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Design Of Pipelines For The Simultaneous Flow Of Oil And Gas

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ABSTRACT

Measurements were made of pressure drops for oil and gas flowing simultaneously through four to ten inch diameter pipelines. These results were correlated with data from the literature for one-half to three inch diameter pipelines. Previously published data were used to construct a generalized chart for predicting the type of flow pattern in the pipeline. The flow patterns described are bubble flow, plug flow, stratified flow, wavy flow, slug flow, annular flow, and spray flow. Our experimental data were shown to be consistant with the generalized flow pattern chart.

The calculation method proposed by Lockhart and Martinelli[12] for designing two-phase pipelines was shown to be inadequate for larger diameter lines and also for some flow patterns. Modifications of the above method in the form of separate equations for each type of flow pattern are presented.

INTRODUCTION

In recent years there has been an increasing tendency to transport both oil and gas simultaneously through pipeline gathering systems. Trends toward central separator batteries, crude stabilizations, and total well stream processing in distillate type fields have emphasized the need for better information on the effects of liquid flowing along with the gas in pipelines.

The manyfold increase in pressure drop when liquid is deliberately put into a gas pipeline is illustrated by Van Wingen's[1] data in Figure 1. A look at the mechanism of two-phase flow will explain why pressure drops increase so much.

When two phases flow through the same pipe the gas usually flows faster than the liquid. The liquid accumulates in the pipe and reduces the cross sectional area available for gas flow. The pressure loss of a fluid flowing through a pipe is inversely proportional to the fifth power of the pipe diameter. The accumulated liquid in the line has the effect of reducing the diameter. A reduction of 20 per cent in diameter would cause a threefold increase in pressure drop, while a 60 per cent reduction would increase pressure drop one hundred times.

At some ratios of gas to liquid flowing, the surface of the liquid is very disturbed. There are many projections of liquid waves or ridges into the gas stream. For a half century studies have been made of the effect of roughness of inside pipe surface on pressure drops. Various friction factor charts taking roughness into account are available.[2] These charts, depending on the Reynolds number level, show that pressure drop for very rough pipes may be two to

References and illustrations at the end of paper tenfold greater than for smooth pipes.

The major factor causing high pressure drop is the energy required to move the liquid through the line. For most cases this is energy supplied by the gas. Additional energy is used in the violent rising and falling of the liquid in the line. This energy must come from a reduction in pressure. When this factor is combined with the effects of decreased pipe diameter and roughness it becomes more apparent why large pressure drops occur.

FLOW PATTERNS

Several flow patterns have been recognized in two-phase flow. Sketches[3] of the various types are shown in Figure 2. Alves[3] has described these as follows: "Assume a horizontal pipe with liquid flowing so as to fill the pipe and consider the types of flow that occur as gas is added in increasing amounts.

a. "Bubble Flow: Flow in which bubbles of gas move along the upper part of the pipe at approximately the same velocity as the liquid." This type is similar to Froth Flow where the entire pipe is filled with a froth similar to an emulsion.

b. "Plug Flow: Flow in which alternate plugs of liquid and gas move along the upper part of the pipe.

c. "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.

d. "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.

e. "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 the pipe at a much greater velocity than the average liquid velocity.

f. "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 as a central core.

g. "Spray Flow: Flow in which most or nearly all of the liquid is entrained as spray by the gas." This has also been called Dispersed Flow.

Most investigators of two-phase flow have used glass or plastic sections in their pipes and observed the types of flow. They have reported the flow rates of gas and liquid for each flow type at their experimental conditions. This was not practical for the high pressures used in our experiments. Figure 3 is a generalized plot of the flow pattern regions which we have prepared from the data of Jenkins, Gazley, Alves and Kosterin.[4,5,3,6,7]

Figure 3 shows the boundaries of the various flow pattern regions as functions of G, the mass velocity of the gas phase, and L/G, the ratio of mass velocities of the liquid and gas phase.

Since most of the available data were for the air-water system at atmospheric pressure, correction factors have been introduced to adjust for other liquids and gases. Holmes suggested[8] these terms for correlating the flooding point in wetted wall distillation columns. The gas mass velocity is divided by $\lambda = \sqrt{(\rho_G/0.075)(62.3/\rho_L)}$ and the L/G ratio is multiplied by $\lambda\psi$ where

$$\psi = \frac{73}{\gamma}\left[\mu_L + \left(\frac{62.3}{\rho_L}\right)^2\right]^{1/3}.$$

ρ_G and ρ_L are the gas and liquid densities at flowing conditions in pounds per cubic foot. The surface tension of the liquid γ, is in dynes per centimeter and the liquid viscosity, μ_L, is in centipoise.

Although the borders of the various flow pattern regions in Figure 3 are shown as lines, in reality these borders are rather broad transition zones. Each observer probably selected the transition at slightly different points. Not all observers have used the same nomenclature and it was necessary to equate terms in some cases. Figure 3 is based on data from one, two and four inch pipe.

LITERATURE ON PRESSURE DROP

Much of the early work on two-phase pressure drop was published by workers at the University of California starting in 1939.[9,10,11,12,13] These studies resulted in the correlation presented by Lockhart and Martinelli in 1949.[12] Their method is as follows: Calculate the pressure drop of the liquid phase assuming that it is the only fluid flowing in the pipeline. A similar calculation is made for the gas phase. The term $X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$ is evaluated from the above results. The two-phase pressure drop is given by the equation:

$$\Delta P_{TP} = \Delta P_G \, \phi_G^2 \qquad \ldots (1)$$

where ΔP_G is the pressure drop for the gas phase calculated above and ϕ_G is a function of X. A similar equation

$$\Delta P_{TP} = \Delta P_L \, \phi_L^2 \qquad \ldots (2)$$

based on the pressure drop for the liquid phase may be used. The multipliers ϕ_G and ϕ_L are functions of the factor X. The terms are related by the equation:

$$\phi_G = X \, \phi_L \qquad \ldots (3)$$

Lockhart and Martinelli presented the relationship between X, ϕ_L and ϕ_G in 1/2 inch and one inch pipe for four cases. These cases were: (1) liquid turbulent and gas turbulent; (2) liquid viscous and gas turbulent; (3) liquid turbulent and gas viscous; and (4) liquid viscous and gas viscous. The Reynolds number at which the flow changes from viscous to turbulent is uncertain but a value between 1000 and 2000 is used. We have arbitrarily used a value of 1000 in this paper. Lockhart and Martinelli did not consider the effect of the various flow patterns in their correlation although their basic assumptions tended to limit it to annular flow.

When their paper was presented the data of Jenkins[4,12] cited during the discussion, indicated that additional terms would be required to accurately predict the pressure drop. Later in 1949 Gazley and Bergelin[14,15] at the University of Delaware presented data on stratified and wave flow in a two inch pipe. They obtained pressure drops considerably lower than those predicted by the Lockhart and Martinelli curve. Their results suggested that the Lockhart analysis was not valid for stratified flow or that two inch pipe had a different relationship between ϕ_G and X.

Data for crude oil well streams at various gas-oil ratios were reported by Van Wingen in 1949.[1] He measured two-phase pressure drops in the gathering system of a San Joaquin Valley field. His data in three inch pipe at 450 psig showed slightly higher pressure drops than an earlier correlation of Martinelli.[10] Van Wingen's data were in terms of stock tank oil and gas off the high and low stage separators. The phase behavior of this field was reported by Olds, Sage and Lacey.[16] Their data have been used to convert Van Wingen's oil and gas volumes to those present at pipeline conditions, as shown in Figure 1.

Johnson and Abou-Sabe[17] presented data on pressure drop for the air-water systems in 0.87 inch ID pipe. Their data, like that of Jenkins[4] in a 1.02 inch ID pipe, demonstrated that more accurate correlations were obtained if curves similar to Lockhart's were drawn through points having the same liquid flow rate.

Alves[3] concluded from his one-inch pipe studies that it was necessary to take into account the type of flow pattern when estimating two-phase pressure drop. He gave several suggestions for adjusting the pressure drop calculated by the Lockhart and Martinelli correlation.

EXPERIMENTAL DATA

Rodman Jenkins of Magnolia Petroleum Company's Field Research Laboratories, was of the opinion that the pressure drops would not be as great as predicted in large diameter pipelines used in condensate gathering systems. To test this point, in 1951, H. R. Inman of the Magnolia Natural Gas Department made experiments 13 through 27 listed in Table I. These experiments included 4, 6, 8 and 10 inch lines. The test conditions were the ordinary operating conditions in the field at the time. Although gas-oil ratios in the lines were varied somewhat by shutting in some wells, the tests did not cover a broad enough range to be conclusive.

