^{1/2.4}$$

G = total mass flow rate

When G/η is plotted vs α (Fig. 10), it is seen that G/η is almost invariant with α over a wide range of α . Further, since η varies little with system properties for normal fluids, the transition will appear to take place at a nearly constant G providing line size variations are moderate. This is in accord with the empirical correlation of Equation (1).

It may be noted that Equation (8) predicts a slight variation of G with pipe diameter although Equation (1) does not. This may be an indication that the effect of the (D/D_p) ratio is not the same for solid particles and bubble suspensions. Alternatively, it may indicate that the Hinze prediction of the size of dispersed particles requires modification when applied to gas bubbles. Presently available data is insufficient for resolution of this question.

3.0 Relationship Between Homogeneous Flow and Frictional Pressure Drop in Straight Pipes

3.1 Comparison With Specific Data Sets

It was previously observed that several investigators claim that, under appropriate conditions, two-phase frictional pressure drop can be reasonably predicted by using a homogeneous flow model. Owens⁽¹⁶⁾ suggested this approach for steam-water systems at high pressures. Meyer and Wallis⁽¹⁾ showed that the homogeneous flow predictions were in reasonable agreement with the pressure drop data for annular-mist flow. Wallis⁽¹⁷⁾ pointed out that in steam-water systems, there is a

drift from Martinelli-Nelson predictions towards homogeneous theory as mass velocity is increased.

If homogeneous-non-homogeneous flow pattern transition occurs at $G_T \approx 2 \times 10^6$ lbs/hr ft² then, when $G_T \geq 2 \times 10^6$, two phase frictional Δp should then be predicted by homogeneous flow theory. This hypothesis was tested against several sets of data. In doing so, it was assumed that the average fluid density could be calculated from the quality assuming a slip ratio of one. Further, in obtaining friction factors, it was assumed that the average fluid viscosity, μ , could be obtained, as suggested by McAdams⁽¹⁸⁾, from

$$\frac{1}{\mu} = \frac{x}{\mu_g} + \frac{1-x}{\mu_l} \quad (9)$$

where x = quality

Figs. 11 and 12 compare the data of Fitzsimmons⁽²⁾ for steam water flows in horizontal tubes to the frictional pressure drop predicted by homogeneous flow theory. It is seen that the data at $G \geq 2 \times 10^6$ lbs/hr ft² agrees well with the homogeneous flow predictions.

At high velocities, one may expect that orientation of the test section would not affect the flow pattern transition. Hence, one would expect that, although obtained from horizontal flow data, the $G_T \geq 2 \times 10^6$ lbs/hr ft² criterion would also be valid for vertical flow. This would appear to be borne out by the frictional pressure drop data of Petrick⁽¹⁹⁾ for the air-water system. Fig. 13 compares these vertical flow data to the homogeneous flow prediction. Again it is seen that, so long as $G_T \geq 2 \times 10^6$ lbs/hr ft², the homogeneous flow theory predicts the data.

In Figs. 14 through 19, the extensive steam-water frictional ΔP data for vertical flow which was compiled by Muscettola⁽²⁰⁾ are compared with homogeneous flow predictions. For each system examined, the homogeneous predictions agree, within the scatter of the data, providing G_T is in excess of 2×10^6 lbs/hr ft². The agreement of these data would appear to support the use of Equation (1) as the criterion for the homogeneous to non-homogeneous flow transition.

3.2 Statistical Analysis of World Data On Pressure Drop Through Horizontal Pipes

Although the previously cited straight pipe pressure drop data, support the use of the homogeneous model at mass velocities above 2×10^6 lbs/hr ft², it is not entirely clear that there is no further effect of mass velocity. To further examine this question, the data bank of world pipe line pressure drop data compiled by Mandhane et al⁽²¹⁾ has been secured. The steam-water data of Fitzsimmons⁽²⁾, air-water data of Petrick⁽¹⁹⁾ and very high velocity steam-water data of Miropolski et al⁽²²⁾ were added to the data bank.

