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Gordon Albaugh Hughmark

Louisiana State University and Agricultural & Mechanical College

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HOLDUP AND PRESSURE DROP WITH GAS-LIQUID FLOW IN A VERTICAL PIPE

A Dissertation

Submitted to the Graduate Faculty of the
Louisiana State University and
Agricultural and Mechanical College
in partial fulfillment of the
requirements for the degree of
Doctor of Philosophy

in

The Department of Chemical Engineering

by

Gordon A. Hughmark

Ch. E., University of Cincinnati, 1947

M.S. (Ch. E.), Louisiana State University, 1955

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ABSTRACT

Vertical upward cocurrent air-liquid flow was investigated with a test section of one inch schedule 40 pipe. Pressure drop was measured with a mercury manometer connected to two pressure taps 20 ft apart in the section. The liquid under test was used as the seal fluid on the mercury. Two solenoid valves supplied air to two quick shut-off valves outside the pressure taps. Liquid drained from the section between the valves provided the holdup data. Six liquids were used in the investigation to determine the effect of liquid density, viscosity and surface tension.

The experimental holdup and pressure drop data could not be correlated with the Lockhart-Martinelli correlation for horizontal flow. Mass velocity was shown to be an important variable that is not considered in their correlation.

A correlation for holdup was developed to include physical properties of the fluids, total mass velocity and mass flows entering the pipe. Liquid volume fraction in the pipe was correlated with a function x .

$$x = \left(\frac{W_L}{W_G} \right)^{0.9} \frac{\mu_L^{0.19} \sigma^{0.205} \rho_G^{0.70} \mu_G^{2.75}}{G^{0.435} \rho_L^{0.72}}$$

W = mass rate entering pipe, lb/sec.

G = total mass velocity, lb/sq ft sec. When G exceeds 50, a value of 50 is used

μ = viscosity, centipoises

ρ = density, lb/cu ft

σ = surface tension, dynes/cm

sub L = liquid

sub G = gas

The correlation is in close agreement with the experimental and literature data. The results indicate that holdup is independent of pipe diameter in the range of 0.4 to 2.34 inches.

A pressure drop correlation was developed in which the two-phase pressure drop is a function of the difference in velocity between the two phases in the pipe with parameters of ψ .

$$\psi = \frac{1}{G^{0.70} \mu_L^{0.147} \sigma^{0.194}}$$

G = total mass velocity, lb/sq ft sec.

μ_L = liquid viscosity, centipoises

σ = surface tension, dynes/cm

Correlation with literature data for 0.5 and 2 inch pipe indicates that ψ is inversely proportional to the square root of the pipe diameter. An average percentage error of less than 15% was obtained with the correlation between the observed data for total pressure drop and the calculated pressure drop data.

INTRODUCTION

The principle of air-water two-phase flow was applied as early as 1797 to an air lift pump invented by a German, Carl Loscher. Two-phase flow was again applied in 1927 in the Seminole, Oklahoma, oil fields to produce oil by gas lift. Two-phase flow is of continuing interest in air lift pumps, gas-oil transmission in pipe lines, and in chemical process equipment. The operation of thermosiphon reboilers depends upon two-phase flow. Lack of a good correlation for two-phase flow has limited design of reboilers of this type to a "rule of thumb" basis.

A large amount of data and many papers have been published on horizontal and vertical cocurrent two-phase flow. Correlations have been published that can be used with reasonable agreement for horizontal flow but there is no satisfactory method for vertical flow. Almost all vertical flow data have been obtained with the air-water system. The purpose of this study is to obtain and correlate data on two-phase systems in vertical flow with emphasis on the study of liquids with different physical properties.

CHAPTER I

Review of Theory

Vertical cocurrent upward isothermal flow of a gas and liquid in a pipe extend from the condition of total liquid through the entire range of liquid-gas mixtures to the condition of total gas. If liquid and gas traveled at the same velocity in a pipe, two-phase flow analysis would be simple. The volume of gas in the pipe would be the same as the gas volume fed and the same with the liquid. This is not the situation. The gas phase, because of its lower density, travels at a higher velocity than the liquid phase. Thus a difference in velocity or slip velocity exists between the two phases. The difference in velocity results in an energy loss in addition to the friction loss at the wall of the pipe. Thus the friction loss for two-phase flow is usually greater than the friction loss for either fluid flowing alone.

Two-Phase Flow Regimes

Cromer⁷ in 1936 visually observed the flow of mixtures of air and water in a vertical pipe. He reported four types or regimes of flow.

1. Bubble flow exists when gas bubbles rise through a continuous liquid phase. This occurs at extremely low gas rates.
2. Mist flow occurs at extremely low liquid rates when droplets of liquid are suspended in a continuous gas phase.
3. Slug flow occurs when pistons of gas are interspersed with slugs of liquid. Neither phase is continuous and a pulsating flow exists.

4. Annular flow occurs at high mass velocities and low to moderate liquid rates. There is an annular flow of liquid along the pipe wall with a core of gas relatively free of liquid.

Both phases are continuous.

The transition between flow regimes is indistinct and it is sometimes difficult to determine the regime for a given condition.

A mathematical analysis of annular flow was attempted by Calvert.⁵ Theoretical and mathematical analysis of the entire range of two-phase flow is difficult if not impossible because of the indefinite flow regimes and because of the unsteady state condition of slug flow. Thus a semi-theoretical and empirical correlation is the general approach.

Horizontal Flow Correlations

Lockhart-Martinelli.

The work of Martinelli¹⁷ and co-workers provides one of the best correlations proposed for horizontal flow. Their basic postulates were stated in 1944 as follows:

1. The pressure drop for the liquid phase and the pressure drop for the gas phase must be equal regardless of the types of flow as long as appreciable pressure differences do not exist along any pipe diameter.
2. The volume occupied by the liquid and the volume occupied by the gas at any instant must equal the total volume of the pipe.

Lockhart and Martinelli¹⁵ in 1949 presented a correlation that enabled the prediction of pressure drops for horizontal flow to within plus or minus about thirty percent. The correlation for turbulent flow in both liquid and gas phases is of greatest interest. They correlated holdup, R_L , and pressure drop by means of a parameter X .

$$X = \left(\frac{W_L}{W_G} \right)^{0.9} \left(\frac{\rho_G}{\rho_L} \right)^{0.5} \left(\frac{\mu_L}{\mu_G} \right)^{0.1} = \frac{\left(\frac{\Delta P}{\Delta L} \right)_L}{\left(\frac{\Delta P}{\Delta L} \right)_G} \quad \text{--- (1)}$$

W_L = weight rate of liquid flow, lb/sec

W_G = weight rate of gas flow, lb/sec

ρ_G = gas density, lb/cu ft

ρ_L = liquid density, lb/cu ft

μ_L = liquid viscosity, lb/sec ft

μ_G = gas viscosity, lb/sec ft

$\left(\frac{\Delta P}{\Delta L} \right)_L$ = pressure drop per unit length if liquid phase flows alone

$\left(\frac{\Delta P}{\Delta L} \right)_G$ = pressure drop per unit length if gas phase flows alone

Holdup was plotted versus X . Pressure drop was contained in a function ϕ that was plotted vs. X .

$$\phi_L^2 = \frac{\left(\frac{\Delta P}{\Delta L} \right)_{TP}}{\left(\frac{\Delta P}{\Delta L} \right)_L}$$

$\left(\frac{\Delta P}{\Delta L} \right)_{TP}$ = pressure drop per unit length for two phase flow

The data included in the study by Lockhart and Martinelli are for horizontal flow in pipes from 0.62 inch to 1 inch in diameter and for pressures from 0 to 50 psig. A limited number of vertical runs were made in 0.0816 inch and 0.0586 inch tubes. The tube diameters are so small for vertical flow and the data are so limited that the correlation is considered only for horizontal flow.

Jenkins¹³ obtained two-phase pressure drop data from the flow of air and water at constant liquid rates. In attempting to correlate the data using Lockhart and Martinelli parameters, he found that a plot of X

versus ϕ showed distinct lines for each liquid rate crossing the curve of Lockhart and Martinelli. This indicated that the mass velocity is an important variable and is not included in the Lockhart-Martinelli correlation.

Chenoweth-Martin.

Chenoweth and Martin⁶ in 1955 published a correlation for horizontal turbulent flow. They plotted the ratio of $\frac{\Delta P_{TP}}{\Delta P_L^*}$ versus y_L with parameters of $\frac{\Delta P_G^*}{\Delta P_L^*} = \frac{f_G \rho_L}{f_L \rho_G}$

ΔP_{TP} = two phase pressure drop

ΔP_L^* = pressure drop for the liquid flowing at a mass velocity equal to that of the liquid plus the gas

ΔP_G^* = pressure drop for the gas flowing at a mass velocity equal to that of the liquid plus gas

y_L = liquid volume fraction entering the pipe

ρ_L = liquid density

ρ_G = gas density

f_G = friction factor for gas based on mass velocity of gas and liquid

f_L = friction factor for liquid based on mass velocity of liquid and gas

Experimental data were obtained in 1-1/2 inch and 3 inch pipes at pressures up to 100 psia. The correlation showed agreement within 20% for these data and the data covered by Lockhart and Martinelli.

Reid²⁰ and co-workers obtained horizontal flow data with the air-water system in 4 inch and 6 inch pipes. The 4 inch data are well represented by the Lockhart-Martinelli correlation but the 6 inch data fell about 20% low. The data showed excellent agreement with the Chenoweth-Martin correlation and showed no noticeable effects of pipe diameter.

Vertical Flow Data and Correlations

Schurig²¹ in 1934 published data on air-water in a 47 mm diameter tube as part of a study of two-phase flow in steam boilers.

Martinelli's success in 1944 in developing a correlation for predicting pressure drop for two-phase flow in horizontal tubes encouraged interest in the application of this correlation to vertical flow.

Kegel¹⁴ obtained data with air and water in a copper tube one inch in diameter and 11 feet long. The fluids were maintained in the turbulent region and the data were found to agree reasonably well with the Martinelli correlation.

Dukler⁸ and McElwee¹⁶ obtained vertical flow data which showed that the Martinelli correlation was satisfactory at high gas rates but useless at low air-water ratios. This observation was similar to that of Jenkins for horizontal flow in that a trend occurs with mass velocity which is not considered in the correlation.

Yagi²⁴ and co-workers correlated the liquid volume fraction, R_L , and the gas volume fraction, R_G , to the ratio of the apparent velocities of the liquid and gas. An excellent correlation is obtained over the limited range of two-phase flow investigated. The data were obtained on air-water and air-oil systems in a 10.25 mm diameter glass tube. Later data²⁵ with tube diameters of 8, 12.5 and 17.5 mm agreed with the original equation. Pressure drop data were also obtained and it was found that the pressure drop from the bottom to the top of the tube was linear with length.