In 1953, additional experiments were made jointly by Magnolia Petroleum Company and Continental Oil Company. These tests were designed to cover gas-oil ratios from 1000 to 50,000. A section of a condensate field gathering system was selected for the test. It consisted of about two miles of eight inch line and eight miles of ten inch line in series. A tank farm was built at the inlet end of the test section. Metering equipment and pumps for injecting the oil into the 1000 psig line were installed. Liquid condensate was accumulated until the tanks were full and then pumped into the gas line. Gas and liquid flow rates were adjusted to the desired values and held constant until measured pressures along the line remained the same for two hours. The gas and liquid were separated and metered at the central separator station at the outlet of the test section. Pressures were measured at seven test points along the line with calibrated dead weight testers. The pipeline was substantially level. There was a one foot rise from inlet to outlet in the eight inch section and a four foot rise in the ten inch section. Corrections for elevation are less than the accuracy of the dead weight testers. Samples of the high pressure liquid and gas were analyzed. Viscosities of the high pressure liquid were determined in the laboratory. Other properties of the fluids were calculated from the analyses[18,19,20,21] Compositions are given in Table II. Fluid compositions for the Inman tests were estimated from flash calculations of well streams. The results are given in Figure 4 and Table I, Experiments 1 through 12.

Pressure drops for the separate liquid and gas phases were calculated by the following modifications of the Fanning equation:

$$\triangle P_G = \frac{Q^2 L' fSTZ}{20{,}000\ D^5 P_{avg}} \quad \dots (4)$$

$$\triangle P_L = \frac{fL'(Bbl/Day)^2(Lb/Gal)}{181{,}916\ D^5} \quad \dots (5)$$

The friction factors used are given in Table 3. They are our best estimates for the probable condition of the pipeline.

The liquid hold-up in the test section was determined for each run by a material balance. It was not possible to segregate the hold-up in the eight and ten inch lines. Figure 5 shows the liquid hold-up as a per cent of the total volume of the test section.

DISCUSSION OF RESULTS

It is apparent from Figure 4 that pressure drops for large pipes are 40 to 60 per cent less than predicted by the Lockhart and Martinelli curve for a broad range of X values. The sudden change in slope in Figure 4 suggests that some radical change has taken place in the type of flow. The divergence of the data into two distinct lines, one for eight inch and one for ten inch pipe, implies that the pipe diameter has a marked effect for this second type of flow. The fact that the ϕ_{GTT} values are higher for the ten inch pipe is significant.

When the type of flow pattern for each experimental point was estimated from Figure 3, it was found that the section of the curve parallel to the Lockhart and Martinelli correlation was in annular flow. The change of slope corresponded to the start of the slug flow pattern and Figure 3 indicated that the fluids in the ten inch pipe entered slug flow at lower values of X than the eight inch pipe. These data demonstrate that it is not practical to disregard the flow pattern regions in devising a two-phase pressure drop correlation.

The data points available were segregated into groups having the same flow pattern and each set of data studied separately.

SLUG FLOW

For one inch and smaller pipe the data of Jenkins[4], Alves[3], and Abou-Sabe[17] were considered. The Lockhart and Martinelli curve probably is the best line that can be drawn through these data, although deviations as great as ± 25 per cent from the predicted ϕ_{GTT} values may be expected. Van Wingen's[1] data for three inch pipe and the Magnolia-Continental test data for eight inch and ten inch pipe are shown in Figure 6. It is apparent that ϕ_{GTT} is inversely proportional to the square root of the liquid mass velocity. These data are correlated by the equation:

$$\text{for slug flow } \phi_{GTT} = \frac{1190\ X^{0.815}}{L^{0.5}} \quad \ldots (6)$$

with average deviations of +8.9 per cent, -3.7 per cent and maximum deviations of +17.4 per cent, -8.7 per cent. Figure 7 compares the equation with the Lockhart and Martinelli curve for various liquid mass velocities. Since all of the data used in arriving at this equation were for hydrocarbon systems, it should be used with caution on dissimilar cases.

ANNULAR FLOW

Data for annular flow are shown in Figure 8. In this case pipe diameter is a controlling factor. At constant values of X, as diameter increases the ϕ_{GTT} value becomes smaller. Data for 0.87 inch[17], 1 inch[4], 3 inch[1], 4 inch, 6 inch, 8 inch and 10 inch show this trend. The straight lines drawn through the data points for each diameter appear to be converging to a point. At X = 0.01 the ϕ_{GTT} value becomes 1.0. In gas transmission formula terms this means that the pipeline efficiency becomes 100 per cent. Obviously this cannot be true since very small amounts of liquids in gas transmission lines will reduce efficiencies to as low as 70 per cent.

The relation between ϕ_G and E[22], the fractional efficiency of a gas pipeline is

$$\phi_G = \sqrt{\frac{1}{E}} \quad \ldots (7)$$

At E = 0.9, ϕ_G will equal 1.0585, and E = 0.7 would give a ϕ_G value of 1.196 and an E = 0.1 is required for ϕ_G of 3.16.

The following formula is suggested for correlating the data in Figure 8.

For Annular Flow

$$\phi_{GTT} = (4.8 - 0.3125D)X^{0.343-0.021D} \quad \ldots (8)$$

The average deviations for the data used are +15 per cent, - 7 per cent, and the maximum deviations are +71 per cent, -13 per cent.

FROTH OR BUBBLE FLOW

Abou-Sabe[17] and Van Wingen's[1] data for 0.87 inch and three inch pipe in the froth flow pattern are correlated in Figure 9. As pointed out by Alves[3], the Lockhart and Martinelli curve represents these data very well. The introduction of a small correction for liquid mass velocity improves the accuracy slightly. Equation (9), below, has an average deviation of +8.8 per cent, -6.2 per cent and a maximum deviation of +21 per cent, -14 per cent.

For Froth Flow

$$\phi_{GTT} = \frac{14.2\ X^{0.75}}{L^{0.1}} \quad \ldots (9)$$

PLUG FLOW

Only seven data points were available for plug flow.[1,17] However, the data showed a consistant effect of liquid mass velocity. Equation (10) is proposed for this case. Data are limited to 0.87 inch and three inch pipe.

For Plug Flow

$$\phi_{GTT} = \frac{27.315\ X^{0.855}}{L^{0.17}} \quad \ldots (10)$$

The average deviation of the experimental points is +9.7 per cent, -7.7 per cent, while the maximum deviations are +15 per cent, -20 per cent.

WAVE FLOW

The available wave flow data are plotted in Figure 11.[4,14] These data indicate that the Lockhart and Martinelli curve give pressure drops 1.5 to 1.7 times the experimental values. At high values of gas mass velocity Jenkins[4] data for one inch pipe approach a single line that may be approximated by an equation of the form $\phi_{GTT} = \text{Cst.}\ X^{0.25}$. At lower gas mass velocity values the data diverge into separate curves for each liquid mass velocity. An equation of the form

$$\phi_{GTT} = \frac{\text{Cst.}\ X^{0.75}}{L^{0.65}}$$ will represent these data satisfactorily.

Additional experimental work at higher X values in larger pipe is needed to evaluate pressure drops for wave flow. This type of flow occurs often in gathering lines in condensate fields so we may expect to learn more about it in the future.

STRATIFIED FLOW

Data for 1, 2, 8 and 10 inch pipe for fluids in stratified flow are shown in Figure 12.[4,14] More experimental work is needed for this flow pattern. Indications are that the pressure drop calculated by the Lockhart and Martinelli curve may be 1.5 to 2.5 times greater than experimental values. However, the data points for one inch and ten inch pipe fall very close to the Lockhart and Martinelli correlation.

OTHER TYPES OF FLOW

In this paper all the experimental data considered were for both gas and liquid phases in turbulent flow. This is the usual case in industrial applications. For data covering the other cases the references should be consulted. Alves[3] and Gazley[14] give helpful data for one and two inch pipe sizes.

DESIGN SUGGESTIONS

The modifications of the basic Lockhart and Martinelli correlation proposed in this paper are probably fairly reliable, but the similarity of new designs to our experimental conditions should be considered carefully in each case. The limits of error for the various equations should be studied before a safety factor or load factor for the flowing quantities is selected.

It will be noted that the measured pressure drops were always small fractions of the total pressure. It is suggested that if the calculated pressure drop is more than 10 per cent of the downstream absolute pressure the pipeline be calculated in two or more sections. The magnitude of the errors involved for gas flow are discussed by Poettmann[23] and Clinedinst[24].

It should be emphasized that the quantities and physical properties of the fluids used in the calculations must be those at pipeline conditions. Equilibrium flash calculations[25,26] or experimental phase studies will provide the best volume data. For quick preliminary estimates generalized correlations of gas solubility and formation volume factors[20] are useful. Estimates of liquid viscosity may be made with fair accuracy by the method of Katz and Bicher[18] if component composition data is available, or by the Beal method[21] if it is not. For surface tension values the Katz, Monroe and Trainer method[19] is suggested.

The data considered in this paper were taken on horizontal pipes only. Additional pressure drop may be expected in hilly country. The liquid being carried along with the gas will tend to run back down the hills and accumulate in the valleys. When the pipe gets full enough of liquid it will slug over the hill and into the next valley. Considerable pressure surges may be expected at this time.