All data in the revised data bank obtained at mass velocities equal to or greater than 2×10^6 lbs/hr ft² were compared with predictions obtained assuming homogeneous flow. In arriving at this comparison, it was assumed that the average fluid density could be calculated from the quality assuming a slip ratio of one. To obtain the mixture viscosity, both the approach of McAdams (see Equation 9) and Dukler et al. were examined. Dukler et al.⁽²⁴⁾ suggested

$$\mu = \frac{V_{sg}}{V_m} \mu_g + \frac{V_{sl}}{V_m} \mu_l \quad (10)$$

Fig. 11

Ratio of Two-Phase to Single Phase Pressure Drop
for Horizontal Flow of Steam-Water Mixtures in Straight Pipes as
Reported by Fitzsimmons (2).

Fig. 12

Fig. 13 Ratio of Two-Phase to Single-Phase Pressure Drop for Air-Water Flow in Vertical Straight Channels. (Petrick)

Fig. 14 Comparison of Steam Water Data (Vertical Flow) Compiled by Muscatella with Homogeneous Flow Predictions.

Ratio of Two-Phase to Single Phase Pressure drop for
Steam-Water Mixtures in Vertical Straight Pipes

Fig. 15. Comparison of Steam Water Data (Vertical Flow) Compiled by Musettola with Homogeneous Flow Predictions,

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Ratio of Two-Phase to Single Phase Pressure drop for
Steam Water Flow in Vertical Straight Pipes

Fig. 16. Comparison of Steam Water Data (Vertical Flow) Compiled by Muscettola with Homogeneous Flow Predictions.

Ratio of Two Phase to Single Phase Pressure drop for Steam-Water Flow in Vertical Straight pipes

Fig. 17: Comparison of Steam Water Data (Vertical Flow) Compiled by Muscetola with Homogeneous Flow Predictions.

Ratio of Two-Phase to Single Phase Pressure drop for
Steam-Water Flow in Vertical Straight Pipes. (Muscetola)

Fig. 18. Comparison of Steam Water Data (Vertical Flow) Compiled by Muscettola with Homogeneous Flow Predictions.

Ratio of Two-Phase to Single Phase Pressure Drop for Steam-Water Flow in Vertical Straight Pipes. (Muscettola)

Fig. 19. Comparison of Steam Water Data (Vertical Flow) Compiled by Musettola with Homogeneous Flow Predictions.

Ratio of Two-Phase to Single Phase Pressure drop for Steam Water Flow in Vertical Pipes

where V_{sl} , V_{sg} = superficial velocities as of liquid and gas respectively

$$V_m = V_{sl} + V_{sg}$$

and other symbols have their previous meaning. Both approaches give nearly the same result.

All of the available data for horizontal tubes, with the exception of the results of Miropolski et al.⁽²²⁾, are compared to the predictions in Fig. 20. In this comparison Equation (9) was used for computation of mixture viscosity. It may be seen that relatively good agreement is attained between measurements and predictions although there is a tendency for the measurements to be slightly higher than predictions.

The comparison may be quantified by a statistical analysis of the data. Following Dukler et al.⁽²³⁾ we define a fractional deviation d_i such that

$$d_i = (P_i - M_i)/M_i \quad (11)$$

where P_i = predicted value for i^{th} data point

M_i = measured value for i^{th} data point

The average fraction deviation, $\bar{d}$, is then

$$\bar{d} = \left(\sum_{i=1}^n d_i \right) / n \quad (12)$$

An estimate, $s(d)$, of the standard deviation of d is then given by

$$s(d) = \sqrt{\frac{\sum d_i^2}{n} - \bar{d}^2} \quad (13)$$

where n = no. of data points.

The evaluation of the 679 data points give the following results .

Table 1

	Viscosity from Eq (9)	Viscosity from Eq (10)
$\bar{d}$	-.163	-.197
$s(d)$	.403	.358

No particular trend is observed for differing systems. If the data for horizontal flow is divided into three groups, we find (using μ from Eq. 9)

Table 2

	Steam-Water	Air-Water	Non-Water Systems
No. Pts.	94	187	398
$\bar{d}$	-.2793	-.192	-.122
$s(d)$	.356	.55	.317

No clear trend is observed with mass velocity in the range of $2.0 \times 10^6 \leq G \leq 5.5 \times 10^6$ lbs/hr ft². Table 3 shows the analyses obtained when the data is categorized according to mass velocity.