Galegar,¹⁰ Stovall and Huntington presented a correlation for the pressure drop in vertical flow based on a plot of a dimensionless flow ratio versus a friction factor.

$$\text{flow ratio} = \frac{G_L \mu_L}{G_G \mu_G}$$

G = mass rate of gas flow referred to liquid or gas, lb/sq ft
sec

μ = viscosity, lb/ft hr

$$f = \frac{3600 \Delta P D^{0.5} g_c \rho_G G_L^2}{2 L G_G^2} \text{ --- (2)}$$

f = friction factor f

ΔP = total differential pressure across the test section

D = internal diameter of pipe, ft

g_c = 32.17 (lb force)(ft)/(lb mass)(sec²)

L = length of test section, ft

The correlation is based on data from test sections of smooth pipe 0.1667 ft diameter by 72.25 ft long and 0.04325 ft diameter by 19.14 ft long.

Holdup and pressure drop data for the air-water system are reported by Galegar⁹ and pressure drop data for air-kerosene are reported by Stovall²³. The correlation represents the data to within plus or minus 23 percent.

Schwarz²² presented holdup data for the steam-water system at pressures from 20 to 80 atmospheres. The data were obtained with a tube 59.5 mm diameter and 4.035 meters long.

Radford¹⁹ obtained holdup and pressure drop data on the air-water system with a smooth tube 1.049 inch diameter and 22.88 feet long. Data are reported for seven series of runs at constant water rates. Govier,¹¹ Radford and Dunn report holdup and pressure drop data for air-water with a smooth tube 1.025 inch diameter and 22.88 feet long. These data are also obtained as a series of runs at various constant water rates.

Isbin,¹² Sher and Eddy studied the steam-water system in a 0.872 inch tube at essentially atmospheric pressure. Holdup data are reported.

Fundamental Vertical Two-Phase Flow Concepts

The concept of vertical two-phase flow analysis is based on Martinelli's¹⁷ postulates that the pressure drop for the two phases are

equal and that the volume of liquid plus the volume of gas equals the total volume of the pipe. As shown by Figure 1, a known mass per unit time of liquid, W_L , and gas, W_G , enter the base of a pipe section at a pressure p_1 . The volumes per unit time at the inlet conditions are calculated. Let

$$y_L = \text{liquid volume fraction} = \frac{W_L v_L}{W_L v_L + W_G v_G}$$

v_L = specific volume of liquid, cu ft/lb

v_G = specific volume of gas, cu ft/lb

y_G = gas volume fraction, $1 - y_L$

Inside the section the gas travels at a higher velocity than the liquid. As a result the gas volume fraction in the pipe, R_G , is less than y_G and the liquid volume fraction in the pipe, $R_L = 1 - R_G$, is greater than y_L . Pressure, p_1 , is greater at the base of the pipe than the pressure, p_2 , at the top because of head and friction drop. Assuming isothermal flow, the gas volume corresponding to W_G is greater at the top than at the base. Since the liquid density is constant, the gas volume fraction, R_G , must be greater at p_2 than at p_1 . All experimental data indicate that R_G increases with decreasing y_L for a constant total mass velocity. The gas fraction, R_G , will then be greater at p_2 than at p_1 . This means that the fraction of the tube cross section area for gas flow increases upward through the pipe.

The data of Petrick¹⁸ indicate an increase in gas volume fraction upward through the pipe. Petrick studied the effect of sudden flow area changes on density. A radioactive technique was used to determine local density. Plots of the data indicate a gradual increase in gas volume fraction after recovery from the effect of sudden flow change.

FIGURE 1

Diagram of Vertical Cocurrent Two-Phase Flow

Energy Balance for Two-Phase Vertical Flow

A mechanical energy balance for the section shown by Figure 1 is made with the assumption that steady state exists. Then W_L pounds of liquid and W_G pounds of gas enter at 1 and leave at 2 in a given time interval. The mechanical energy balance form of the Bernoulli theorem⁴ is written as

$$\frac{g}{g_c} Z_1 + \frac{V_1^2}{2g_c} + W_O - \int_{p_1}^{p_2} v dp - F = \frac{g}{g_c} Z_2 + \frac{V_2^2}{2g_c} \quad \text{--- (3)}$$

Z = elevation, ft

V = average linear velocity of fluid flow, ft/sec

g = local acceleration of gravity, ft/sec²

$g_c = 32.17 \text{ (lb force)(ft)/(lb mass)(sec}^2\text{)}$

p = absolute pressure at elevation Z , lb force/sq ft

W_O = external work, ft-lb force/lb mass

v = specific volume of system fluid cu ft/lb mass

F = energy loss to friction, ft-lb force/lb mass

This equation applies to single phase flow; for liquid-gas flow the equation can be applied separately to each phase. Assume W_L pounds of liquid and W_G pounds of gas enter per unit time and $W_O = 0$.

For the liquid phase,

$$W_L \frac{g}{g_c} Z_1 + W_L \frac{V_{L1}^2}{2g_c} - W_L \int_{p_1}^{p_2} v_L dp - W_L F_L = W_L \frac{g}{g_c} Z_2 + W_L \frac{V_{L2}^2}{2g_c} \quad \text{--- (4a)}$$

For the gas phase,

$$W_G \frac{g}{g_c} Z_1 + W_G \frac{V_{G1}^2}{2g_c} - W_G \int_{p_1}^{p_2} v_G dp - W_G F_G = W_G \frac{g}{g_c} Z_2 + W_G \frac{V_{G2}^2}{2g_c} \quad \text{--- (4b)}$$

Since v_L can be assumed constant and v_G continually increases almost linearly as p decreases, $-\int_{p_1}^{p_2} v_L dp$ reduces to $-v_L(p_2 - p_1)$ and $-v_{G_{am}}(p_2 - p_1)$ is a good approximation of $-\int_{p_1}^{p_2} v_G dp$. Where

$v_{G_{am}}$ = arithmetic mean of v_{G_1} and v_{G_2} , cu ft/lb mass

Substituting and adding the equations for liquid (4a) and gas (4b) the following equation is obtained.

$$(W_L + W_G) \frac{g}{g_c} (Z_2 - Z_1) + \frac{W_L(V_{L_2}^2 - V_{L_1}^2)}{2g_c} + \frac{W_G(V_{G_2}^2 - V_{G_1}^2)}{2g_c} + W_L F_L + W_G F_G - (W_L v_L + W_G v_{G_{am}})(p_1 - p_2) = 0 \quad (5)$$

Assuming that $\frac{g}{g_c} = 1$ and dividing equation (5) through by $(W_L v_L + W_G v_{G_{am}})$,

$$\frac{(W_L + W_G)(Z_2 - Z_1)}{(W_L v_L + W_G v_{G_{am}})} + \frac{W_L(V_{L_2}^2 - V_{L_1}^2) + W_G(V_{G_2}^2 - V_{G_1}^2)}{2g_c(W_L v_L + W_G v_{G_{am}})} + \frac{W_L F_L + W_G F_G}{(W_L v_L + W_G v_{G_{am}})} - (p_1 - p_2) = 0 \quad (6)$$

Then defining

$$\Delta P_{TP} = \frac{W_L F_L + W_G F_G}{(W_L v_L + W_G v_{G_{am}})} \quad \text{where}$$

ΔP_{TP} = two phase pressure drop, lb force/sq ft

the following general equation is obtained for liquid-gas two-phase flow:

$$\frac{(W_L + W_G)(Z_2 - Z_1)}{(W_L v_L + W_G v_{G_{am}})} + \frac{W_L(V_{L_2}^2 - V_{L_1}^2) + W_G(V_{G_2}^2 - V_{G_1}^2)}{2g_c(W_L v_L + W_G v_{G_{am}})} + \Delta P_{TP} - (p_1 - p_2) = 0 \quad (7)$$

Equation (7) as derived in this Chapter will be applied later to correlate the new data reported herein for vertical liquid-gas two-phase flow.

CHAPTER II

Equipment, Method of Operation and Materials

Holdup and pressure drop data for cocurrent vertical air-liquid flow were obtained for a 20 foot section of one inch schedule 40 steel pipe. Flow rates studied ranged from 0.01 to 40 SCFM of air and from 0.05 to 7 gpm of liquid.

Equipment

The equipment is shown by Figure 2.

Air was drawn from the receiver of an air compressor which operated to deliver air between the limits of 70 and 95 psig. Air from the receiver passed through a Commercial Filters Corporation Fulflo Filter to remove entrained oil and water. The air was then regulated by a pressure reducer to supply 50 psig air to the rotameters. The most appropriate of three rotameters, connected in parallel, was used to give maximum accuracy of measurement in the range of air flow desired for a specific test. A valve downstream from each meter was used for control. The metered air then entered the base of the test section as shown by Figure 2.

Liquid was drawn from a 55 gallon drum by a rotary pump and discharged through either of two parallel rotameters. A by-pass from the pump discharge to the suction was used to circulate the liquid discharged by the positive displacement pump but which was not fed to the test section. Flow through each rotameter was controlled by a valve downstream from the meter. The metered liquid then entered the base of the test section.

FIGURE 2
Diagram of Experimental Equipment

The vertical test section itself was a 24.2 foot long run of one inch schedule 40 steel pipe with pressure taps 20 feet apart. The bottom pressure tap was 28 inches above the air-liquid mixing nozzle. Surge pots 3 inches in diameter by 6 inches long were installed with the tap from the test section entering the side of the pot. The outlet from the bottom of each pot was connected with copper tubing to a 30 inch differential manometer which used mercury as the metering element. The portion of the pots below the taps and the copper tubing were filled with the test liquid as a seal fluid in the mercury. The diameter of the surge pots was selected so that the cross sectional area of the pots was about 200 times the cross sectional area of the manometer. A change in manometer level thus reflected a negligible change in level in the surge pot.

One quick shut-off valve was installed below the bottom pressure tap and a second above the top pressure tap. Both valves were one inch Valvair Corporation two-way valves, normally open. Each valve was operated by a three way solenoid air-valve. Air at 50 psig was supplied to the solenoid valves. Both solenoids were operated from a single electrical connection. When this was closed, air was supplied to the air operated valves and caused immediate, simultaneous and tight shut off. The air operated valves were opened by breaking the electrical connection, the resultant action blocking the air at the solenoid valves and venting the air from the valves. To insure simultaneous and uniform closure, the solenoid valves were installed so that the air piping between the solenoid valve and its component air operated valve was identical for both.

A pressure gage was installed in the test section at the elevation of the bottom pressure tap. A drain connection was provided above the bottom valve and a vent was installed below the top valve. After the valves were closed, the vent valve was opened and the liquid in the section was drained through the drain connection.

