Two-phase Pressure Drops in Large-diameter Pipes

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In gas-liquid two-phase flow one may imagine four possible *flow types*: viscous-turbulent, turbulent-viscous, viscous-viscous, and turbulent-turbulent, each being defined by the type of flow in liquid and gas phases respectively. Flow in either phase with Reynolds numbers greater than 2,000 is arbitrarily considered turbulent.

In addition to characterizing flow by the type, it is common to identity it also by some graphic description which defines a *flow pattern*. Flow patterns appear to vary both with mass velocity and with the mass velocity ratios of gas-to-liquid (*1, 2*). Approximately in the order of occurrence, as the ratio of gas-to-liquid flow increases, seven specific flow patterns are observed (*1*):

1. *Bubble flow*: Flow in which bubbles of gas move along the upper part of the pipe at approximately the same velocity as the liquid.
2. *Plug flow*: Flow in which alternate plugs of liquid and gas move along the upper part of the pipe.
3. *Stratified flow*: Flow in which the liquid flows along the bottom of the pipe and the gas flows above, over a smooth liquid-gas interface.
4. *Wavy flow*: Flow which is similar to stratified flow except that the gas moves at a higher velocity and the interface is disturbed by waves traveling in the direction of flow.
5. *Slug flow*: Flow in which a wave is picked up periodically by the more rapidly moving gas to form a frothy slug which passes through a pipe at a much greater velocity than the average liquid velocity.
6. *Annular flow*: Flow in which the liquid forms in a film around the inside wall of the pipe and the gas flows at a high velocity in a central core.
7. *Spray flow*: Flow in which most or nearly all the liquid is entrained as spray by the gas.

Isben, Moen, and Mosher (*4*) have discussed the complex problems of predicting or correlating two-phase pressure drops in pipes. Engineering estimates using correlating methods such as those proposed by Chenoweth and Martin (*3*) and by Lockhart and Martinelli (*6*) are usually satisfactory, but may be improved if additional experimental parameters are included (*2, 4, 5*).

CHENOWETH AND MARTIN CORRELATION

The general correlation proposed by Chenoweth and Martin, shown in Figure 1, is limited to the turbulent-turbulent type of flow and does not specifically include the concept of flow pattern.

The abscissa denotes the superficial liquid-volume fraction and the ordinate represents the ratio of the two-phase pressure drop to a fictitious all-liquid pressure drop, $\Delta P_{TP}/\Delta P_L^*$. The calculation of ΔP_L^* is described below. The family of lines on the plot refers to different values of the ratio of the fictitious all-gas to all-liquid pressure drops, $\Delta P_G^*/\Delta P_L^*$. Each fictitious single-phase pressure drop is calculated from single-phase friction-factor correlations and is based on the *total mass* flow rate of both liquid and gas. Calculating ΔP_L^* requires use of the physical properties of the liquid phase and ΔP_G^*, the physical properties of the gas phase.

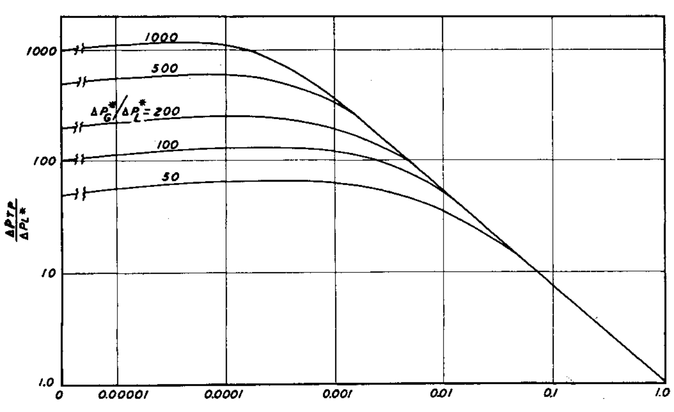

Fig. 1. Chenoweth and Martin correlation.

LOCKHART AND MARTINELLI CORRELATION

In the method proposed by Lockhart and Martinelli (*6*), two-phase pressure-drop data are correlated by plotting log ϕ_G as a function of the parameter, log X. Four different plots are given by Lockhart and Martinelli to cover the four possible flow types. Flow patterns are not specifically considered.