It seems probable that lines in annular, froth, or dispersed flow will not allow the liquid to accumulate in the valleys. Lines in stratified or wave flow probably allow liquid to accumulate until slug flow starts, carrying the liquid over the hill. Perhaps pipe diameters may be selected that will keep uphill flow in a safe pattern. Kosterin[7] states that in a one inch pipe at a 72° angle with the horizontal, the flow patterns were not affected by the inclination at velocities exceeding ten feet per second. For his air-water system at atmospheric pressure this velocity corresponds to the start of froth flow. He found the greatest pressure surges in stratified flow at velocities in the range of 1.6 to 6.6 feet per second.

Data for vertical two-phase flow have been presented by various authors[8,27,28,29,30,31] Perhaps their data would be helpful in evaluating inclined two-phase flow.

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NOMENCLATURE

D = Inside diameter of pipe, inches
f = Friction factor for fluid flow - see Table 3
G = Mass velocity of the gas phase, pounds per hour per square foot of total pipe cross section area
L = Mass velocity of the liquid phase, pounds per hour per square foot of total pipe cross section area
L' = Length of pipeline in feet
P_{avg} = Average of inlet and outlet pressures of the pipeline, psia
ΔP_G = Pressure drop for the gas phase flowing in the pipe. The full cross sectional area of the pipe is used in evaluating the pressure drop, PSI
ΔP_L = Pressure drop for the liquid phase flowing in the pipe. The full cross sectional area of the pipe is used in evaluating the pressure drop.
ΔP_{TP} = Pressure drop for the two phases flowing simultaneously, pounds per square inch.
Q = Gas flow rate, thousands of standard cubic feet of gas per day at 14.65 psia and 60 F
Re = Reynolds number, $\dfrac{DG}{12\mu(2.42)}$
S = Specific gravity of flowing gas (Air = 1)
$X = \sqrt{\dfrac{\Delta P_L \text{ (PSI)}}{\Delta P_G \text{ (PSI)}}}$
Z = Compressibility factor of flowing gas

$$\phi_G = \sqrt{\frac{\Delta P_{TP}}{\Delta P_G}} = \phi_L X$$

ϕ_{GTT} = ϕ_G when both liquid and gas phases are in turbulent flow with Reynolds numbers above 1000.

$$\phi_L = \sqrt{\frac{\Delta P_{TP}}{\Delta P_L}} = \frac{\phi_G}{X}$$

$$\lambda = \sqrt{\frac{\rho_G}{0.075}\ \frac{62.3}{\rho_L}}$$

ρ_G = Density of gas phase, pounds per cubic foot

ρ_L = Density of liquid phase, pounds per cubic foot

$$\psi = \frac{73}{\gamma}\left[\mu_L + \left(\frac{62.3}{\rho_L}\right)^2\right]^{1/3}$$

γ = Surface tension of liquid phase, dynes per centimeter

μ_L = Viscosity of liquid phase, centipoise

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Table 2

Analysis of Fluid Streams from Outlet Separator (Run No. 3) Magnolia - Continental Tests

Component	Mol % Gas	Mol % Liquid
CO_2	0.10	—
C_1	95.03	22.95
C_2	3.26	4.32
C_3	0.84	2.71
iC_4	0.22	1.49
nC_4	0.26	1.77
iC_5	0.06	1.49
C_{6+}	0.14**	63.26*
	100.00	100.00

Engler Distillation of Oil

Volume % Distilled	°F
I.B.P.	127
5	185
10	229
20	267
30	310
40	356
50	411
60	468
70	516
80	579
90	627
EP	688
Res.	3.25
Loss	2.00
API @ 60 F	48.3

* MW C_{6+} = 149.5
22.9 gallons/mol

**MW C_{6+} = 98.9
16.677 gallons/mol

Table 2 (Cont.)

Viscosity of Liquid (Composition Change With Pressure)

PSIG	Viscosity Centipoise @ 77 F
600	0.657
700	0.635
800	0.614
900	0.593
1000	0.572
1070	0.557

Table 3

Friction Factors Used for Single Phase Flow Calculations

Reynolds Number, Re	Friction Factors - Commercial Pipe 1 - 4 inch	6 - 24 inch
1,000	0.0157	0.0157
2,000	0.0132	0.0126
3,000	0.0119	0.0110
10,000	0.0087	0.0078
40,000	0.0064	0.0056
100,000	0.0054	0.0046
150,000	0.0050	0.0042
400,000	0.0042	0.0037
1,000,000	0.0036	0.0032
4,000,000	0.0029	0.0027
10,000,000	0.0026	0.0023

Table 1
Experimental Data

Experiment No.	1	2	3	4	5	6	7
Run No.	6	3	2	4	1	5	6
MSCFD	26,970	25,552	12,050	11,886	6,474	4,348	26,970
Bbls/Day	514	5,484	4,167	6,592	4,970	5,420	514
Length Line, Ft	11,317	11,317	11,317	11,317	5,534	11,317	11,333
Inside Diameter, In.	7.750	7.750	7.750	7.750	7.750	7.750	10.136
Inlet Pressure, PSIG	983	1,007	972	977	964	940	964
Outlet Pressure, PSIG	964	975	962	960	958	930	945
Gas Gravity (Air=1)	0.59	0.59	0.59	0.59	0.59	0.59	0.59
Gas Density #/cuft	3.42	3.48	3.38	3.32	3.38	3.28	3.36
Liquid Density #/Gal	6.499	6.499	6.525	6.499	6.53	6.103	6.499
Gas Viscosity, cp	0.014	0.014	0.014	0.014	0.014	0.014	0.014
Liquid Viscosity, cp	0.577	0.574	0.58	0.578	0.58	0.587	0.577
Surface Tension, Dynes/cm	16.7	16.7	16.7	16.7	16.7	16.7	16.7
ψ	4.27	4.27	4.27	4.27	4.27	4.5	4.27
λ	7.62	7.7	7.58	7.58	7.58	7.71	7.62
Line Temperature, °F	75	80	69	78	66	82	75
Two Phase Δ P (PSI)	19.0	32	10	17	6	10	19
Gas Δ P (PSI)	11.39	10.04	2.63	2.48	0.380	0.407	11.47
Liquid Δ P (PSI)	0.0307	2.0	1.265	2.85	0.848	1.92	0.0322
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$	0.0518	0.446	0.694	1.0725	1.494	2.17	0.0529
ϕ_{GTT}	1.292	1.786	1.948	2.62	3.98	4.96	1.288
Re_G, Thousands	2,840	2,750	1,268	1,250	681	456.5	2,172
Re_L, Thousands	8.26	88.5	67.0	105.5	79.2	80.3	6.31
G #/hr/sq ft	155,500	147,307	69,500	68,500	37,400	25,000	90,500
L #/hr/sq ft	17,850	191,000	145,000	228,500	173,000	177,000	10,440
$\frac{G}{\lambda}$	20,430	19,200	9,160	9,030	4,930	3,240	12,000
$\frac{L \lambda \Psi}{G}$	3.73	42.7	67.5	108	149.5	245	3.70
Type Flow	Annular	Annular	Annular	Slug	Slug	Slug	Annular