Flow from the top of the test section passed to the holdup drum. Air separated from the liquid in the drum and was vented to the atmosphere through a coupling in the top of the drum. Liquid was recirculated from a connection in the bottom of the drum.

The air rotameters were calibrated by a timed air displacement of water in an inverted transparent container. Liquid rotameters were calibrated for each liquid by a timed weighing of the liquid. The zero points for the manometer were obtained for each liquid by obtaining a reading with the test section filled with liquid at zero flow and from a reading with liquid completely drained from the section. When the difference between the two readings, corrected for the seal liquid, corresponded to the calculated liquid head in the section, it was assumed that the leads were uniformly filled with the seal fluid.

Method of Operation

The desired liquid flow for a run was metered to the test section through use of the liquid rotameter. After the manometer had reached a steady reading, the desired gas flow was metered with the gas rotameter. Steady state of most of the two-phase runs was obtained in less than one minute. Runs at high liquid rates and very low air rates required several minutes to reach steady state conditions as shown by the manometer. The rotameter readings, manometer reading and pressure gage reading were recorded for each run. The valves were closed, the vent was opened and the liquid was drained from the test section to a two liter graduated cylinder. The volume of liquid drained from the test section was recorded.

Data for the runs are tabulated in Appendix A. Air flow is recorded as the CFM of air flowing through the test section at a pressure equal to the average of the pressures at the top and bottom taps. Liquid flow is in cubic feet per minute flowing through the section. Pressure as

tabulated is the reading of the pressure gage at the bottom of the section. Pressure drop is the actual pressure differential between the two taps. This was determined from the manometer reading.

The volume of the test section was determined to be 3570 cc. Complete drainage of water showed a holdup of 3550 cc above the drain connection. Calculated holdup between the bottom valve and the drain connection was 20 cc.

Two corrections were applied to the liquid volume drained from the section to obtain the total liquid holdup. These were:

1. Twenty cc was added to account for the liquid remaining in the section between the valve and drain connection.
2. Volume was added to account for the liquid adhering to the wall of the pipe after draining. This measurement was made for each liquid. The section was filled with the liquid and then drained. The difference between this volume and the known volume of 3550 cc for complete drainage was assumed to be the amount of liquid still adhering to the pipe wall after draining for each run with that liquid. These data are tabulated in Table I.

TABLE I

Volume Drained from Test Section Initially Full of Liquid

Liquid	Liquid Drained cc		Liquid Adhering to Pipe Wall cc
	Test 1	Test 2	
Water	3550	3550	0
Sodium Carbonate Solution	3520	3525	25
Varsol	3480	3480	70
Oil Blend No. 1	3480	3480	70
Oil Blend No. 2	3470	3470	80
Trichloroethylene	3550	3550	0

Air and liquid temperatures during the entire series of runs varied between 70° and 80°F. Air fed to the test section was saturated with water at the 70 psig air receiver pressure. Vaporization of liquid in the test section was considered to be negligible for all runs.

Materials

Runs were made with air and six different liquids. The liquids were selected to vary the physical properties of density, viscosity and surface tension. Because of the extremely turbulent nature of two-phase flow, dynamic rather than static surface tension would be expected to be an important variable. Therefore, use of surface active agents was purposely avoided in this investigation.

Table II lists the six liquids tested and gives their density, viscosity and surface tension.

TABLE II
Physical Properties of Liquids Used
in Two-Phase Flow Investigation

Material	Density Lb/Cu Ft	Viscosity Centipoises	Surface Tension Dynes/cm
Water	62.4 at 25°C	0.90 at 25°C	72.5 at 25°C
Sodium Carbonate Solution	7 ⁴ at 25°C	3.15 at 23.9°C	75.9 at 25°C
Varsol	48.5 at 25.5°C	1.07 at 25°C	24.9 at 25°C
Oil Blend No. 1	51.5 at 24°C	5.83 at 24.4°C	27.4 at 25°C
Oil Blend No. 2	54.2 at 25°C	28.6 at 25°C	29.9 at 25°C
Trichloroethylene	88.9 at 24°C	0.55 at 25°C	28.8 at 25°C

Density was determined by weighing one liter of the liquid in a one liter volumetric flask. The viscosity of water and trichloroethylene and the surface tension of water were obtained from data reported in the literature.^{3,4} The remaining viscosity and surface tension measurements were made on samples of the liquid by the Physical Data Group of the Ethyl Corporation. Viscosity was obtained with a Brookfield Viscosimeter with a special adapter for low viscosities. Surface tension was determined

with the Findlay surface tension apparatus using the capillary rise method.

Water runs were made with Baton Rouge city water. A solution of about 18 weight percent soda ash in water was used for the sodium carbonate solution runs. Varsol was used in a series of tests. Oil Blend No. 1 was prepared by adding 10 gallons of SAE 20 oil to about 10 gallons of Varsol. Six gallons of SAE 50 oil was added to about 10 gallons of Oil Blend No. 1 to obtain Oil Blend No. 2. Trichloroethylene for these runs was provided by the Ethyl Corporation.

CHAPTER III

Discussion and Correlation of Results

Single-phase pressure drop results for the test section are presented and discussed. The reliability of the two-phase data is evaluated. Observed effects of liquid physical properties are discussed. Hold-up data are compared with the Lockhart-Martinelli correlation and a new correlation for holdup is developed. Pressure drop data are compared with the Lockhart-Martinelli and Chenoweth-Martin correlations. A new pressure drop correlation is developed.

Single-Phase Pressure Drop

As a limiting case and as a means of comparing with single-phase flow, pressure drops were determined for the 20 ft test section with liquid flow and zero air flow. The results are tabulated in Table III.

TABLE III

Liquid Phase Pressure Drop

Liquid	Mass Velocity Lb/Sq Ft Sec.	Pressure Drop	Reynolds Number	Friction
		to Friction, In. Hg		Factor f
Water	150	0.57	20,300	0.031
Na ₂ CO ₃ Solution	150	0.60	6,200	0.039
Varsol	150	0.70	18,200	0.0303
Varsol	100	0.325	12,100	0.0314
Oil Blend No. 1	150	0.88	3,770	0.040
Oil Blend No. 2	150	2.0	680	0.090
Oil Blend No. 2	100	1.35	455	0.145
Oil Blend No. 2	50	0.68	225	0.293
Trichloroethylene	150	0.355	35,400	0.028

These data are shown as a typical friction factor plot in Figure 3. The solid line for the turbulent flow region represents the

friction factor for a relative roughness of 0.002 ,² corresponding to a one inch commercial steel pipe. Data for the laminar flow region are shown in comparison to the line for $f = \frac{64}{Re}$ which is Poiseuille's law in comparable units.

Reliability and Reproducibility of Experimental Data

Govier, Radford, and Dunn¹¹ present data for the air-water system at a water mass velocity of 128 lb/sq ft sec . Their test section was a 22.88 ft length of 1.025 inch smooth pipe. In the present study, a series of check runs were made at a constant water mass velocity of 128 lb/sq ft sec . to compare with this published data. The resulting holdup data are shown by Figure 4 and the pressure drop data by Figure 5. Very close agreement can be noted for the holdup data. The observed pressure drops are approximately 20% higher than those of Govier et al., as would be expected for data from a commercial pipe in comparison to a smooth pipe.

Reproducibility of the experimental data was ascertained by obtaining data from five runs with identical liquid and air feed rates. Four groups of five identical runs were made for several conditions of air-water flow and one group of five identical runs was made for each of the other liquids tested. The data are presented in Appendix B and a summary of the statistical analysis of the data in Table IV. In order to avoid bias, runs in the first seven groups were purposely made so that no two runs in the same group were consecutive. Data for several groups were taken on three different days. Groups 1, 2 and 3 show the reproducibility for the air-water system with three different values of y_L . Groups 1 and 4 compare total mass velocities of 25 and 50 for air-water with the same y_L . Groups 1, 5, 6, 7, 8, and 9 show the effect of the different liquids. Ninety percent of the data would be expected to fall within plus or minus two standard deviations of the mean for a "t" distribution. For R_L data,

TABLE IV
Analysis of Data for Reproducibility

Group	Run Number	Liquid	G	y_L	Mean R_L	Standard Deviation R_L	Mean ΔP , In. Hg	Standard Deviation ΔP , In. Hg
1	248, 251, 256, 259, 262	Water	25	0.378	0.5854	0.0055	9.89	0.055
2	249, 255, 258, 260, 266	Water	25	0.053	0.2574	0.0087	4.34	0.23
3	245, 250, 257, 261, 268	Water	25	0.0042	0.08082	0.0125	3.67	0.09
4	269, 275, 281, 288, 296	Water	50	0.378	0.5050	0.0111	8.66	0.13
5	346, 405, 408, 424, 430	Na ₂ CO ₃ Solution	25	0.378	0.631	0.0207	12.06	0.44
6	588, 591, 610, 613, 616	Varsol	25	0.378	0.5256	0.0111	7.07	0.097
7	805, 808, 811, 814, 817	Oil Blend No. 1	25	0.378	0.5657	0.0086	7.82	0.083
8	957-961	Oil Blend No. 2	25	0.363	0.7258	0.0088	10.08	0.067
9	1025-1029	Trichloroethylene	37.5	0.378	0.5232	0.0056	18.24	0.043

two standard deviations vary from a minimum of 0.011 for Group 1 to a maximum of 0.041 for Group 5. For pressure drop, two standard deviations vary from 10.6% of the mean for Group 2 to 0.47% for Group 9. There does not appear to be any significant effect of liquid characteristics, total mass velocity or y_L on the reproducibility of the data.

The standard deviations were obtained from

$$\sigma_o^2 = \frac{\sum X^2}{(N-1)} - \frac{(\sum X)^2}{N(N-1)} \quad \text{--- (8)}$$

σ_o = standard deviation

$\sum X^2$ = sum of squares of R_L or ΔP for a group

$\sum X$ = sum of R_L or ΔP for a group

N = number of runs in the group.

Holdup As A Function of Liquid Physical Properties and Flow Rate

The object for the experimental program was as follows:

1. To obtain holdup and pressure drop data for a range of y_L at a constant total mass velocity with a given liquid.
2. To repeat 1 at several total mass velocities with the same liquid.
3. To repeat 1 and 2 for each liquid tested.

Mass velocity was considered as an important variable because Jenkins'¹³ data showed a parallel trend from the Lockhart-Martinelli correlation with liquid mass velocity as a parameter. Review of vertical data in the literature^{10,11} showed the same trend with mass velocity. However, total mass velocity was found to give a more parallel trend than liquid mass velocity alone. For this reason, total mass velocity was used in obtaining the experimental data.