DISCUSSION OF CORRELATIONS

The correlations developed by Lockhart and Martinelli were based on experimental two-phase pressure-drop data taken in small-diameter pipes at pressures up to 50 lb./sq. in. abs. Chenoweth and Martin correlated these data as well as data of their own taken in pipes to 3 in. and at pressures up to 100 lb./sq. in. abs. The low-pressure Chenoweth and Martin data, even in the 3-in. pipe, agreed reasonably well with the Lockhart-Martinelli correlation, but the high-pressure data showed definite deviations, in some cases by as much as 250%. The Chenoweth and Martin correlation, while empirical and applicable only when the flow type is turbulent-turbulent, is reported to apply to most of the applicable experimental two-phase pressure-drop data within ±20 to 30%; the data tested resulted from experiments in pipes from 0.5- to 3.5-in. diam., with pressures varying from atmospheric to 100 lb./sq. in abs., and with fluid-gas systems involving air and water, kerosene, benzene, and diesel oil.

Pressure-drop data for pipe sizes above 3.5 in. are very scarce. Baker (*2*) has presented a few results for gas-oil flow in 4- to 10-in. pipes at pressures around 1,000 lb./sq. in. abs. To extend Baker's results and to determine the validity of the Chenoweth and Martin and Lockhart-Martinelli correlations for low-pressure air-water flow in large-diameter pipes were the purposes of this investigation.

TABLE 1. SUMMARY OF DATA AND CALCULATIONS

Flow rates		Pressure drops			P_{system},
Air, std. cu. ft./min.	Water, gal./min.	ΔP_{TP}, in. H_2O	ΔP_L, in. H_2O	ΔP_G†, in. H_2O	lb./sq. in. abs. (avg.)
4-in. pipe					
100	98	21.26	4.68	0.230	21.3
200	100	37.25	4.86	0.780	22.0
300	106	57.14	5.42	1.60	22.2
400	106	71.57	5.42	2.76	21.4
300	103	51.68	5.17	1.63	21.8
100	155	28.86	10.53	0.165	30.2
200	155	45.24	10.53	0.575	29.9
300	155	61.43	10.53	1.20	29.6
400	155	79.95	10.53	2.00	29.8
300	155	67.28	10.53	1.21	29.5
100	198	42.51	15.11	0.200	24.9
150	189	52.26	16.38	0.344	29.8
200	198	87.75	16.38	0.714	24.1
300	198	105.7	16.38	1.47	24.3
400	185	117	14.43	1.94	30.6
6-in. pipe					
100	192	5.85	1.86	0.043	27.1
200	192	8.40	1.86	0.17	27.6
300	192	10.08	1.86	0.36	27.0
400	192	14.62	1.86	0.645	26.9
485	192	15.25	1.86	0.900	27.4
400	192	14.43	1.86	0.630	27.4
100	415	17.48	8.50	0.042	27.9
200	415	23.90	8.50	0.17	27.8
300	432	29.80	8.98	0.35	27.7
400	415	27.90	8.50	0.628	27.7
485	415	33.20	8.50	0.894	27.6
100	398	16.20	8.80	0.046	28.9
200	399	19.80	8.80	0.16	28.8
300	399	24.90	8.80	0.34	28.3
400	398	30.90	8.80	0.595	29.1
455	397	31.60	8.80	0.744	29.1
200	396	22.50	8.80	0.17	28.3
400	396	31.20	8.80	0.627	28.1
100	498	24.0	12.4	0.041	28.8
100	500	23.4	12.4	0.042	28.3
150	500	24.5	12.4	0.094	28.3
200	500	33.2	12.4	0.17	28.2
300	502	41.7	12.4	0.37	26.5
300	500	34.4	12.4	0.33	29.1
400	502	47.2	12.4	0.66	26.4
400	500	47.4	12.4	0.66	26.3
200	495	29.6	12.3	0.17	28.1
400	500	37.4	12.4	0.59	29.6

†Corrected to system pressure by Fanning equation.