Table 1 (Cont)
Experimental Data

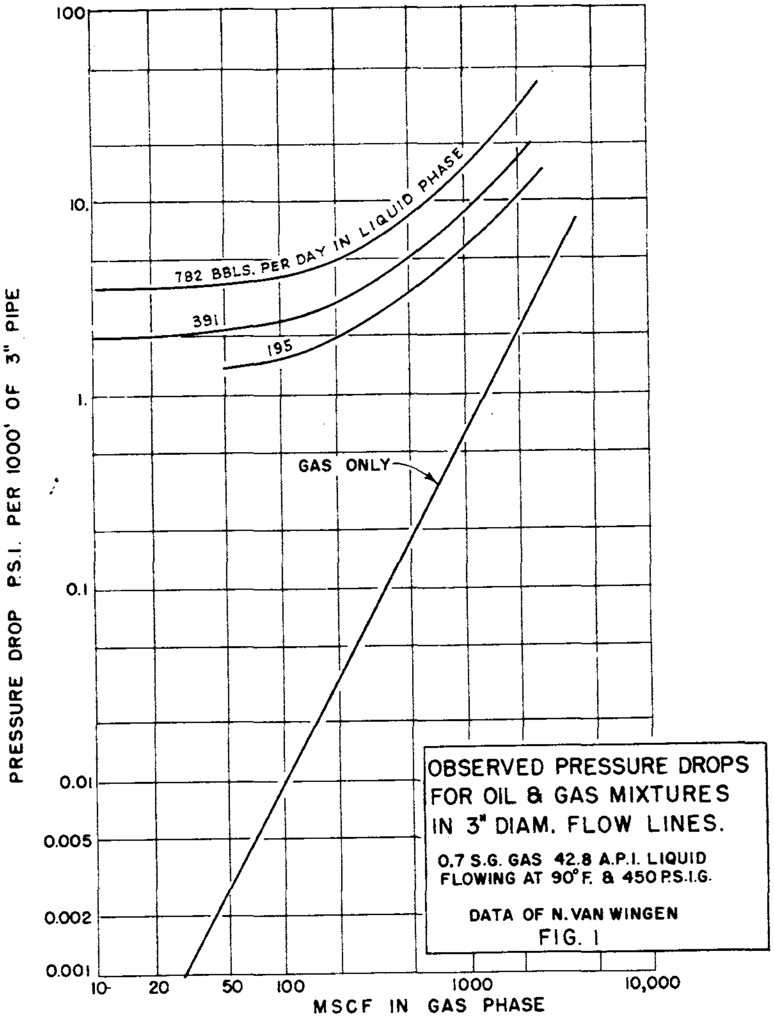
Experiment No.	8	9	10	11	12	13	14
Run No.	3	2	4	1	5	8	7
MSCFD	25,552	12,050	11,886	6,474	4,348	7,471.6	9,338.4
Bbls/Day	5,484	4,167	6,592	4,970	5,420	627	721
Length Line, Ft.	41,333	31,115	41,333	41,333	41,333	3,666	3,666
Inside Diameter, In.	10.136	10.136	10.136	10.136	10.136	4.026	4.026
Inlet Pressure, PSIG	975	962	960	952	930	1,087	1,096
Outlet, Pressure, PSIG	946	948	936	936	912	1,067	1,075
Gas Gravity (Air=1)	0.59	0.59	0.59	0.59	0.59	0.625	0.625
Gas Density #/cuft	3.37	3.38	3.32	3.32	3.28	4.31	4.31
Liquid Density #/Gal	6.499	6.525	6.499	6.53	6.103	5.11	5.11
Gas Viscosity, cp	0.014	0.014	0.014	0.014	0.014	0.0145	0.0145
Liquid Viscosity, cp	0.577	0.58	0.578	0.58	0.589	0.557	0.557
Surface Tension, Dynes/cm	16.7	16.7	16.7	16.7	16.7	16.7	16.7
ψ	4.27	4.27	4.27	4.27	4.5	4.98	4.98
λ	7.62	7.58	7.58	7.58	7.71	9.66	9.66
Line Temperature, °F	80	69	78	66	82	79	80
Two Phase $\triangle$ P (PSI)	29	14	24	16	18	20	21
Gas $\triangle$ P (PSI)	9.90	1.94	2.50	0.784	0.404	7.83	11.78
Liquid $\triangle$ P (PSI)	2.08	.964	2.876	1.74	1.937	0.298	0.386
$X = \sqrt{\frac{\triangle P_L}{\triangle P_G}}$	0.458	0.705	1.0725	1.49	2.19	0.195	0.181
ϕ_{GTT}	1.71	2.685	3.1	4.52	6.59	1.60	1.335
Re_G, Thousands	2,100	991	977	532	357	1,510	2,020
Re_L, Thousands	67.6	51.1	80.7	60.6	61.2	15.8	18.2
G #/hr/sq ft	85,727	40,400	39,800	21,700	14,600	158,000	210,000
L #/hr/sq ft	111,400	84,500	133,300	101,000	110,000	63,600	76,500
$\frac{G}{\lambda}$	11,400	5,320	5,250	2,790	1,890	16,300	21,700
$\frac{L\lambda\psi}{G}$	42.3	67.7	108	151	262	21.6	17.55
Type Flow	Annular	Slug	Slug	Slug	Slug	Annular	Annular

Table 1 (Cont)
Experimental Data

Experiment No.	15	16	17	18	19	20	21
Run No.	8	7	10	9	7	9	8
MSCFD	6,979.6	9,338.4	11,767	6,483.5	4,440.7	6,483.5	6,668.0
Bbls/Day	627	721	192	136	76	136	107
Length Line, Ft.	14,790	14,790	11,427	11,427	11,427	10,617	10,617
Inside Diameter, In.	5.937	5.937	7.750	7.750	7.750	7.750	7.750
Inlet Pressure, PSIG	1,070	1,076	712	705.5	1,075.5	703	1,067
Outlet Pressure, PSIG	1,055	1,060	703	703	1,075.0	701.5	1,065.5
Gas Gravity (Air=1)	0.625	0.625	0.62	.60	0.60	0.60	0.60
Gas Density #/cuft	4.28	4.28	2.62	2.62	3.99	2.62	3.99
Liquid Density #/Gal	5.11	5.11	6.15	6.15	5.68	6.15	5.68
Gas Viscosity, cp	.0145	0.0145	.0143	.014	0.014	0.014	0.014
Liquid Viscosity, cp	0.557	0.557	0.63	0.63	0.557	0.63	0.557
Surface Tension, Dynes/cm	16.7	16.7	18	18	15	18	15
ψ	4.98	4.98	4.25	4.25	5.17	4.25	5.17
λ	9.66	9.66	6.86	6.86	8.82	6.86	8.82
Line Temperature, °F	72	73	65	65	69	65	70
Two Phase $\triangle$ P (PSI)	15	16	9	2.5	0.5	1.5	1.5
Gas $\triangle$ P (PSI)	3.51	6.70	3.50	1.1	0.354	1.022	0.701
Liquid $\triangle$ P (PSI)	0.1685	0.2165	0.00586	0.00328	0.00113	0.003045	0.0019
$X = \sqrt{\frac{\triangle P_L}{\triangle P_G}}$	0.22	0.1795	0.0408	0.0545	0.0564	0.0545	0.052
ϕ_{GTT}	2.06	1.543	1.605	1.506	1.19	1.212	1.462
Re_G, Thousands	1,022	1,370	1,330	722	494	722	746
Re_L, Thousands	10.7	1,235	2.67	1.88	1.10	1.88	1.55
G #/hr/sq ft	72,300	96,300	71,100	37,800	25,900	37,800	38,900
L #/hr/sq ft	29,200	35,100	6,310	4,470	2,310	4,470	3,250
$\frac{G}{\lambda}$	7,480	10,000	10,350	5,500	2,940	5,500	4,410
$\frac{L\lambda\psi}{G}$	19.4	17.6	2.58	3.45	4.07	3.45	3.81
Type Flow	Annular	Annular	Wave	Wave	Stratified	Wave	Stratified

Table 1 (Cont)
Experimental Data

Experiment No.	22	23	24	25	26	27
Run No.	7	8	7	9	8	7
MSCFD	6,797.9	11,950	9,476.7	6,483.5	11,950	9,476.7
Bbls/Day	102	236	244	136	236	244
Length Line, Ft.	10,617	11,313	11,313	22,044	41,317	41,317
Inside Diameter, In.	7.750	7.750	7.750	7.750	10.136	10.136
Inlet Pressure, PSIG	1,075	1,064	1,074	705.5	1,063	1,068
Outlet Pressure, PSIG	1,074	1,062	1,067	701.5	1055.5	1,058
Gas Gravity (Air=1)	0.60	0.60	0.60	0.60	0.60	0.59
Gas Density #/cuft	3.99	3.99	3.99	2.62	3.99	3.88
Liquid Density #/Gal	5.68	5.68	5.68	6.15	5.68	5.68
Gas Viscosity, cp	0.014	0.014	0.014	0.014	0.014	0.014
Liquid Viscosity, cp	0.557	0.557	0.557	0.63	0.557	0.557
Surface Tension, Dynes/cm	15	15	15	18	15	16.7
ψ	5.17	5.17	5.17	4.25	5.17	4.65
λ	8.82	8.82	8.82	6.86	8.82	8.7
Line Temperature, °F	69	70	69	66	70	69
Two Phase $\triangle$ P (PSI)	1.0	2.0	7.0	4.0	7.5	10
Gas $\triangle$P(PSI)	0.721	2.22	1.38	2.12	2.195	1.413
Liquid $\triangle$ P (PSI)	0.00175	0.0074	0.00784	0.00633	0.00791	0.00818
$X = \sqrt{\frac{\triangle P_L}{\triangle P_G}}$	0.0491	0.0576	0.0753	0.0545	0.060	.076
ϕ_{GTT}	1.179	0.95	2.25	1.37	1.85	2.66
Re_G, Thousands	756	1,340	1,038	722	1,023	794
Re_L, Thousands	1.48	3.44	3.56	1.88	2.62	2.71
G #/hr/sq ft	39,600	69,600	55,200	37,800	40,600	31,800
L #/hr/sq ft	3,090	7,150	7,400	4,470	4,170	4,320
$\frac{G}{\lambda}$	4,500	7,900	6,260	5,500	4,610	3,660
$\frac{L \lambda \psi}{G}$	3.56	4.69	6.11	3.45	4.68	5.4
Type Flow	Stratified	Annular	Wave	Wave	Wave	Stratified