Figure 6 shows holdup, R_G , as a function of y_L at a total mass velocity of 150 lb/sq ft sec. for all of the liquids tested. The dotted

line indicates R_G for no difference in velocity between the phases, which is the limiting case in which $R_G = y_G = 1 - y_L$. The observed data approach this condition as y_L approaches 1. As shown by Figure 6 the liquids appear to give the following order of increasing difference between R_G and y_G for a constant value of y_L . The properties of these liquids, as measured, are included to explain their behavior.

	<u>ρ</u>	<u>μ</u>	<u>σ</u>
1. Trichloroethylene	88.9	0.55	28.8
2. Varsol	48.5	1.07	24.9
3. Oil Blend No. 1	51.5	5.83	27.4
4. Water	62.4	0.9	72.5
5. Sodium Carbonate Solution	74.0	3.15	75.9
6. Oil Blend No. 2	54.2	28.6	29.9

Varsol, Oil Blend No. 1 and Oil Blend No. 2 have approximately the same density and surface tension. From a comparison of these, it can then be concluded that the difference between R_G and y_G increases with liquid viscosity. In fact, except for water and sodium carbonate solution, consideration of the tabulation shows an increase in this difference with increasing liquid viscosity. The fact that water and sodium carbonate solution show a greater difference between R_G and y_G that is inconsistent in terms of viscosity, also indicates that another physical property must contribute. Both these liquids have high surface tensions, which suggests that the difference between R_G and y_G must increase with increasing surface tension. The effect of liquid density is not obvious.

Figure 7 shows the holdup for water at three different total mass velocities. It is apparent that the difference between R_G and y_G increases as the total mass velocity decreases.

FIGURE 7
Holdup Data for Air-Water at
Various Total Mass Velocities

Figures 6 and 7 represent only a small part of the experimental data. However, the observations for other flow rates are consistent with Figure 6 and those for other fluids are consistent with Figure 7. In summary, the difference between R_G and y_G is indicated to behave qualitatively as follows:

1. $R_G = y_G$ at $y_L = 1$. As y_L decreases, the difference between R_G and y_G increases to a maximum and then decreases to 0 at $y_L = 0$.
2. Since R_G is always less than y_G , the gas velocity is always greater than the liquid velocity.
3. At constant y_L , the difference between R_G and y_G tends to be greater at low total mass velocities than at high total mass velocities.
4. At constant y_L and constant total mass velocity, the difference between R_G and y_G appears to increase with increasing liquid viscosity and with increasing surface tension.

Holdup Data Correlation

Figure 8 shows experimental data for the air-water system at total mass velocities of 5, 25, and 150 lb/sq ft sec. presented in the form used by Lockhart and Martinelli.¹⁵ From the relative consistency of the data for each flow rate and their position relative to the best line drawn for the Lockhart-Martinelli data, it is apparent that mass velocity is an important variable that is not included in their correlation.

The experimental data appear to be substantially parallel to the best line Lockhart-Martinelli correlation and the deviations appear to be a function of total mass velocity. The correlating factor used by Lockhart and Martinelli and shown by equation (1) was

$$X = \left(\frac{W_L}{W_G} \right)^{0.9} \left(\frac{\rho_G}{\rho_L} \right)^{0.5} \left(\frac{\mu_L}{\mu_G} \right)^{0.1}$$

The viscosity term is constant for each liquid tested in this investigation. Therefore, $\left(\frac{W_L}{W_G} \right)^{0.9} \left(\frac{\rho_G}{\rho_L} \right)^{0.5}$ was used as the initial correlating factor. This was plotted versus R_L with a separate plot for each liquid. Figure 9 shows the plot for the air-water data. The plots for the other liquids were similar. Data for total mass velocities of 50, 100 and 150 were found to fall together. The function $\frac{1}{G^{0.435}}$ was found to correlate the parameters for total mass velocities of 5, 10, 25 and 50. A separate plot was then made for each liquid of R_L versus $\frac{1}{G^{0.435}} \left(\frac{W_L}{W_G} \right)^{0.9} \left(\frac{\rho_G}{\rho_L} \right)^{0.5}$. When G was greater than 50, a value of 50 was used. A best line was drawn for each plot and these lines were then drawn on a single plot shown by Figure 10.

The parameters are essentially parallel for the liquids. At $R_L = .60$, the value of $\frac{1}{G^{0.435}} \left(\frac{W_L}{W_G} \right)^{0.9} \left(\frac{\rho_G}{\rho_L} \right)^{0.5}$ was obtained for each liquid. This was divided by $\left(\frac{\rho_G}{\rho_L} \right)^{0.5}$ at the average conditions of the runs to eliminate density from the terms and thus obtain the best value of $\frac{1}{G^{0.435}} \left(\frac{W_L}{W_G} \right)^{0.9}$ for each liquid. This value was correlated by the equation:

$$\log \left[\frac{1}{G^{0.435}} \left(\frac{W_L}{W_G} \right)^{0.9} \right] = \log a_0 + a_1 \log \rho_L + a_2 \log \mu_L + a_3 \log \sigma \quad \text{-----} \quad (9)$$

A least squares calculation was made with the LGP-30 computer at the Ethyl Corporation to obtain the best fit to the equation. The correlation coefficient for the least squares fit was 0.98 which indicates a high level of significance of the fit of the equation to the data. The data obtained are shown by Table V.

TABLE V

Summary of Coefficients and Standard Deviations
for Holdup Correlation - Experimental Data

<u>Physical Property</u>	<u>Coefficient</u>	<u>Standard Deviation</u>
ρ_L	-0.70989	0.265
μ_L	+0.18971	0.040
σ	+0.20448	0.106

The coefficients are twice the standard deviation or higher. All three coefficients are therefore at the 95% confidence level or better.

Correlation of the experimental data did not consider either density or viscosity of the gas phase because these did not vary significantly. The data of Schwarz²² report holdup data on water-steam at 20 to 80 atmospheres pressure. These data were worked up in the same manner as the experimental data and were pooled with the experimental data for correlation to obtain coefficients for ρ_G and μ_G . The results are shown by Table VI.

TABLE VI

Summary of Coefficients and Standard
Deviations for Holdup Correlation -
Schwarz Data and Experimental Data

<u>Physical Property</u>	<u>Coefficient</u>	<u>Standard Deviation</u>
ρ_L	-0.7205	0.22
μ_L	+0.18773	0.032
σ	+0.2054	0.09
ρ_G	+0.70266	0.10
μ_G	+2.7539	0.56

The coefficients for the liquid physical properties are essentially the same as for the correlation of the experimental data alone. All coefficients are above the 98% confidence level. The correlation coefficient was 0.99.

The correlating function for R_L is thus found to be

$$\left(\frac{W_L}{WG}\right)^{0.9} \frac{\mu_L^{0.19} \sigma^{0.205} \rho_G^{0.70} \mu_G^{2.75}}{G^{0.435} \rho_L^{0.72}}$$

where G is a maximum of 50 lb/sq ft sec. The experimental data were correlated using this function but neglecting μ_G which was constant at 0.018 centipoises. The calculated function x and R_L are tabulated in Appendix C for all of the experimental data. The correlated data are shown by Figure 11. Gas viscosity is then added to the function by the factor

$$\left(\frac{\mu_G}{0.018}\right)^{2.75} = 6 \times 10^4 \mu_G^{2.75}$$

The final correlation shown by Figure 12 plots R_L against the function x .

$$x = 6 \times 10^4 \left(\frac{W_L}{WG}\right)^{0.9} \frac{\mu_L^{0.19} \sigma^{0.205} \rho_G^{0.70} \mu_G^{2.75}}{G^{0.435} \rho_L^{0.72}} \quad (10)$$

W = weight rate of flow, lb/sec.

μ = viscosity, centipoises

ρ = density, lb/cu ft

G = total mass velocity, lb/sq ft sec. All values greater than 50 are assumed to equal 50.

σ = surface tension, dynes/cm.

This proposed correlation was compared with the experimental data by determining R_L (observed) - R_L (calculated). The correlation was also compared in the same way with published holdup data. The results are tabulated by Table VII.

Excellent agreement is obtained with the proposed correlation over a wide range of fluid physical properties and for pipe diameters from 0.4 to 2.34 inches in diameter.

TABLE VII
Summary of $R_L(\text{Observed}) - R_L(\text{Calculated})$
for Proposed Holdup Correlation

Investigator	System	Pipe Diameter Inches	No. of Runs	Average ΔR_L	Range of ΔR_L
This investigation	Air-Water	1.049	176	0.0221	-0.11 to +0.038
This investigation	Air- Na_2CO_3 Solution	1.049	108	0.0228	-0.074 to +0.065
This investigation	Air-Varsol	1.049	97	0.0225	-0.067 to +0.052
This investigation	Air-Oil Blend No. 1	1.049	90	0.021	-0.057 to +0.06
This investigation	Air-Oil Blend No. 2	1.049	55	0.0239	-0.033 to +0.064
This investigation	Air-Trichloroethylene	1.049	25	0.0218	-0.038 to +0.053
Govier et al. ¹¹	Air-Water	1.025	26	0.024	-0.058 to +0.023
Yagi et al. ²⁴	Air-Water	0.405	24	0.014	-0.04 to +0.028
Yagi et al. ²⁵	Air-Water	0.69	16	0.012	-0.03 to +0.02
Schurig ²¹	Air-Water	1.85	29	0.0455	-0.105 to +0.032
Isbin et al. ¹²	Steam-Water	0.872	23	0.073	-0.18 to +0.02
Schwarz ²²	Steam-Water	2.34	15	0.025	-0.078 to +0.025
Galegar et al. ¹⁰	Air-Water	2.0	32	0.0445	-0.14 to +0.027
Galegar et al. ¹⁰	Air-Water	0.52	15	0.045	-0.075 to -0.015

Two-Phase Pressure Drop

The term $\frac{W_L(V_{L2}^2 - V_{L1}^2) + W_G(V_{G2}^2 - V_{G1}^2)}{2g_c(W_L V_L + W_G V_{Gam})}$ of equation (7) was found

to be negligible for the experimental data. Equation (7) then reduces to

$$\frac{(Z_2 - Z_1)(W_L + W_G)}{(W_L V_L + W_G V_{Gam})} - (p_1 - p_2) + \Delta P_{TP} = 0 \quad \text{--- (11)}$$

Equation (11) was used to calculate ΔP_{TP} from the experimental data. The first term, the pressure change due to potential head, was calculated based on the air and liquid rates entering the pipe. The differential pressure ($p_1 - p_2$) was that measured between the two pressure taps.

Figure 13 and Figure 14 show the results, using the methods proposed by Lockhart and Martinelli¹⁵ and Chenoweth and Martin⁶ to correlate the experimental data in the turbulent range for the air-water system. A definite trend with mass velocity is shown in each case.

The two-phase pressure drop, ΔP_{TP} , is due to two mechanisms both of which cause energy losses.