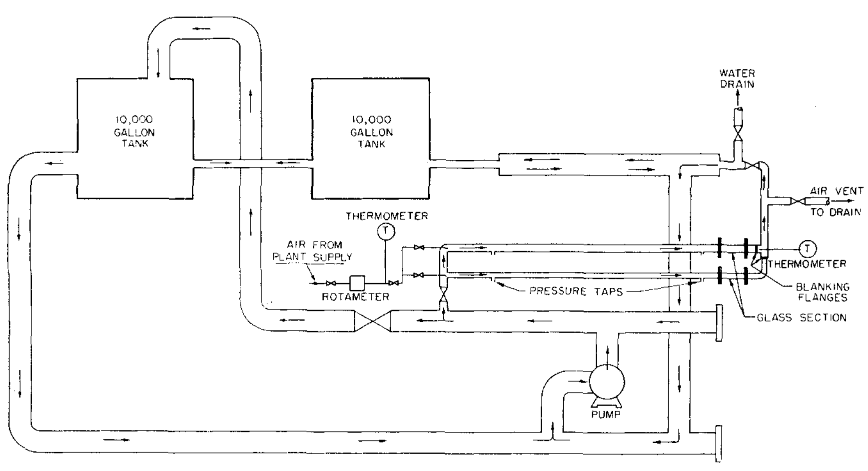

Fig. 2. Flow sheet of system.

EXPERIMENT

To investigate two-phase flow in large pipes, both 4- and 6-in. commercial steel pipes were installed in parallel on a straight horizontal run of 76 ft. A flow sheet of the system is shown in Figure 2. The pipes were interconnected at both ends but provision was made for the insertion of blank flanges at the downstream end to permit the full flow to be directed through each pipe separately. A water supply was connected to the upstream end, and an orifice meter and gate valve were installed in the supply line to measure and control the water-flow rate. Water was circulated through the system and back to surge tanks by a pump. Air, taken from a plant air header, was throttled by a gate valve and measured by a rotameter.

To measure the pressure differential down the length of a pipe, taps were placed along the bottom of each pipe 56 ft. apart, with calming sections of 10 ft. at each end to minimize the effects which bends leading into and away from the pipes might have had on flow within the test section. The pressure differential between the bottom taps was transmitted by water through ⅛-in. copper tubing to a vertical 50-in. manometer containing 2.95 specific gravity Meriam oil. The vertical manometer could be read to the nearest 0.05 in. Snubbers at the manometer terminals dampened the fluctuating pressure of the system, allowing measurement of the average pressure differential; however the snubbers could be by-passed to permit readings with partial or no snubbing. The absolute system pressure was measured by Bourdon-tube (1 to 100 lb./sq. in. gauge) pressure gauges at the first tap of each test section.

Glass sections 12 in. long were installed in the downstream end of each pipe, so that the flow pattern might be observed. Bimetallic thermometers indicated the temperature of the air before it passed to the pipe test sections and the temperature of the two phases as the fluid left the test section. In almost all the experimental runs the temperature of the air and water remained between 20° and 25°C.

In each series of runs the water-flow rate was kept constant and the air-flow rate was increased from 100 to about 500 std. cu. ft./min. Pressure differentials were measured with snubbers in the transmitting lines to damp out any oscillations caused by the slugging flow of the air and water. To ensure that equilibrium was attained with the snubbers in the line, measurements were made only after the manometer level became constant. For each run two pressure differentials were measured and averaged. To indicate the reproducibility of the data, at the end of each series of runs every fourth run of the series was repeated.

Pressure drops resulting from the slight increase in gas volume due to water vaporization or to the increase in momentum of the exit gas stream were estimated and found to be negligible compared to the measured pressure drop.

Single-phase pressure drops were experimentally determined and were compared with values calculated by use of roughness factors of 0.00015 and 0.0015 ft., corresponding to new and old steel pipe. The experimental pressure drops were found to be between the calculated values, an indi-