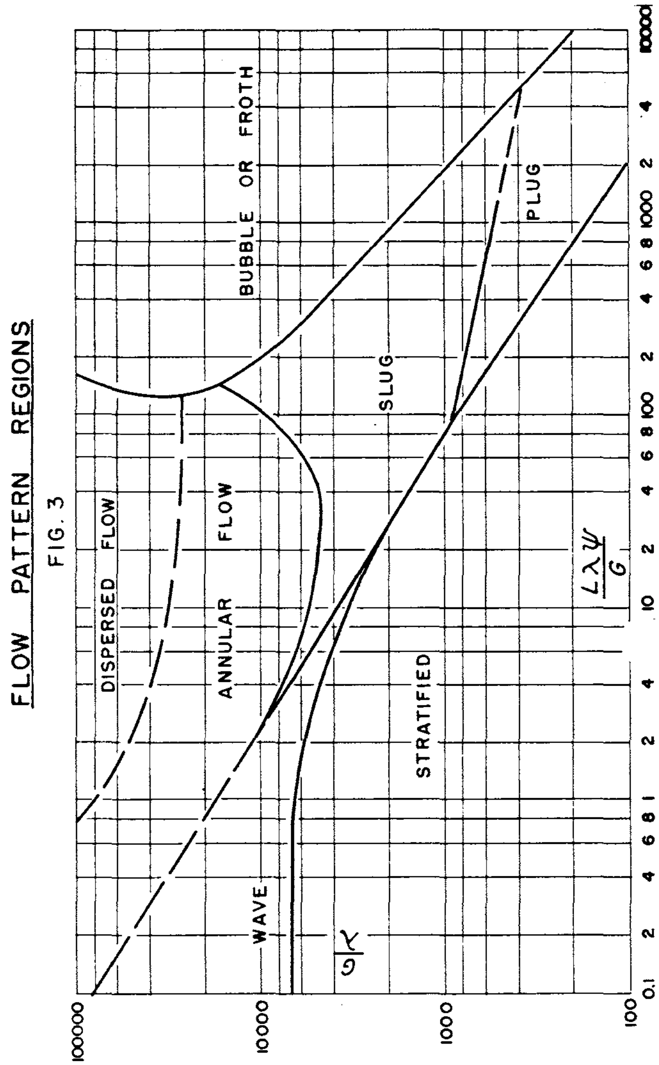


OBSERVED PRESSURE DROPS FOR OIL & GAS MIXTURES IN 3" DIAM. FLOW LINES.

0.7 S.G. GAS 42.8 A.P.I. LIQUID FLOWING AT 90°F. & 450 P.S.I.G.

DATA OF N. VAN WINGEN

FIG. 1



FLOW PATTERN REGIONS

FIG. 3

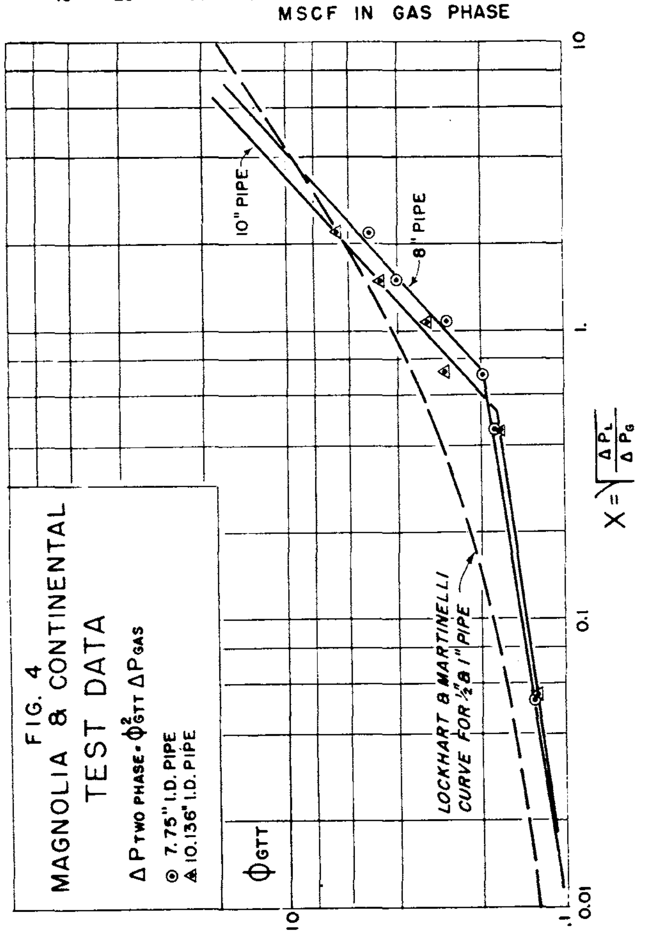
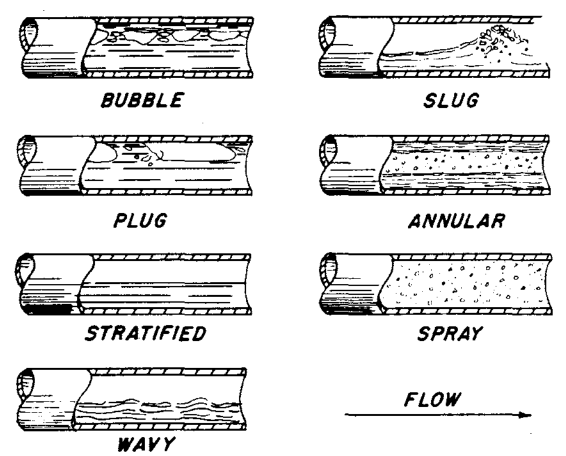


FIG. 4
MAGNOLIA & CONTINENTAL TEST DATA

$\Delta P_{TWO\ PHASE} = \phi^2_{GTT}\ \Delta P_{GAS}$

⊙ 7.75" I.D. PIPE
▲ 10.136" I.D. PIPE



FLOW PATTERN SKETCHES

FIG. 2

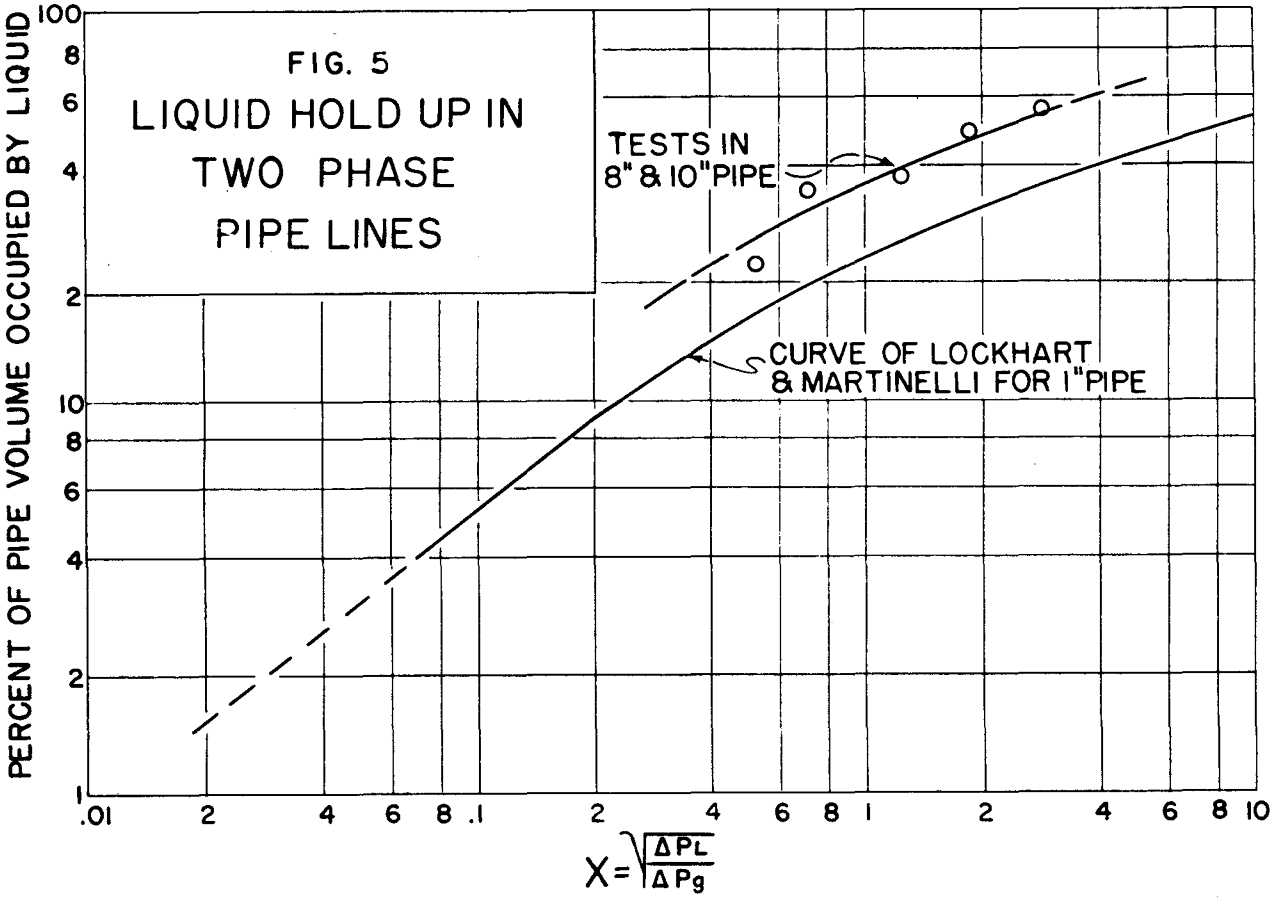


FIG. 5
LIQUID HOLD UP IN
TWO PHASE
PIPE LINES
TESTS IN
8" & 10" PIPE
CURVE OF LOCKHART
& MARTINELLI FOR 1" PIPE
PERCENT OF PIPE VOLUME OCCUPIED BY LIQUID
$X=\sqrt{\frac{\Delta P_L}{\Delta P_g}}$