1. Friction between the fluid and the pipe wall.
2. Turbulence between the two phases.

The latter of these could be expected to be a function of the difference in velocity between the two phases which is defined as the slip velocity. Figure 15 shows the two-phase pressure drop per foot of pipe versus slip velocity for the air-water runs. Similar plots were obtained with the data for the other fluids studied. Calculated slip velocity and two-phase pressure drop per unit length for the experimental runs are tabulated in Appendix C. Figure 16 shows $\Delta P_{TP}/L$ versus slip velocity for all systems for tests made at a total mass velocity of 150 lb/sq ft sec.

The ordinate intercept is the pressure drop for 100% liquid flow. To force all series to pass through zero $\Delta P_{TP}/L$ at zero slip velocity, the ordinate was revised to $\frac{\Delta P_{TP} - \Delta P_L}{L}$ where ΔP_L is the pressure drop for liquid flow at the total mass velocity of the specific series of

runs. The data were replotted and at a slip velocity of 40 ft/sec. a value of $\frac{\Delta P_{TP} - \Delta P_L}{L}$ was obtained for each system at each total mass velocity. These values were correlated by the equation:

$$\log \frac{\Delta P_{TP} - \Delta P_L}{L} = \log a_0 + a_1 \log \rho_L + a_2 \log \mu_L + a_3 \log \sigma + a_4 \log G \quad \text{--- (12)}$$

The computer was used to obtain the best least squares fit in the same manner as for the holdup correlation. A correlation coefficient of 0.984 was obtained with $a_1 = 0$. This indicates that the correlation is independent of liquid density. The data are shown by Table VIII.

TABLE VIII

Summary of Coefficients and Standard Deviations for Pressure Drop Correlation

<u>Physical Property</u>	<u>Coefficient</u>	<u>Standard Deviation</u>
μ_L	-0.1476	0.019
σ	-0.1944	0.057
G	-0.6973	0.033

The coefficients are at least three times the standard deviation which indicates that each is very significant. The parameter to correlate physical properties and total mass velocity is therefore

$$\psi = \frac{1}{\mu_L^{0.149} \sigma^{0.194} G^{0.70}} \quad \text{--- (13)}$$

Plots were then made of $\frac{\Delta P_{TP} - \Delta P_L}{L}$ versus ψ at intervals of slip velocity of 20 ft/sec. Figure 17 shows a typical plot for this step. Best values of $\frac{\Delta P_{TP} - \Delta P_L}{L}$ were then obtained and plotted with ψ as parameters. The final graph is shown by Figure 18.

Total liquid-gas pressure drop can then be calculated with equation (11). The two-phase pressure drop, ΔP_{TP} , is calculated by adding ΔP_L to $\Delta P_{TP} - \Delta P_L$ obtained from the correlation for a specific fluid and total mass velocity which fixes the value of ψ . The proposed correlation procedure was checked with the experimental data by determining the

percentage error between the observed and calculated total pressure drop. The results are tabulated in Table IX.

TABLE IX
Comparison of Observed and Calculated
Total Pressure Drop for Experimental Runs

<u>System</u>	<u>No. of Runs</u>	<u>Average % Absolute Error</u>
Air-Water	182	14.2
Air-Na ₂ CO ₃ Solution	113	10.5
Air-Varsol	96	14.3
Air-Oil Blend No. 1	90	9.2
Air-Oil Blend No. 2	57	6.4
Air-Trichloroethylene	25	8.4

Average Absolute Error of 563 Runs = 11.4%

This correlation is based on a 1.049 inch pipe diameter with a relative roughness of 0.002. Published data are for smooth pipes of different diameters and were used to extend the correlation.

Galegar et al.¹⁰ presented pressure drop data for smooth pipes 2 inches and 0.52 inches in diameter. Govier et al.¹¹ presented data for a 1.025 inch diameter smooth pipe. These data were correlated with a diameter correction to ψ .

$$\psi = \frac{1}{G^{0.7} \mu_L^{0.147} \sigma^{0.194}} \left(\frac{D}{0.0873} \right)^{0.5} \quad \text{--- (14)}$$

where D = inside diameter of pipe, ft.

A correction for pipe roughness was made with a correction to equation (11) to give the following equation for total pressure drop.

$$(P_1 - P_2) = \frac{(Z_2 - Z_1) (W_L + W_G)}{(W_L V_L + W_G V_{G_{am}})} + \frac{f_L}{(f_L)_{0.002}} \Delta P_{TP} \quad \text{--- (15)}$$

f_L = friction factor for liquid at total mass velocity
in the pipe

$(f_L)_{0.002}$ = friction factor for liquid at total mass velocity
in pipe with relative roughness of 0.002.

The absolute average percentage error between the observed and calculated data are shown by Table X.

TABLE X
Comparison of Observed and Calculated
Total Pressure Drop for Literature Data

<u>Investigator</u>	<u>System</u>	<u>Pipe Diameter, Inches</u>	<u>No. of Runs</u>	<u>Average % Error</u>
Govier et al. ¹¹	Air-Water	1.025	27	10.4
Galegar et al. ¹⁰	Air-Water	2.0	24	14.5
Galegar et al. ¹⁰	Air-Water	0.52	15	8.3

Thus the correlation for total pressure drop in vertical two-phase flow appears to be accurate to within 15% for the range of data investigated.

SUMMARY AND CONCLUSIONS

Vertical upward cocurrent two-phase flow data were obtained with a one inch pipe for air and six different liquids. The holdup and pressure drop data obtained showed significant deviations from the correlations proposed by Lockhart and Martinelli for horizontal two-phase flow. Therefore, correlations were developed for holdup and pressure drop based on the experimental data.

Gas and liquid physical properties, mass velocity of flow, and mass flows entering the pipe were found to be the important variables for holdup in the pipe. Liquid volume fraction in the pipe was correlated with a function x .

$$x = \left(\frac{W_L}{W_G} \right)^{0.9} \frac{\mu_L^{0.19} \sigma^{0.205} \rho_G^{0.70} \mu_G^{2.75}}{G^{0.435} \rho_L^{0.72}}$$

W = mass rate entering pipe, lb/sec.

G = total mass velocity, lb/sq ft sec. When G exceeds 50, a value of 50 is used.

μ = viscosity, centipoises

ρ = density, lb/cu ft

σ = surface tension, dynes/cm

sub L = liquid

sub G = gas

The correlation is in close agreement with the experimental data and literature data for holdup. The literature data indicate that holdup is independent of pipe diameter in the range of the available data, 0.4 to 2.34 inches.

The same variables that effect holdup were found to be important in the correlation of two-phase pressure drop data. This correlation shows two-phase pressure drop to be a function of the difference in velocity between the two phases in the pipe with parameters of ψ .

$$\psi = \frac{1}{G^{0.70} \mu_L^{0.147} \sigma^{0.194}}$$

G = total mass velocity, lb/sq ft sec.

μ_L = liquid viscosity, centipoises

σ = surface tension, dynes/cm

Correlation with literature data for 0.5 and 2 inch pipe indicates that ψ is inversely proportional to the square root of the pipe diameter.

The procedure for calculating the total pressure drop for vertical cocurrent upward flow is based on the following equation developed from the Bernoulli theorem.

$$\frac{(Z_2 - Z_1) (W_L + W_G)}{(W_L v_L + W_G v_{G_{am}})} - (P_1 - P_2) + \Delta P_{TP} = 0$$

W = mass rate entering pipe, lb/sec.

Z = elevation, ft

v = specific volume, cu ft/lb

p = absolute pressure, lb/sq ft

ΔP_{TP} = two-phase pressure drop, lb/sq ft

sub L, G = liquid, gas

sub G_{am} = arithmetic mean of gas at 1 and 2

sub 1,2 = inlet and outlet respectively

The procedure is as follows:

1. Calculate the pressure change due to potential head, the first term of the general equation.
2. Calculate the liquid volume fraction in the pipe with the holdup correlation.

3. Calculate the liquid and gas velocities in the pipe from the mass flows and 2.
4. Calculate ΔP_{TP} from the difference in velocity between the two-phases from 3 and for ψ for the system.
5. With the head term and ΔP_{TP} thus obtained, calculate $(p_1 - p_2)$ from the general equation.

Recommendations for extending this work are as follows:

1. The holdup correlation is based upon only two gases: air and steam. Additional data should be obtained with other gases to better determine the effect of gas physical properties.
2. The pressure drop correlation is based on air as the gas with experimental data for pressures near atmospheric. The correlation could be extended with work on other gases to determine the effect of gas physical properties and at higher pressures to determine the effect of pressure.

NOMENCLATURE

a_0	constant
a_1, a_2, a_3, a_4	exponents
D	inside diameter of pipe, ft
F	energy loss to friction, ft-lb force/lb-mass
f	friction factor
G	mass velocity, lb mass/sq ft sec.
g_c	32.17 (lb force)(ft)/(lb mass)(sec. ²)
L	length of pipe, ft
N	number of runs in a group
ΔP	pressure drop, lb force/sq ft
p	absolute pressure, lb force/sq ft
R	volume fraction in pipe
V	linear velocity, ft/sec.
v	specific volume, cu ft/lb mass
W	mass rate of flow, lb mass/sec.
W_0	external work, ft-lb force/lb-mass
X	parameter in Lockhart-Martinelli correlation
x	parameter in proposed holdup correlation
y	volume fraction entering the pipe
Z	elevation, ft
Δ	finite change of variable
μ	viscosity, lb mass/ft-hr or centipoises as noted
ρ	density, lb mass/cu ft

ΣX	sum of R_L or ΔP for a group
ΣX^2	sum of squares of R_L or ΔP for a group
σ	surface tension, dynes/cm
σ_0	standard deviation
ψ	parameter in proposed pressure drop correlation
sub G	gas
sub G_{am}	gas-arithmetic mean
sub L	liquid
sub S	slip
sub TP	two-phase
sub 1	point of inlet to section
sub 2	point of outlet of section

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APPENDIX A

Holdup and Pressure Drop Data

<u>Run No.</u>	<u>Air Rate</u> CFM	<u>Water Rate</u> CFM	<u>Holdup</u> cc	<u>Pressure</u> <u>Gage, psig</u>	<u>Pressure</u> <u>Drop, In. Hg</u>
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Air-Water Runs