Table 3

	$2 \times 10^6 \leq G \leq 3 \times 10^6$	$3 \times 10^6 \leq G < 4 \times 10^6$	$4 \times 10^6 \leq G < 5 \times 10^6$	$5 \times 10^6 \leq G < 5.5 \times 10^6$
No. Pts.	410	191	66	12
$\bar{d}$	-.151	-.149	-.285	-.13
$s(d)$	0.47	.29	.19	.05

Fig. 20. Comparison of Measured Straight Pipe Pressure Drop In Horizontal Flow With Homogeneous Model (Viscosity From Eq. 9)

The 225 high G data points for vertical lines are very well correlated by the homogeneous flow model. For runs with $G \leq 2 \times 10^6$ lbs/hr ft², it is found that $\bar{d} = -.0435$ and $s(d) = 0.2018$. Nearly all the points are at high void fractions; less than 10 points have α values below 0.5.

Although the flow pattern transition criterion was based on horizontal flow data, the criterion would seem to apply to vertical flow as well. This would indicate that gravitational effects are not of major importance at the high mass flow rates involved.

It appears that under the conditions where homogeneous flow is predicted, homogeneous theory does a reasonable job of predicting straight pipe frictional pressure drop. The small underprediction of the horizontal pressure drop data by the homogeneous model used may indicate that the viscosities predicted by Equation (9) or (10) are too low. One would expect that the viscosity of a frothy mixture would be higher than the viscosity of the liquid. This effect would be missing from the vertical data examined since this data is all at high void fractions where mist flow occurs.

One way in which the viscosity of a froth may be estimated is to assume that the vapor bubbles may be considered as small, rigid spheres. If this assumption is correct, the viscosity of the froth is a function of the void fraction. The correlation of Happel^(22a) for viscosity of suspensions of solid sphere may be used for this prediction with α replacing the volume fraction of solutions. We then have for, μ_{TP} , the viscosity of the foam

$$\mu_{TP} = \mu_l \exp \left(2.5 \alpha / (1 - 39\alpha/64) \right) \quad (13a)$$

When the viscosity computed by Equation (13a) was used for evaluation of the Reynolds number for all data points where $\alpha < 0.5$, the pressure drop underprediction of the homogeneous model was eliminated.

At high void fractions, the homogeneous mixture moves from a froth to a mist. As α approaches 1.0, the viscosity of the mixture must approach that of the vapor. It was therefore arbitrarily assumed that the peak viscosity occurred at $\alpha=0.5$ and that, at higher α , the viscosity followed a hyperbolic curve which approached μ_g as α approached 1.0. This behavior can be modeled by assuming that for $\alpha \geq 0.5$

$$\mu_{TP} = \mu_g + (\mu_c - \mu_g) \left(\frac{1}{\alpha} - 1 \right)^k \quad (13b)$$

where μ_g = viscosity of vapor

μ_c = μ_{TP} evaluated by Equation (13a) with $\alpha = 0.5$

k = constant

The value of k was varied and the value selected was that which led to the minimum fractional deviation for all of the available data.

When Equation (13a) and (13b) were used to evaluate μ_1 with $k = 3$,

it was found that the homogeneous model more nearly predicts the mean of the data; viz; $d = .035$, $s(d) = .49$.

When all of the data (horizontal and vertical flow) are compared to this revised model it is found that

$$\bar{d} = .0125$$

$$s(d) = .44$$

The only data known to the authors for mass velocities in excess of 5.5×10^6 lbs/hr ft² are those reported by Miropolski et al.⁽²²⁾ These data indicate that at mass velocities greater than about 10^7 lbs/hr ft² and qualities greater than about 0.2, the two phase frictional pressure drop falls substantially below that given by the homogeneous model. At a G of $2 \cdot 10^7$, reported pressure drops are as low as 20% of that of the homogeneous model. Chisholm^{(22b)(22c)} has developed a correlation which fits this very high velocity data. However, Chisholm's correlation overpredicts the pressure drop for mass velocities between 2×10^6 and 5.5×10^6 lbs/hr ft². Comparison of Chisholm's correlation with the available data yields

$$\bar{d} = .2663$$

$$s(d) = .746$$

for this velocity range.