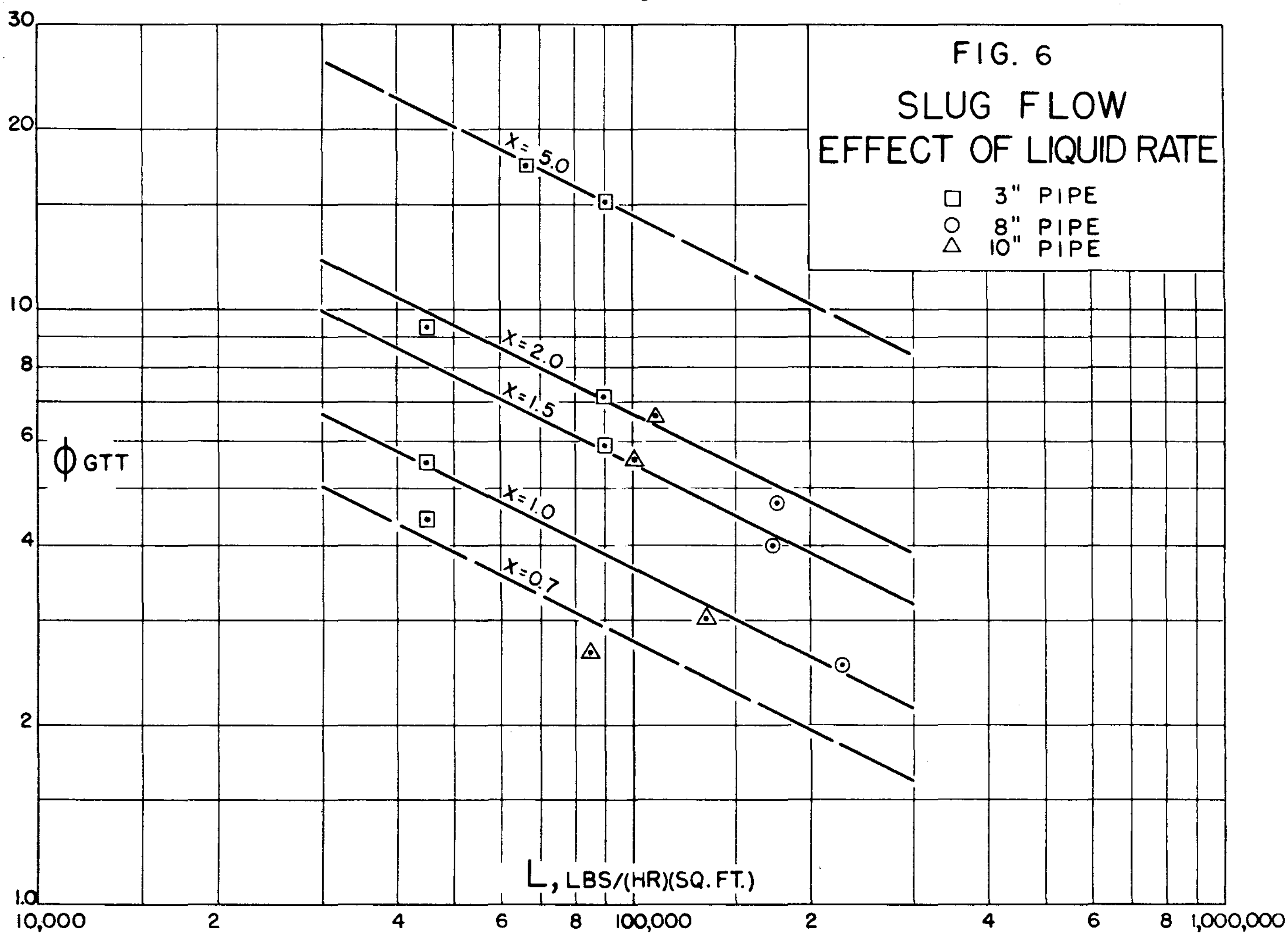


FIG. 6
SLUG FLOW
EFFECT OF LIQUID RATE
3" PIPE
8" PIPE
10" PIPE
X=5.0
X=2.0
X=1.5
X=1.0
X=0.7
ϕ_{GTT}
L, LBS/(HR)(SQ.FT.)

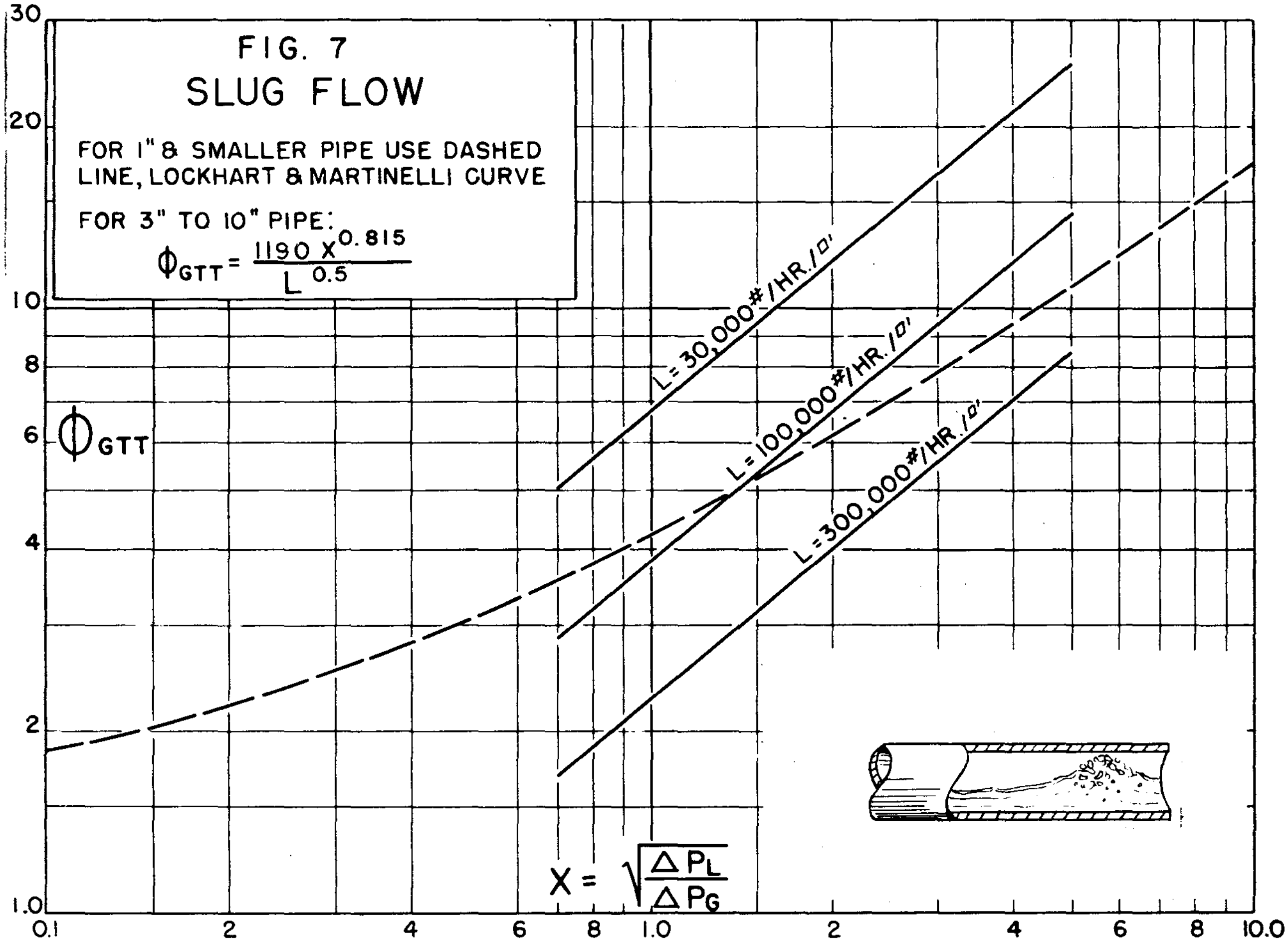


FIG. 7
SLUG FLOW
FOR 1" & SMALLER PIPE USE DASHED LINE, LOCKHART & MARTINELLI CURVE
FOR 3" TO 10" PIPE:
$\phi_{GTT} = \frac{1190\, X^{0.815}}{L^{0.5}}$
ϕ_{GTT}
L=30,000#/HR./□'
L=100,000#/HR./□'
L=300,000#/HR./□'
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$
30
20
10
8
6
4
2
1.0
0.1
2
4
6
8
1.0
2
4
6
8
10.0