G = 5

135	3.02	0.0243	800	2.8	3.8
136	4.17	0.0227	725	2.7	3.6
137	5.33	0.0208	620	2.5	3.2
138	6.45	0.0195	580	2.4	3.0
139	7.63	0.0179	500	2.3	2.7
140	0.00095	0.0289	3490	10.0	17.6
141	0.0195	0.0289	3185	9.4	16.2
142	0.0473	0.0289	2910	8.4	14.5
143	0.110	0.0289	2440	7.0	11.45
144	0.187	0.0289	2110	6.2	9.7
145	0.268	0.0289	1820	5.3	8.1
146	0.353	0.0289	1650	4.5	6.8
147	0.355	0.0285	1620	4.6	7.4
148	0.485	0.0281	1325	3.7	5.5
149	0.74	0.0278	1055	3.1	4.25
150	1.19	0.0272	875	2.6	3.35
151	1.70	0.0265	810	2.5	3.2
152	1.98	0.0262	815	2.6	3.4
153	2.28	0.0256	800	2.7	3.6
194	4.73	0.0219	660	2.5	3.1
195	7.45	0.0187	505	2.3	2.9
196	10.45	0.0156	350	2.0	2.0
197	13.4	0.011	210	2.0	1.5
198	19.9	0.0070	125	1.9	1.1
199	23.2	0.0	0	-	0.9

G = 10

54	0.625	0.0575	1095	3.6	4.9
55	0.852	0.0575	1045	3.0	4.1
56	1.22	0.0575	890	3.7	3.6
57	1.69	0.0575	870	3.0	3.45
58	2.05	0.0575	815	3.5	3.45
59	2.58	0.0575	805	2.7	3.7
60	3.38	0.0575	780	3.0	4.2
61	4.2	0.0575	790	3.1	4.2

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate CFM</u>	<u>Water Rate CFM</u>	<u>Holdup cc</u>	<u>Pressure Gage, psig</u>	<u>Pressure Drop, In. Hg</u>
<u>Air-Water Runs (Continued)</u>					
78	0.00265	0.0575	3495	10.0	17.7
79	0.0101	0.0575	3340	9.7	16.9
80	0.0288	0.0575	3150	9.2	16.4
81	0.047	0.0575	2990	8.7	15.0
82	0.109	0.0575	2610	7.4	12.6
83	0.186	0.0575	2240	6.5	10.5
84	0.264	0.0575	2015	5.7	9.0
85	0.343	0.0575	1730	5.1	7.9
187	4.75	0.051	710	2.8	3.7
188	7.4	0.0475	550	2.5	3.2
189	10.5	0.0445	400	2.1	2.7
190	13.6	0.040	300	2.0	2.1
191	16.5	0.036	220	2.0	1.9
192	20.0	0.0306	150	2.0	1.7
193	23.6	0.0242	120	2.1	1.7
224	0.413	0.057	1460	4.5	6.5
225	0.60	0.0565	1235	3.7	5.1
226	0.90	0.0565	1010	3.0	4.1
227	1.20	0.056	930	2.8	3.9
228	1.52	0.0555	810	2.7	3.6
229	1.79	0.054	820	2.8	3.5
230	2.24	0.054	810	2.8	3.7
278	26.4	0.020	105	2.5	1.3
279	30.3	0.0144	80	2.4	1.2
280	35.5	0.0078	50	2.5	1.2

G = 25

33	0.823	0.144	1225	3.8	5.5
34	1.66	0.144	930	3.3	4.4
35	2.62	0.144	970	3.2	4.6
36	3.30	0.144	890	3.5	5.1
37	4.12	0.144	840	3.5	4.9
86	0.0053	0.144	3490	10.0	17.7
87	0.0193	0.144	3325	9.6	16.8
88	0.0465	0.144	3110	9.0	15.6
89	0.1065	0.144	2760	8.1	13.8
90	0.183	0.144	2450	7.1	12.0
91	0.26	0.144	2145	6.5	10.6
92	0.336	0.144	1960	6.0	9.4
158	0.405	0.144	1820	5.5	8.9
180	4.55	0.138	810	3.5	4.8
181	7.3	0.134	710	3.1	4.4
182	10.1	0.131	575	3.0	4.0
183	13.0	0.127	490	3.1	3.9

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate CFM</u>	<u>Water Rate CFM</u>	<u>Holdup cc</u>	<u>Pressure Gage, psig</u>	<u>Pressure Drop, In. Hg</u>
<u>Air-Water Runs (Continued)</u>					
184	16.0	0.122	395	3.0	3.8
185	19.2	0.118	335	3.1	3.7
186	22.7	0.111	290	3.2	3.7
218	0.595	0.144	1530	4.3	6.9
219	0.87	0.144	1200	3.8	5.5
220	1.18	0.144	1105	3.5	5.0
221	1.50	0.144	1000	3.2	4.5
222	1.83	0.144	940	3.0	4.6
223	2.23	0.144	900	3.4	4.6
245	25.6	0.1075	260	3.6	3.7
246	29.1	0.102	215	3.7	3.7
247	33.9	0.0935	180	4.0	3.9

G = 50

27	0.791	0.289	1440	5.0	7.3
28	1.61	0.289	1180	4.2	5.7
29	2.54	0.289	1070	4.0	5.7
30	3.14	0.289	1010	4.4	5.9
31	3.97	0.289	975	4.4	6.0
47	0.796	0.289	1460	5.0	7.3
99	0.0054	0.289	3485	10.0	17.55
100	0.0191	0.289	3360	9.9	17.1
101	0.046	0.289	3195	9.5	16.3
102	0.105	0.289	2930	8.7	14.7
103	0.178	0.289	2620	8.0	13.3
104	0.25	0.289	2450	7.4	12.2
105	0.334	0.289	2270	6.7	11.0
120	2.88	0.289	1060	4.2	6.3
121	3.95	0.289	1000	4.0	6.0
122	5.21	0.289	910	4.0	6.0
123	6.35	0.289	830	4.0	6.0
124	7.43	0.289	835	4.0	5.8
157	0.39	0.289	2145	6.5	10.3
173	4.4	0.28	865	4.5	6.3
174	7.05	0.278	810	4.1	6.0
175	9.8	0.275	680	4.0	5.7
176	12.25	0.270	605	4.1	5.7
177	15.15	0.267	550	4.5	5.8
178	18.2	0.261	520	4.8	6.2
179	21.3	0.254	465	4.8	5.7
212	0.573	0.289	1775	5.5	8.8
213	0.845	0.289	1420	4.8	7.3
214	1.14	0.289	1265	4.6	6.7
215	1.43	0.289	1260	4.2	6.1

Holdup and Pressure Drop Data
(Continued)

Run No.	Air Rate CFM	Water Rate CFM	Holdup cc	Pressure Gage, psig	Pressure Drop, In. Hg
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Air-Water Runs (Continued)

216	1.73	0.289	1090	4.2	6.1
217	2.31	0.289	1090	4.0	5.8
252	23.8	0.251	430	5.4	6.0
253	26.8	0.247	390	5.8	6.3
254	31.2	0.238	375	6.2	6.6

G = 100

93	0.010	0.58	3430	10.2	17.75
94	0.046	0.58	3320	10.0	17.2
95	0.104	0.58	3115	9.5	16.0
96	0.176	0.58	2950	9.0	15.2
97	0.25	0.58	2795	8.6	14.2
98	0.323	0.58	2615	8.1	13.4
115	2.58	0.58	1230	6.8	8.5
116	3.66	0.58	1090	6.0	8.6
117	4.63	0.58	1035	6.0	8.6
118	5.63	0.58	950	6.0	8.6
119	6.63	0.58	905	6.5	8.8
156	0.37	0.58	2510	8.0	12.8
166	4.2	0.57	1095	6.0	8.8
167	6.45	0.57	940	6.5	9.3
168	8.95	0.565	835	6.5	9.3
169	11.3	0.56	770	7.0	9.7
170	13.8	0.555	700	7.2	9.9
171	16.3	0.55	660	8.0	10.4
172	18.8	0.543	630	8.5	10.6
206	0.56	0.58	2225	7.2	11.25
207	0.80	0.58	1870	6.6	10.1
208	1.07	0.58	1660	6.4	9.5
209	1.36	0.58	1560	6.0	8.9
210	1.65	0.58	1440	5.7	8.5
211	2.21	0.58	1275	5.5	8.6
263	21.5	0.541	550	9.0	11.0
264	23.8	0.537	530	9.5	11.3
265	26.9	0.528	505	10.5	11.5

G = 150

40	0.355	0.87	2730	9.0	14.5
41	0.720	0.87	2300	8.1	12.5
42	1.46	0.87	1725	7.3	10.7
43	2.16	0.87	1510	7.5	10.2
44	2.83	0.87	1340	7.2	10.2
62	0.54	0.87	2450	8.5	13.3

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate CFM</u>	<u>Water Rate CFM</u>	<u>Holdup cc</u>	<u>Pressure Gage, psig</u>	<u>Pressure Drop, In. Hg</u>
<u>Air-Water Runs (Continued)</u>					
63	1.11	0.87	1905	7.5	11.3
64	1.82	0.87	1605	7.3	10.3
72	0.01	0.87	3450	10.4	18.1
73	0.045	0.87	3390	10.4	17.7
74	0.10	0.87	3240	10.1	17.0
75	0.167	0.87	3110	9.9	16.3
76	0.235	0.87	2950	9.5	15.6
77	0.303	0.87	2840	9.2	14.9
106	2.57	0.87	1480	7.5	10.5
107	3.47	0.87	1330	7.7	10.6
108	4.3	0.87	1170	8.1	10.8
109	5.33	0.87	1100	8.0	11.4
110	5.95	0.87	1095	8.5	11.6
155	0.36	0.87	2730	9.0	14.4
159	3.9	0.855	1165	8.0	11.2
160	6.05	0.855	1040	8.5	12.1
161	8.15	0.85	950	9.2	12.4
162	10.2	0.845	870	10.0	13.0
163	12.2	0.841	820	10.8	13.9
164	14.5	0.838	760	11.5	14.4
165	16.6	0.833	690	12.5	15.3
200	0.525	0.87	2485	8.3	13.0
201	0.76	0.87	2230	8.0	12.2
202	1.03	0.87	2050	7.7	11.5
203	1.29	0.87	1840	7.5	11.8
204	1.55	0.87	1710	7.3	10.7
205	2.08	0.87	1500	7.0	10.3
272	18.8	0.83	680	13.0	15.8
273	21.1	0.827	625	13.5	16.5
274	24.2	0.82	580	14.5	17.7

Holdup and Pressure Drop Data
(Continued)

Run No.	<u>Air Rate</u> CFM	<u>Liquid Rate</u> CFM	<u>Holdup</u> cc	<u>Pressure</u> Gage, psig	<u>Pressure</u> Drop, In. Hg
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Air-Sodium Carbonate Solution Runs