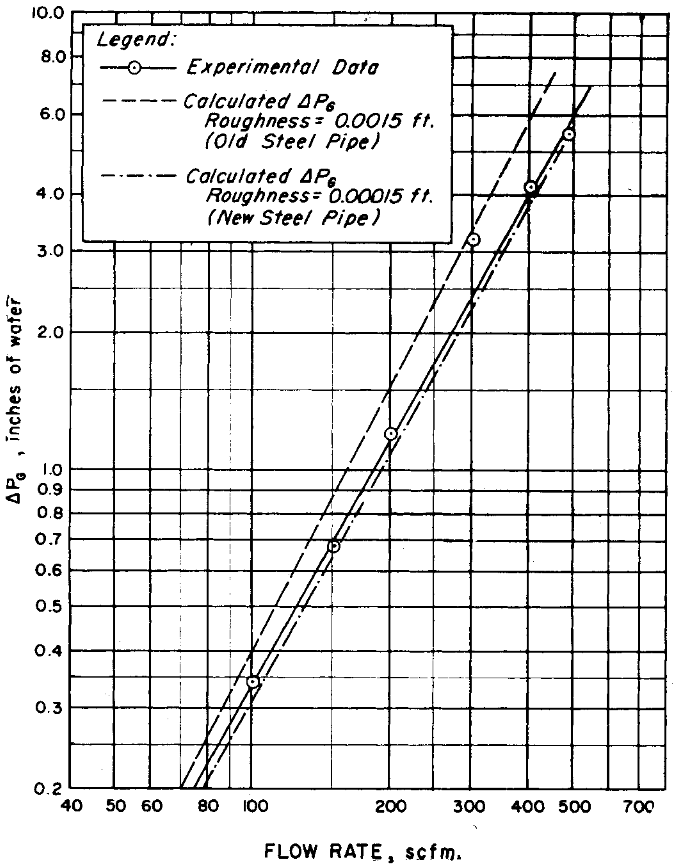

Fig. 3. ΔP_G vs. flow rate, gas phase, 4-in. pipe.
NOTE: Air flow at atmospheric pressure.

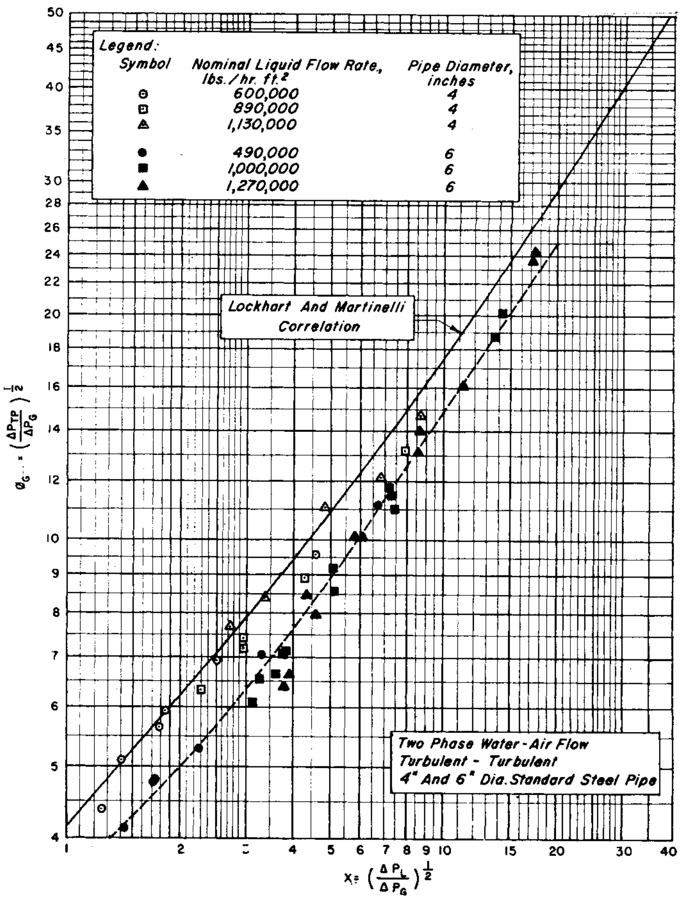

Fig. 4. ϕ_G vs. X for 4- and 6-in.-diam. standard steel pipes.

cation that the roughness of the wall was typical of steel pipe in service. Figure 3 shows the results of one such comparison.

RESULTS AND DISCUSSION OF RESULTS

The experimental data and calculated results are summarized in Table 1. The range of water- and air-flow rates investigated was 100 to 198 gal./min. with 100 to 400 std. cu. ft./min. for the 4-in. pipe and 192 to 500 gal./min. with 100 to 485 std. cu. ft./min. for the 6-in. pipe. With these high flow rates the flow type investigated here was turbulent-turbulent throughout.