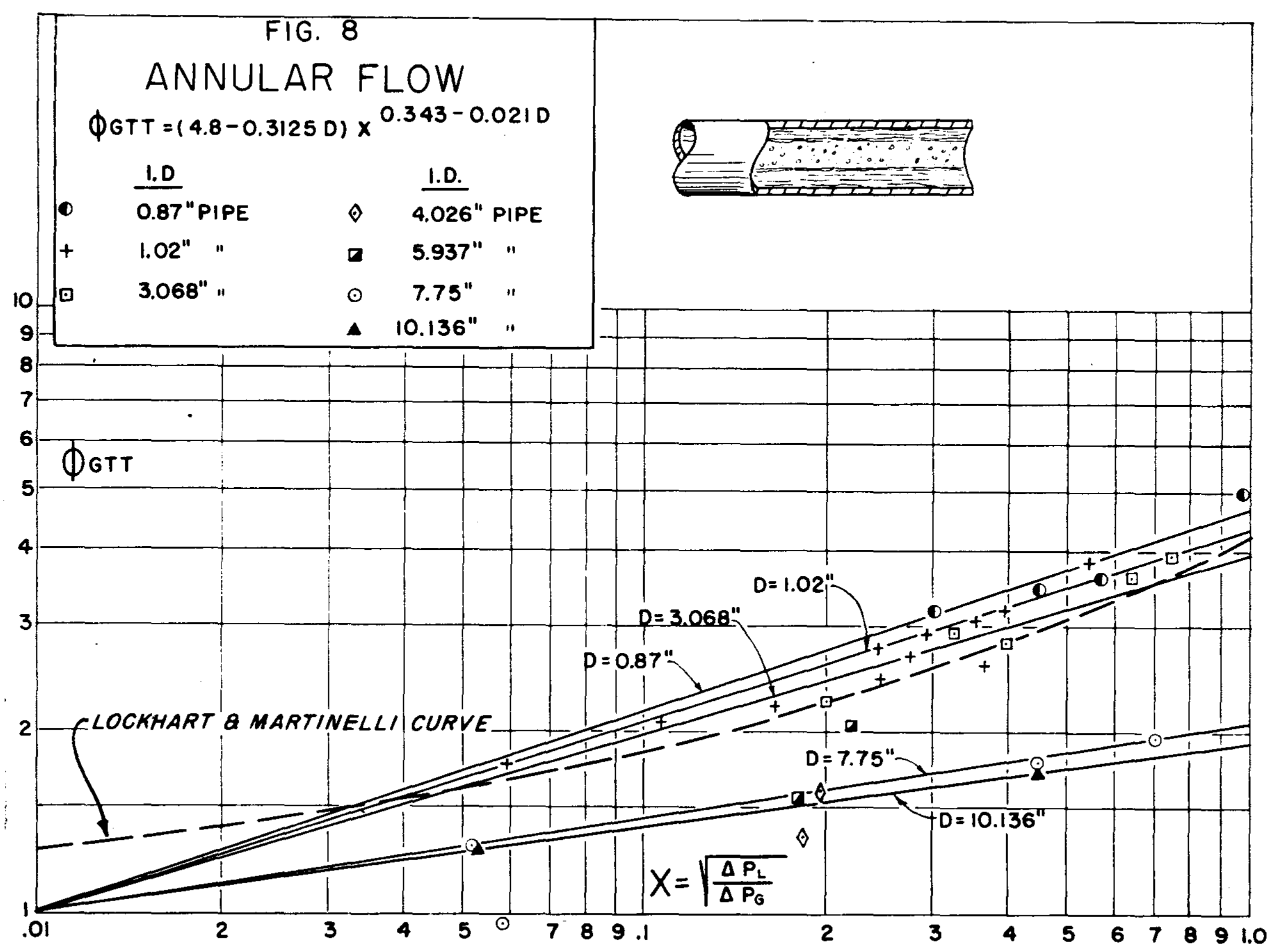


FIG. 8
ANNULAR FLOW
$\phi_{GTT} = (4.8 - 0.3125\,D)\, X^{0.343 - 0.021\,D}$
I.D
0.87" PIPE
1.02" "
3.068" "
I.D.
4.026" PIPE
5.937" "
7.75" "
10.136" "
ϕ_{GTT}
D=1.02"
D=3.068"
D=0.87"
LOCKHART & MARTINELLI CURVE
D=7.75"
D=10.136"
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$
10
9
8
7
6
5
4
3
2
1
.01
2
3
4
5
7
8
9
.1
2
3
4
5
6
7
8
9
1.0

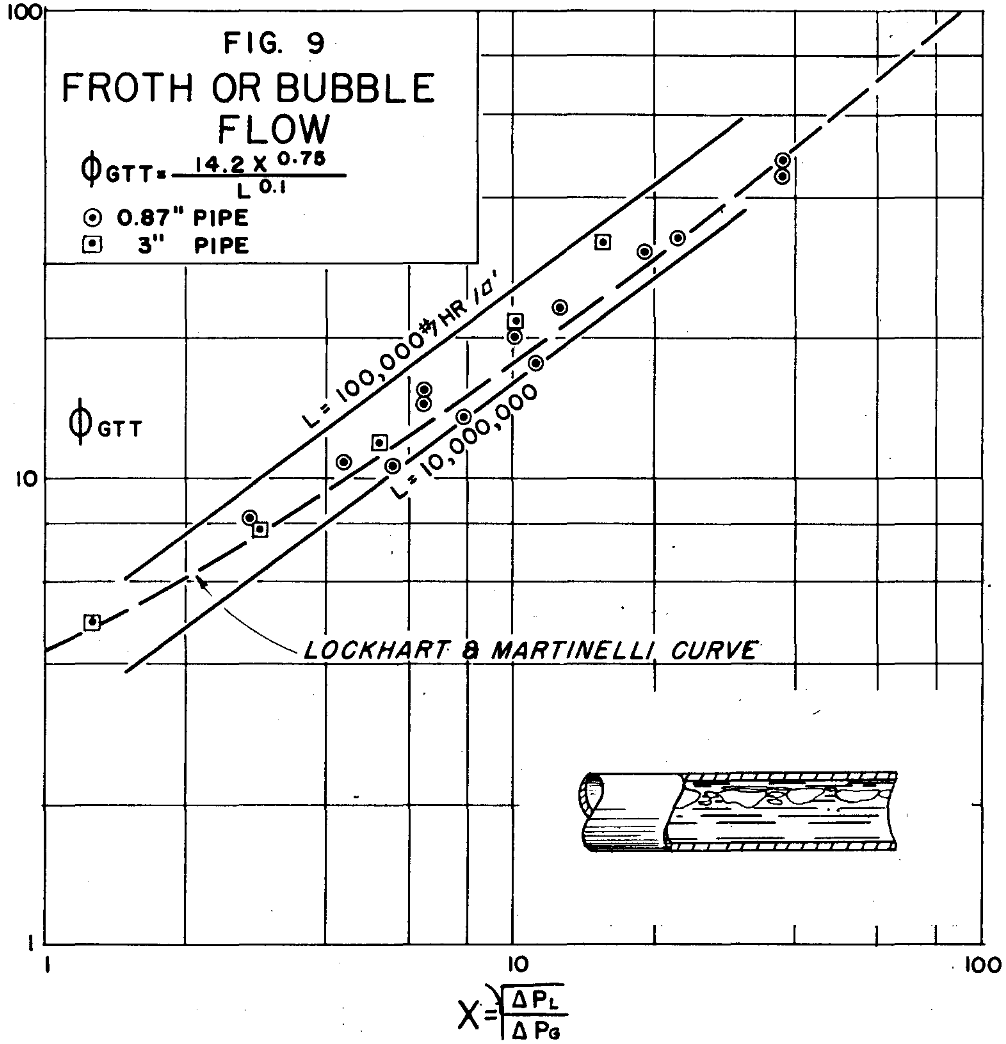


FIG. 9
FROTH OR BUBBLE FLOW
$\phi_{GTT} = \frac{14.2\,X^{0.75}}{L^{0.1}}$
0.87" PIPE
3" PIPE
L = 100,000 #/HR./10'
L = 10,000,000
LOCKHART & MARTINELLI CURVE
ϕ_{GTT}
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$
1
10
100