G = 5

371	0.0096	0.0244	3380	11.4	20.1
372	0.044	0.0244	2970	10.0	17.4
373	0.130	0.0244	2120	7.3	11.1
374	0.210	0.0241	1685	5.8	8.8
375	0.325	0.0240	1350	4.8	7.1
376	0.413	0.0237	1305	4.3	6.1
377	0.735	0.0234	1075	3.5	4.7
378	1.19	0.0229	910	3.1	3.8
379	1.65	0.0223	875	2.9	3.6
380	1.98	0.0218	800	2.8	3.6
381	2.25	0.0210	775	2.9	3.4
382	3.5	0.0202	730	3.0	3.6
383	5.98	0.0171	585	2.6	3.3
384	10.4	0.0124	415	2.2	2.3
402	13.5	0.0065	325	2.3	2.0
403	16.4	0.004	220	2.0	1.5
404	18.0	0.003	180	2.0	1.1

G = 10

357	0.0099	0.0485	3400	11.4	20.9
358	0.044	0.0485	3000	10.0	17.8
359	0.132	0.0485	2235	7.8	12.6
360	0.205	0.0483	1925	6.5	11.0
361	0.315	0.0482	1540	5.4	7.75
362	0.405	0.0481	1470	4.8	7.25
363	0.725	0.0479	1090	3.7	5.0
364	1.175	0.0473	970	3.3	4.3
365	1.64	0.0468	850	3.0	4.0
366	1.96	0.0463	890	3.0	3.9
367	2.24	0.0460	870	3.0	3.8
368	3.46	0.0445	780	3.2	4.2
369	5.85	0.0417	645	3.0	3.75
370	10.1	0.037	470	2.4	2.9
395	10.1	0.037	485	2.6	3.0
396	13.5	0.0333	380	2.4	2.4
397	16.4	0.0297	295	2.1	2.1
398	19.3	0.026	240	2.5	2.1
399	23.5	0.0213	200	2.4	1.8
400	26.6	0.0175	165	2.4	1.9
401	33.0	0.0087	130	2.9	1.9

Holdup and Pressure Drop Data
(Continued)

Run No.	Air Rate CFM	Liquid Rate CFM	Holdup cc	Pressure Gage, psig	Pressure Drop, In. Hg
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Air-Sodium Carbonate Solution Runs (Continued)

G = 25

343	0.0096	0.122	3425	11.5	21.5
344	0.0435	0.122	3120	10.5	18.5
345	0.130	0.122	2550	8.9	14.7
346	0.201	0.122	2200	7.6	12.6
347	0.305	0.122	1905	6.6	10.4
348	0.39	0.122	1745	6.1	9.2
349	0.69	0.122	1370	5.0	6.9
350	1.15	0.122	1200	4.1	5.6
351	1.60	0.122	1025	3.8	5.3
352	1.94	0.122	990	3.6	5.0
353	2.18	0.120	925	3.7	4.85
354	3.45	0.116	770	3.5	5.4
355	5.8	0.113	750	3.5	4.8
356	9.95	0.109	560	3.2	4.1
391	12.8	0.107	490	3.1	3.9
392	15.6	0.103	410	3.4	3.6
393	18.7	0.099	350	3.5	3.9
394	22.6	0.0941	310	3.4	3.6
417	25.6	0.0925	265	4.9	3.8
418	31.0	0.0865	215	4.3	4.1

G = 50

329	0.0096	0.243	3450	11.7	20.7
330	0.043	0.243	3225	11.0	19.2
331	0.124	0.243	2800	9.6	16.7
332	0.195	0.243	2500	8.7	15.0
333	0.296	0.243	2210	7.7	12.3
334	0.38	0.243	2080	7.1	11.5
335	0.673	0.243	1680	6.1	9.0
336	1.07	0.243	1330	5.2	7.3
337	1.52	0.243	1155	4.8	7.2
338	1.86	0.243	1055	4.6	6.6
339	2.12	0.243	1040	4.5	6.5
340	3.36	0.237	975	4.4	6.9
341	5.6	0.234	765	4.5	6.8
342	9.43	0.231	690	4.5	5.8
389	14.7	0.225	515	5.0	5.9
390	21.0	0.213	440	5.3	6.2
415	25.4	0.21	380	5.7	6.6
416	28.9	0.204	320	6.6	6.9

Holdup and Pressure Drop Data
(Continued)

Run No.	Air Rate CFM	Liquid Rate CFM	Holdup cc	Pressure Gage, psig	Pressure Drop, In. Hg
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Air-Sodium Carbonate Solution Runs (Continued)

G = 100

315	0.0096	0.485		12.2	21.0
316	0.043	0.485	3340	11.7	20.1
317	0.122	0.485	3025	10.6	18.3
318	0.185	0.485	2820	10.1	16.9
319	0.280	0.485	2570	9.4	15.2
320	0.357	0.485	2460	9.0	14.4
321	0.66	0.485	2060	7.7	11.9
322	1.05	0.485	1650	6.8	11.5
323	1.46	0.485	1495	6.5	9.6
324	1.76	0.485	1410	6.3	9.4
325	1.99	0.485	1270	6.3	9.0
326	2.15	0.477	1270	6.2	9.0
327	5.30	0.472	930	6.0	9.3
328	8.9	0.470	720	6.5	9.6
387	13.5	0.466	610	7.5	9.5
388	19.2	0.458	570	8.0	10.4
413	22.5	0.452	505	9.0	10.4
414	25.9	0.445	475	10.0	11.5

G = 150

300	0.0094	0.73	3460	12.2	21.35
301	0.0416	0.73	3410	12.0	20.9
302	0.094	0.73	3300	11.9	19.9
303	0.159	0.73	3075	11.3	18.9
304	0.207	0.73	2960	11.0	18.2
305	0.273	0.73	2790	10.5	17.4
306	0.345	0.73	2700	10.2	16.7
307	0.610	0.73	2340	9.2	14.75
308	0.994	0.73	2025	8.5	13.1
309	1.39	0.73	1735	8.0	12.0
310	1.66	0.73	1610	7.8	11.65
311	1.90	0.73	1510	7.7	11.4
312	2.06	0.72	1505	7.5	11.3
313	4.95	0.717	1070	8.2	12.1
314	8.1	0.714	925	9.2	13.1
385	12.8	0.711	765	9.6	14.1
386	17.1	0.695	625	11.5	14.6
411	20.3	0.695	605	12.5	15.2
412	23.2	0.690	530	13.5	15.6

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate</u> CFM	<u>Varsol Rate</u> CFM	<u>Holdup</u> cc	<u>Pressure</u> Gage, psig	<u>Pressure</u> Drop, In. Hg
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Air-Varsol Runs

G = 5

594	0.0107	0.037	3270	7.9	13.0
595	0.0483	0.037	2780	6.6	10.8
596	0.146	0.037	1840	4.4	6.7
597	0.23	0.037	1450	3.6	5.1
598	0.346	0.0365	1170	3.0	3.9
599	0.441	0.0363	970	2.6	3.2
600	0.773	0.0357	935	2.3	2.6
601	1.24	0.0351	810	2.2	2.5
602	1.68	0.0344	910	2.4	2.6
603	2.0	0.0336	825	2.3	2.6
604	2.31	0.0334	840	2.3	2.7
605	3.6	0.0312	720	2.3	2.6
606	6.15	0.0271	515	2.0	2.0
607	10.5	0.0198	280	1.6	1.1
608	15.1	0.0111	140	1.5	0.7
609	15.9	0.0097	130	1.5	0.7

G = 10

570	0.0108	0.0737	3260	7.9	13.2
571	0.0487	0.0737	2900	6.9	11.3
572	0.143	0.0737	2020	5.0	7.7
573	0.225	0.0737	1550	4.1	6.1
574	0.342	0.0737	1310	3.3	4.6
575	0.435	0.073	1180	3.0	4.1
576	0.77	0.0722	1020	2.5	3.0
577	1.22	0.0715	920	2.5	2.8
578	1.68	0.0708	950	2.5	2.9
579	2.0	0.070	930	2.5	3.1
580	2.29	0.0695	910	2.6	3.2
581	3.58	0.0675	770	2.5	2.8
582	6.15	0.062	580	2.1	2.3
583	10.3	0.0555	380	2.0	1.7
584	14.8	0.0485	230	1.9	1.4
585	19.7	0.040	160	2.1	1.3
586	27.0	0.027	100	2.2	1.3
587	36.0	0.010	50	2.3	1.3

G = 25

516	0.0107	0.184	3425	8.0	13.5
517	0.049	0.184	3150	7.5	12.4
518	0.139	0.184	2485	6.1	9.9

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate</u> <u>CFM</u>	<u>Varsol Rate</u> <u>CFM</u>	<u>Holdup</u> <u>cc</u>	<u>Pressure</u> <u>Gage, psig</u>	<u>Pressure</u> <u>Drop, In. Hg</u>
<u>Air-Varsol Runs (Continued)</u>					
519	0.209	0.184	2000	5.3	8.3
520	0.330	0.184	1730	4.4	6.5
521	0.42	0.184	1550	4.1	6.1
522	0.745	0.184	1170	3.3	4.4
523	1.14	0.184	1020	2.7	3.7
524	1.67	0.184	910	2.7	3.5
525	1.98	0.184	960	2.7	3.7
526	2.27	0.184	880	2.8	3.8
527	2.4	0.18	855	2.8	3.8
528	6.0	0.174	630	2.7	3.0
529	10.1	0.166	450	2.5	2.7
530	14.7	0.158	360	2.6	2.7
531	19.4	0.149	320	2.9	2.8
532	25.6	0.137	220	3.2	2.9
533	33.4	0.120	160	4.0	3.1

G = 50

534	0.0485	0.368	3280	7.9	13.1
535	0.135	0.368	2820	7.0	11.4
536	0.209	0.368	2490	6.5	10.2
537	0.318	0.368	2110	5.6	8.8
538	0.40	0.368	1960	5.3	8.0
539	0.725	0.368	1510	4.3	6.3
540	1.10	0.368	1230	3.7	5.2
541	1.62	0.368	1140	3.6	4.7
542	1.91	0.368	1050	3.5	4.6
543	2.21	0.368	1010	3.5	4.7
544	3.47	0.362	860	3.3	4.6
545	5.83	0.359	690	3.3	4.5
546	9.8	0.351	530	3.5	4.3
547	13.8	0.343	440	4.0	4.4
548	18.5	0.333	380	4.2	4.7
549	24.1	0.32	310	5.0	5.0
550	30.8	0.304	290	6.1	5.3

G = 100

500	0.132	0.737	3070	8.0	12.6
501	0.204	0.737	2870	7.5	12.0
502	0.305	0.737	2610	7.0	11.0
503	0.387	0.737	2470	6.7	10.3
504	0.685	0.737	2220	6.5	9.0
505	1.03	0.737	1740	5.5	7.7
506	1.54	0.737	1520	5.1	7.2