A plot of the Lockhart-Martinelli correlation for turbulent-turbulent flow is shown in Figure 4, along with the experimental data taken in the present investigation. The 4-in. data are well represented by the correlation. The 6-in. data fell about 20% below the correlation but on a smooth parallel curve.

Comparison of the experimental data with the Chenoweth and Martin correlation in Figure 5 showed excellent agreement and no noticeable effects of pipe diameter. The family of curves shown in Figure 1 for different values of $\Delta P_G^*/\Delta P_L^*$ are not shown in Figure 5 since at high values of the superficial

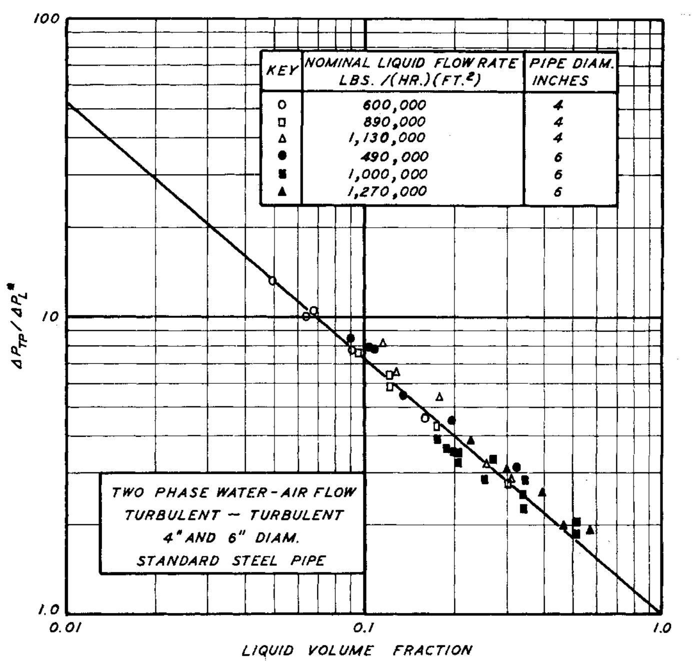

Fig. 5. Turbulent-turbulent two-phase water-air flow in 4- and 6-in.-diam. standard steel pipes.

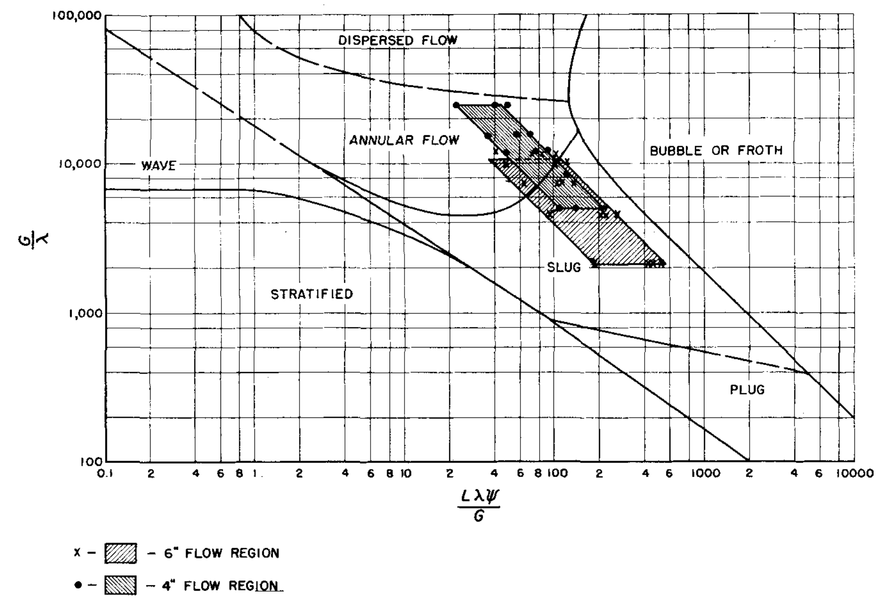

Fig. 6. Baker's correlation for flow-pattern regions.