At the qualities and velocities at which low frictional pressure drop is observed, mist flow would be expected. It is possible that an inverse annular flow (steam on outside, mist in center core) could develop under these conditions. Confirmation of these results by other investigators is needed before any firm conclusions can be drawn about this very high velocity range.

Fig. 21. Comparison of Measured Straight Pipe Pressure Drop With Homogeneous Model (Viscosity from Equations (13a) and (13b))

4.0 Relationship Between Flow Pattern and Pressure Drop Across Area Changes

The pressure differences across abrupt area changes have generally been considered without any reference to the flow pattern existing in the system. One would expect that there should be some relationship between flow pattern and pressure drop. Of particular interest, is the relationship between the conditions under which homogeneous flow is observed and those under which homogeneous theory predicts the observed pressure drop. If the observation that homogeneous flow occurs in straight pipes when $G \geq 2 \times 10^6$ lbs/hr ft² is correct, then one would expect that, as the mass velocity increased, the area change pressure drop should approach that predicted by homogeneous theory.

4.1 Abrupt Expansions

It was previously determined⁽²⁾ that the pressure change Δp_ϵ abrupt expansion may be obtained from

$$\Delta p_\epsilon = \frac{G_T^2 \sigma^2}{\rho_l g_c} \left[\frac{\rho_l}{\rho_g} x^2 \left(\frac{1}{\alpha_1 \sigma} - \frac{1}{\alpha_2} \right) + [(1-x)^2 \left(\frac{1}{\sigma(1-\alpha_1)} - \frac{1}{(1-\alpha_2)} \right)] \right] \quad (14)$$

where

Δp_ϵ = pressure rise between locations 1 and 2 assuming pipe friction is negligible

ρ_g, ρ_l = gas and liquid densities, respectively

G_T = total mass velocity based on area of smaller pipe (lbs/hr ft²)

x = quality (lbs vapor/lb total fluid flowing)

σ = area ratio

α_1, α_2 = void fraction at locations upstream and downstream of expansion, respectively.

Based on analysis of the data of Fitzsimmons⁽²⁾, Ferrell and McGee⁽²³⁾, Mendler⁽²⁴⁾ and Janssen⁽²⁵⁾ it was concluded that the best correlation was obtained when α_1 and α_2 were computed assuming slip flow. The Hughmark⁽²⁶⁾ correlation was used to relate α and quality. The overall result was substantially better than when a homogeneous model was assumed.

Table 5 shows the results of a statistical comparison of the abrupt expansion data when several velocity ranges are considered.

Table 5

Statistical Comparison of Abrupt Expansion Data

	<u>Homogeneous Model</u>		<u>Slip Flow</u>	
	<u>$\bar{d}$</u>	<u>$s(d)$</u>	<u>$\bar{d}$</u>	<u>$s(d)$</u>
$G^* < .5 \times 10^6$	0.60	.94	-0.02	.64
$G \leq 1 \times 10^6$	0.49	.82	-0.03	.54
$G > 1 \times 10^6$	0.05	.11	-0.08	.09
$G \geq 2 \times 10^6$	0.10	.06	-0.00	.08
all G	0.42	.77	-0.04	.49

* G based on mass velocity in largest pipe in above table.
The quantities $\bar{d}$ and $s(d)$ are defined by Equations (12) and (13).

It will be observed that as the mass velocity is increased the average fractional deviation shown by the homogeneous model decreased. For $G > 1 \times 10^6$, the magnitude of $\bar{d}$ for the homogeneous model is less than that for the slip flow model. Similarly, the value of $s(d)$ continuously decreases with increasing mass velocity. Although the value of $\bar{d}$ appears to increase slightly for $G \geq 2 \times 10^6$, the change is probably more apparent than real.

Considering the small number of points in the mass velocity range above 2×10^6 and the value of $s(d)$, it may be shown that the difference seen is within the variability to be expected if the true universe means were the same in each velocity range.