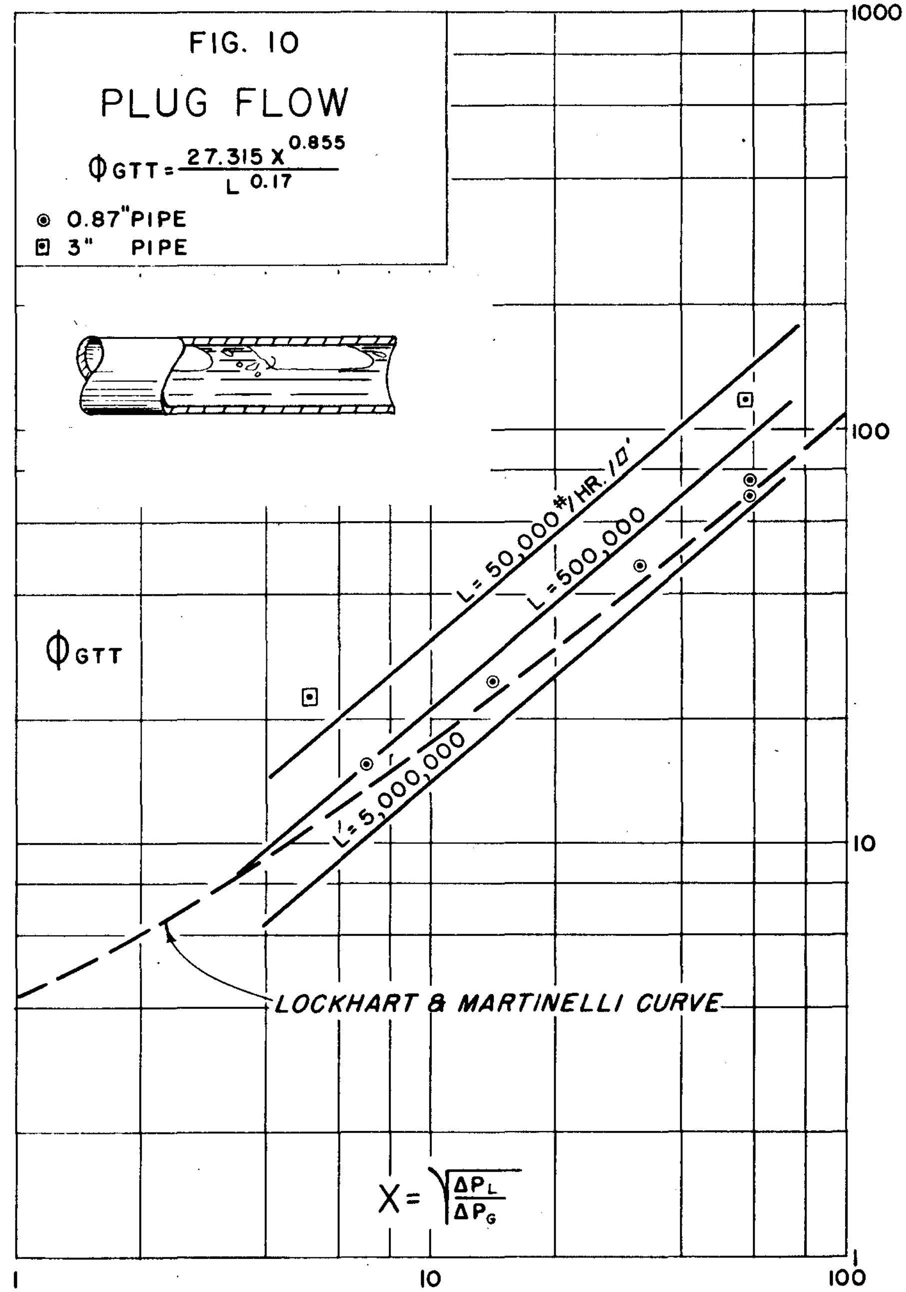


FIG. 10
PLUG FLOW
$\phi_{GTT} = \frac{27.315\,X^{0.855}}{L^{0.17}}$
0.87" PIPE
3" PIPE
L = 50,000 #/HR./10'
L = 500,000
L = 5,000,000
LOCKHART & MARTINELLI CURVE
ϕ_{GTT}
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$
1
10
100
1000

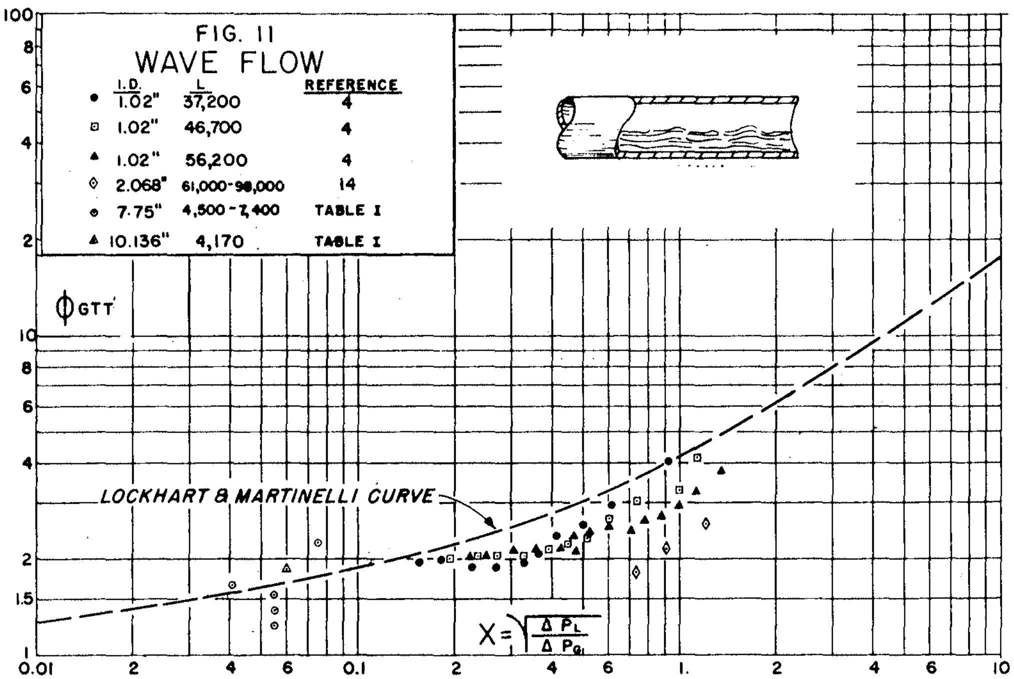


FIG. 11
WAVE FLOW
I.D. L REFERENCE
1.02" 37,200 4
1.02" 46,700 4
1.02" 56,200 4
2.068" 61,000-98,000 14
7.75" 4,500-7,400 TABLE I
10.136" 4,170 TABLE I
ϕ_{GTT}
LOCKHART & MARTINELLI CURVE
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$

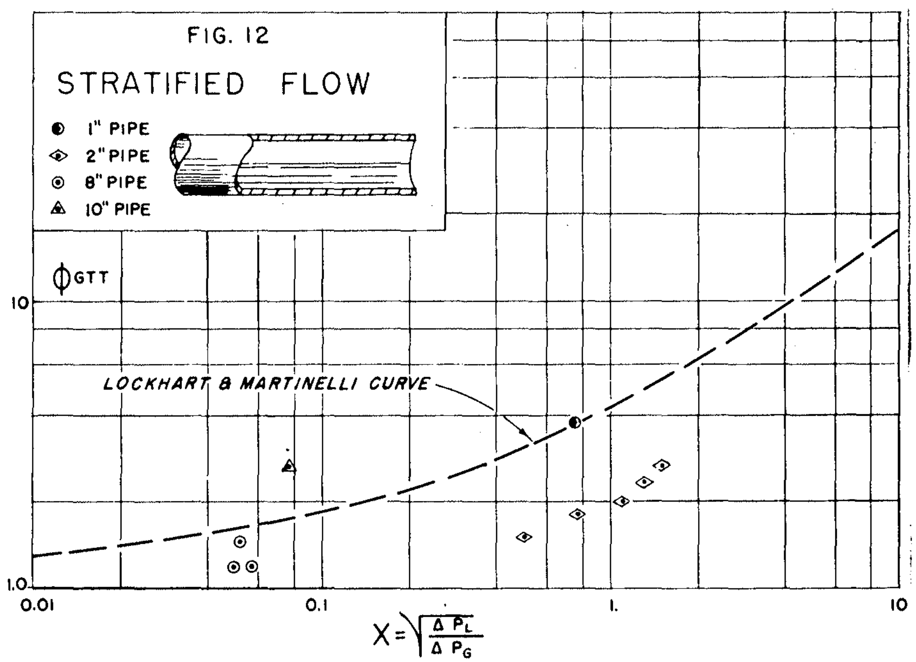


FIG. 12
STRATIFIED FLOW
1" PIPE
2" PIPE
8" PIPE
10" PIPE
ϕ_{GTT}
LOCKHART & MARTINELLI CURVE
$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}}$