liquid-volume fraction, all curves converge to a single line. In fact, above a superficial liquid volume fraction of 0.1, i.e., from 10 to 100% liquid, two-phase pressure drops may be estimated as follows:

$$\Delta P_{TP} = \Delta P_L^* \, (LVF)^{-0.86}$$

This relationship becomes especially interesting when it is observed that if one assumes the single-phase friction factor correlating charts to be applicable to two-phase flow problems, and superficial average densities and velocities are employed, one then obtains a similar relation, i.e.,

$$\Delta P_{TP} = \Delta P_L (LVF)^{-1}$$

COMPARISON WITH BAKER'S RESULTS

A comparison of the experimental results reported here with the data reported by Baker (*2*) on large-diameter pipes involves the concept of flow patterns. As noted previously, the flow pattern is believed to be a function of both the mass velocity and the ratio of the mass velocities of the liquid and gas flow. Baker correlates the mass velocity of the gas and the ratio of the mass velocities of the gas and liquid with the various flow-pattern regions. On Figure 6†, which shows Baker's correlation, experimental data points taken during the present investigation are plotted. If Figure 6 correctly defines the flow-pattern regions, the data taken here should be in both the slug and annular regions, although only slug flow was observed visually. The boundaries between patterns in Figure 6 are not definite and represent the average of the transition region between different patterns. Thus the data reported here may be interpreted as being between slug and annular flow.

Baker found that for turbulent-turbulent, two-phase flow in large-diameter pipes the Lockhart-Martinelli correlation requires modification depending upon what flow pattern is present. For the slug-flow region Baker suggested

$$\phi_G = \frac{1190X^{0.815}}{L^{0.5}}$$

and for the annular flow region

$$\phi_G = (4.8 - 0.3125D)X^{(0.343 - 0.021D)}$$

The largest value of L in the slug-flow region investigated by Baker was 228,000; the minimum used here was 490,000. The data shown on Figure 4 cover a range in L of 490,000 to 1,270,000 and show no noticeable trends due to changes in L.

Baker's annular-flow-pattern correlation predicts that at a constant value of X, ϕ_G decreases as the diameter increases. This prediction agrees with the results reported here, although the use of Baker's equation for values of X much greater than unity yields curves well below the experimental values reported in Table 1. The largest value of X used by Baker for the annular region was 0.694 (8-in. pipe) and the minimum used in the present investigation was 1.2 (4-in. pipe).

†The Ψ and λ terms are introduced in the correlation to allow for variation in the physical properties of different fluids.

$\Psi = (73/\gamma)/[\mu_L(62.3/\rho_L)^2]^{1/3}$
where ρ_L = density of liquid, lb./cu. ft.
γ = surface tension of liquid, dynes/cm.
μ_L viscosity of liquid, centipoises
$\lambda = [(\rho_G/0.075)(\rho_L/62.3)]^{1/2}$
where ρ_G = density of the gas, lb./cu. ft.

CONCLUSIONS

Comparative application of the Lockhart-Martinelli and Chenoweth-Martin correlations indicates that the latter is more reliable in the range of pipe sizes and mass flow rates studied here. Extrapolation of this finding beyond the limits of this investigation, however, cannot be advised in view of the still very incomplete development of flow theory as applied to large pipes.

The comparison of the results of the present investigation with Baker's results may be summarized as follows: the effect of diameter on the pressure drop in the annular-flow-pattern region agrees qualitatively with that predicted by Baker; however, contrary to Baker's conclusions, no effect of liquid-flow rate was noted in the slug-flow region. Baker's data were taken for oil and gas at high pressure and at much lower liquid mass velocities than were the present data and should, perhaps, not be used for a valid comparison.

NOTATION

D = pipe diameter, in.
G = gas flow rate, lb./(hr.)(sq. ft.)
L = liquid flow rate, lb./(hr.)(sq. ft.)
LVF = superficial liquid volume fraction
ΔP_G or ΔP_L = pressure drop, as calculated from single-phase correlations if gas or liquid is flowing alone in pipe at same rate as in two-phase flow
ΔP_G^*, ΔP_L^* = fictitious single-phase pressure drop, as calculated from single-phase correlation if gas or liquid is flowing alone in pipe at the total mass flow rate of both streams. For ΔP_G^* the average physical properties of gas are used; for ΔP_L^* the average physical properties of liquid are used
ΔP_{TP} = two-phase pressure drop
ϕ_G = correlating factor, $(\Delta P_{TP}/\Delta P_G)^{1/2}$
X = correlating factor, $(\Delta P_L/\Delta P_G)^{1/2}$

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