It may be concluded that the behavior of the data on pressure changes across abrupt expansions is in general agreement with visual observations on the onset of homogeneous flow. The abrupt expansion data would appear to indicate that when $G \geq 1. \times 10^6$ lbs/hr ft² (in large pipe) the flow is sufficiently homogeneous for a homogeneous approximation to be used in calculations. However, the data of this investigation for pressure drop across expansion-contraction combination indicate that the homogeneous model still overpredicts ΔP_E at $G=1. \times 10^6$ lbs/hr ft² in the largest size pipe.

4.2 Abrupt Contractions

Based on a one dimensional momentum balance for conditions where x is nearly constant, Janssen concluded that the contraction pressure change, ΔP_c , is given by

$$\Delta P_c = \frac{G_1^2}{2g_c \rho_1} \left[\frac{1}{\sigma^2} \left\{ \frac{\rho_1}{\rho_g} x^2 \bar{\alpha}_1 \left(\frac{1}{C^2 \alpha_3^2} - \frac{1}{\alpha_4^2} \right) + (1-x)^2 (1-\bar{\alpha}_1) \right. \right. \\ \left. \left(\frac{1}{C^2 (1-\alpha_3)^2} - \frac{1}{(1-\alpha_4)^2} \right) \right\} - \frac{2}{\sigma^2} \left\{ \frac{\rho_1}{\rho_g} x^2 \left(\frac{1}{C \alpha_3} - \frac{1}{\alpha_4} \right) + (1-x)^2 \left(\frac{1}{C (1-\alpha_3)} - \frac{1}{(1-\alpha_4)} \right) \right\} \right. \\ \left. + \frac{\rho_1}{\rho_g} x^2 \bar{\alpha}_2 \left(\frac{1}{\sigma^2 \alpha_4^2} - \frac{1}{\alpha_1^2} \right) + (1-x)^2 (1-\bar{\alpha}_2) \left\{ \frac{1}{\sigma^2 (1-\alpha_4)^2} - \frac{1}{(1-\alpha_1)^2} \right\} \right] \quad (18)$$

where

G_1 = total mass velocity at section 1 lbm/ft²hr

C = vena contracta area ratio

σ = channel area ratio

x = mixture quality

$\alpha_1, \alpha_3, \alpha_4$ = void fractions at sections 1, 3 and 4

$$\bar{\alpha}_1 = (\alpha_3 + \alpha_2)/2$$

$$\bar{\alpha}_2 = (\alpha_1 + \alpha_4)/2$$

Subscripts

1 = upstream 2 = at abrupt area change

3 = at vena contracta 4 = downstream

The previously reported analyses by Husain and Weisman⁽³⁾ of their own Freon data and the steam water data of references (2) (23) (25) (27), concluded that the data were best predicted assuming slip flow (α and x related by Hughmark's⁽²⁶⁾ correlation) and mixing at the vena contracta. At the vena contracta the void fraction, α_3 , was obtained from

$$\alpha_3 = \alpha_{\text{homogeneous}} \quad \text{for } \alpha_4 \leq 0.5$$

$$\alpha_3 = \alpha_{\text{slip}} + A(\alpha_{\text{homogeneous}} - \alpha_{\text{slip}}) \quad \text{for } \alpha_4 \geq 0.5$$

where

$$A = 1.5 - \alpha_4$$

α_{slip} = value of α which would be computed for small size pipe in absence of contraction

$\alpha_{\text{homogeneous}}$ = void fraction based on a slip ratio of 1.0

Husain and Weisman also found that a homogeneous model (all α constant at values obtained assuming a slip ratio of one) gave nearly as good a correlation of the data even though it was not in accord with the experimental facts.

Although the homogeneous model provided a good means for approximation of the pressure drop across a contraction, it clearly is not in accord with the actual physical process over most of the data range. One might expect that the homogeneous model would furnish a better approximation at the higher mass velocities. Table 4 compares the homogeneous model to Husain and Weisman's model D for a set of velocity ranges at the upper end of those examined. The performance of the Husain and Weisman model deteriorates slightly at these high mass velocities while that of the homogeneous model remains good. The homogeneous model would appear to be appreciably better than the Husain and Weisman model for $G > 1.5 \times 10^6$ lbs/hr ft². There is, however, only limited data to back this conclusion since only 10 of the 497 data points had mass velocities (in large size pipe) exceeding 1.5×10^6 lbs/hr ft².

Table 4

Statistical Analysis of Abrupt Contraction Data *

	Husain and Weisman Model		Homogeneous Model	
	$\bar{d}$	s(d)	$\bar{d}$	s(d)
$G > .5 \times 10^6$ lbs/hr ft ²	-.092	.275	-.075	.30
$G > 1 \times 10^6$ lbs/hr ft ²	-.166	.117	-.156	.094
$G > 1.5 \times 10^6$ " " "	-.169	.022	-.054	.010
$G > 2.0 \times 10^6$ " " "	-.164	.00	-.046	.00
all G	.0034	.253	.028	.269

5.0 Conclusions

Correlation of visual observations of two-phase flow in straight pipes shows that homogeneous flow occurs when the total mass velocity exceeds 2×10^6 lbs/hr ft². For the froth flow region, this observation is shown

* Note: G based on area of largest size pipe.

to be essentially consistent with a general approach to the suspension of small bubbles and particles. The analysis indicates that fluid physical properties and void fraction affect the mass velocity at which homogeneous flow begins only very slightly.

The onset of homogeneous flow as determined by visual correlations is in accord with frictional pressure drop behavior in straight pipes. For air-water and steam-water systems over a wide range of pressures, it is found that, at mass velocities in excess of 2×10^6 lbs/hr, ft², a homogeneous model can reasonably predict the data. Caution should be used in using the homogeneous model for high void fractions at $G \geq 10^7$ lbs/hr ft², Russian correlations indicate the possibility of inverse annular flow under these conditions.

It may be concluded that the data on pressure drop across abrupt area changes are reasonably consistent with the predicted transition to homogeneous flow in straight pipes. Although contraction pressure drop may be approximated by the homogeneous model over a wide velocity range, the model should only be used for expansion pressure drops at very high mass flow rates. Any further studies of the inter-relationship of flow pattern and area change pressure drop will require void fraction - quality relationships which consider flow patterns.

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Bubble Size Distribution Data

Bubble size distributions were determined from analyses of enlarged high speed photographs. The photographs were taken at low void fractions where individual bubbles could be distinguished.

Pressure = 40 psia

Photo No. 6: Mass Flow rate = 1.96×10^6 lbm/hr ft², $\alpha \approx .04$

Photo No. 14: Mass flow rate = 1.79×10^6 lbm/hr ft², $\alpha \approx .02$

Average bubble dia.	No. of bubbles Photo #6	No. of bubbles Photo #14	Av. bubble dia	No. of bubbles ^s in Photo #6	No. of bubbles in Photo #14
mm.			mm.		
.08	13	0	1.11	1	0
.109	12	1	1.20	1	0
.138	21	5	1.23	1	0
.168	16	20	1.32	1	0
.197	30	38	1.34	1	0
.227	17	24			
.256	8	14		Total 172	146
.285	16	11			
.315	4	11		d = .318 ± .27mm	$\bar{d} = .288 \pm .169\text{mm}$
.344	0	2			
.374	0	2			
.403	2	1			
.432	4	0			
.462	1	1			
.491	2	1			
.522	1	2			
.550	0	0			
.579	0	0			
.610	1	3			
.638	1	0			
.669	2	3			
.698	1	0			
.757	2	0			
.786	1	1			
.800	6	0			
.874	2	2			
.933	1	1			
.963	1	0			
1.021	2	0			
1.08	0	2			

Photo No. 10. Mass Flow Rate = 1.6×10^6 lbm/hr ft²

Pressure = 40 psia

$\alpha \approx .03$

Average dia	No. of bubbles
----------------	----------------

mm.

.1842	2
.216	7
.249	19
.281	28
.312	49
.346	10
.377	15
.410	7
.442	4
.474	2
.507	1
.895	8

Total	152
-------	-----

$\bar{d} = .3456 \pm .142$

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