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate CFM</u>	<u>Varsol Rate CFM</u>	<u>Holdup cc</u>	<u>Pressure Gage, psig</u>	<u>Pressure Drop, In. Hg</u>
<u>Air-Varsol Runs (Continued)</u>					
507	1.82	0.737	1385	5.0	6.8
508	2.10	0.737	1310	5.0	6.8
509	2.21	0.732	1260	4.9	6.5
510	5.47	0.725	830	5.3	6.8
511	8.1	0.719	670	6.3	7.1
512	12.5	0.703	590	6.8	7.9
513	16.2	0.695	510	7.6	8.3
514	21.1	0.69	460	8.6	8.7
515	27.0	0.675	390	10.0	9.6

G = 150

554	0.129	1.11	3045	8.3	13.1
555	0.20	1.11	2960	7.9	12.9
556	0.296	1.11	2790	7.8	12.2
557	0.38	1.11	2665	7.6	12.0
558	0.657	1.11	2290	7.1	10.5
559	0.98	1.11	1990	6.9	9.5
560	1.43	1.11	1740	6.7	8.8
561	1.69	1.11	1660	6.7	8.6
562	1.96	1.11	1580	6.6	8.4
563	3.06	1.1	1300	6.8	8.1
564	5.98	1.1	930	8.0	9.1
565	8.91	1.09	750	9.5	10.1
566	13.3	1.08	620	10.5	10.8

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate</u> CFM	<u>Oil Rate</u> CFM	<u>Holdup</u> cc	<u>Pressure</u> Gage, psig	<u>Pressure</u> Drop, In. Hg
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Air-Oil Blend No. 1 Runs

G = 5

785	0.0107	0.035	3330	8.1	13.8
786	0.0493	0.035	2750	7.0	11.2
787	0.143	0.035	1970	4.9	7.4
788	0.225	0.035	1630	4.0	5.9
789	0.338	0.035	1420	3.5	4.8
790	0.427	0.034	1370	3.4	4.5
791	0.76	0.0335	1190	2.9	3.8
792	1.14	0.0328	1060	2.5	3.1
793	1.68	0.032	910	2.4	2.9
794	2.3	0.0312	880	2.3	2.9
795	3.6	0.029	740	2.0	2.7
796	7.55	0.0227	530	2.0	2.1
797	11.75	0.0154	330	2.0	1.4
798	14.8	0.0105	260	2.0	1.4
799	18.2	0.0049	180	2.0	1.2
800	20.0	0.0021	110	1.9	1.0

G = 10

766	0.0106	0.070	3360	8.3	14.0
767	0.048	0.070	2910	7.3	12.0
768	0.140	0.070	2140	5.5	8.5
769	0.22	0.070	1810	4.6	6.9
770	0.334	0.070	1570	3.9	5.5
771	0.425	0.069	1340	3.7	5.1
772	0.75	0.0685	1240	3.2	4.2
773	1.14	0.068	1100	2.7	3.75
774	1.67	0.067	980	2.6	3.2
775	2.27	0.0662	870	2.6	3.2
776	3.57	0.064	710	2.5	3.1
777	7.4	0.0575	600	2.4	2.7
778	13.4	0.048	410	2.0	2.0
779	19.7	0.037	280	2.3	2.0
780	26.7	0.0238	170	2.5	2.1
781	34.9	0.0094	80	3.0	2.3

G = 25

715	0.049	0.175	3140	7.9	13.2
716	0.137	0.175	2620	6.7	10.8
717	0.216	0.175	2290	5.8	9.3
718	0.327	0.175	2010	5.0	7.9
719	0.417	0.175	1820	4.8	7.2

Holdup and Pressure Drop Data
(Continued)

Run No.	Air Rate CFM	Oil Rate CFM	Holdup cc	Pressure Gage, psig	Pressure Drop, In. Hg
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Air-Oil Blend No. 1 Runs (Continued)

720	0.737	0.175	1530	4.0	5.9
721	1.12	0.175	1340	3.5	5.0
722	1.65	0.175	1135	3.2	4.5
723	2.24	0.175	1090	3.2	4.2
724	3.5	0.170	960	3.2	4.0
725	7.33	0.162	670	2.9	3.7
726	12.8	0.155	525	3.2	3.5
727	18.8	0.142	440	3.5	3.5
728	25.2	0.130	330	3.9	3.6
729	33.0	0.114	270	4.5	4.1

G = 50

732	0.0473	0.35	3250	8.4	13.7
733	0.133	0.35	2940	7.5	12.4
734	0.207	0.35	2700	7.0	11.8
735	0.308	0.35	2390	6.5	10.0
736	0.395	0.35	2260	6.0	9.3
737	0.708	0.35	1830	5.0	7.5
738	1.07	0.35	1570	4.5	6.4
739	1.59	0.35	1390	4.0	5.9
740	2.17	0.35	1170	4.0	5.5
741	3.41	0.344	940	3.8	5.3
742	7.04	0.338	750	4.0	5.4
743	12.2	0.328	540	4.5	5.7
744	17.9	0.317	440	5.0	5.9
745	23.7	0.305	380	5.5	5.8
746	30.6	0.289	350	6.5	6.2

G = 100

700	0.0466	0.70	3200	8.9	14.6
701	0.129	0.70	3050	8.5	13.7
702	0.199	0.70	2945	8.1	13.1
703	0.297	0.70	2715	7.7	12.1
704	0.377	0.70	2580	7.5	11.6
705	0.67	0.70	2220	6.8	10.3
706	1.0	0.70	1920	6.3	9.0
707	1.49	0.70	1640	5.9	8.4
708	2.04	0.70	1410	5.7	7.8
709	3.17	0.692	1140	5.7	7.6
710	6.4	0.686	890	6.5	8.6
711	10.9	0.67	690	7.5	9.4
712	15.7	0.66	600	8.5	9.6
713	20.3	0.655	510	9.7	10.1
714	26.3	0.64	465	10.7	10.6

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate CFM</u>	<u>Oil Rate CFM</u>	<u>Holdup cc</u>	<u>Pressure Gage, psig</u>	<u>Pressure Drop, In. Hg</u>
<u>Air-Oil Blend No. 1 Runs (Continued)</u>					
G = 150					
750	0.125	1.05	3150	9.1	14.3
751	0.195	1.05	3080	8.8	14.0
752	0.29	1.05	2930	8.6	13.5
753	0.366	1.05	2860	8.5	13.2
754	0.64	1.05	2510	8.0	12.1
755	1.03	1.05	2230	7.6	11.0
756	1.41	1.05	1960	7.5	10.4
757	1.92	1.05	1740	7.3	9.9
758	2.96	1.045	1360	7.6	9.7
759	5.87	1.035	970	8.8	10.8
760	9.85	1.025	790	10.6	12.7
761	13.7	1.02	660	12.5	13.4
762	18.2	1.005	590	13.5	14.7

Holdup and Pressure Drop Data
(Continued)

Run No.	Air Rate CFM	Oil Rate CFM	Holdup cc	Pressure Gage, psig	Pressure Drop, In. Hg
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Air-Oil Blend No. 2 Runs

G = 25

942	0.0475	0.167	3160	8.4	13.9
943	0.135	0.167	2780	7.0	11.7
944	0.209	0.167	2570	6.5	10.4
945	0.314	0.167	2300	5.8	9.2
946	0.401	0.167	2170	5.5	8.5
947	0.713	0.167	1900	4.8	7.2
948	1.15	0.167	1630	4.3	6.2
949	1.60	0.167	1450	3.9	5.7
950	2.19	0.167	1280	3.8	5.6
951	3.43	0.161	1100	3.7	5.4
952	7.0	0.156	930	4.0	5.6
953	12.3	0.145	790	4.4	5.8
954	18.1	0.135	600	4.8	6.1
955	24.4	0.124	490	5.0	6.3
956	31.3	0.109	360	6.1	6.9

G = 50

927	0.047	0.333	3270	8.9	15.0
928	0.130	0.333	3000	8.1	13.5
929	0.204	0.333	2800	7.5	12.5
930	0.306	0.333	2570	6.9	10.9
931	0.387	0.333	2430	6.7	10.7
932	0.685	0.333	2140	6.0	9.3
933	1.105	0.333	1810	5.5	8.3
934	1.55	0.333	1630	5.0	7.6
935	2.1	0.333	1440	5.0	7.4
936	3.28	0.327	1230	5.0	7.3
937	6.68	0.321	930	5.3	7.7
938	11.25	0.312	840	6.5	8.7
939	16.6	0.302	730	7.5	10.0
940	22.6	0.29	680	7.5	10.4
941	28.4	0.276	620	9.0	10.9

G = 100

912	0.046	0.663	3340	9.5	16.25
913	0.125	0.663	3180	9.3	15.5
914	0.194	0.663	3040	9.0	14.9
915	0.291	0.663	2870	8.5	14.05
916	0.367	0.663	2740	8.4	13.6
917	0.65	0.663	2450	7.9	12.5
918	1.04	0.663	2230	7.5	11.6

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate CFM</u>	<u>Oil Rate CFM</u>	<u>Holdup cc</u>	<u>Pressure Gage, psig</u>	<u>Pressure Drop, In. Hg</u>
<u>Air-Oil Blend No. 2 Runs (Continued)</u>					
919	1.45	0.663	1990	7.2	11.0
920	1.95	0.663	1660	7.2	10.6
921	3.05	0.656	1330	7.2	10.8
922	6.15	0.651	1060	8.1	12.1
923	10.5	0.641	890	9.5	14.2
924	14.8	0.630	690	11.0	15.0
925	19.4	0.619	640	12.0	15.6
926	25.7	0.605	650	13.5	16.8

G = 150

900	0.123	0.995	3270	10.1	16.7
901	0.188	0.995	3190	10.0	16.4
902	0.28	0.995	3020	9.8	15.9
903	0.355	0.995	2970	9.6	15.5
904	0.62	0.995	2690	9.3	14.8
905	0.99	0.995	2470	9.1	14.0
906	1.36	0.995	2190	8.9	13.5
907	1.86	0.995	1970	8.9	13.3
908	2.82	0.99	1650	9.5	13.3
909	5.6	0.985	1210	10.7	15.0
910	9.1	0.975	1040	13.5	17.3
911	13.5	0.965	890	14.0	18.4

Holdup and Pressure Drop Data
(Continued)

<u>Run No.</u>	<u>Air Rate</u> CFM	<u>Liquid Rate</u> CFM	<u>Holdup</u> cc	<u>Pressure</u> <u>Gage, psig</u>	<u>Pressure</u> <u>Drop, In. Hg</u>
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Air-Trichloroethylene Runs

G = 37.5

1013	0.125	0.153	2470	10.0	17.9
1014	0.197	0.153	2020	8.7	14.5
1015	0.30	0.153	1720	7.2	11.5
1016	0.39	0.153	1540	6.7	10.4
1017	0.705	0.153	1180	5.3	7.9
1018	1.13	0.153	1100	4.8	6.5
1019	1.57	0.153	930	4.8	6.8
1020	2.12	0.153	850	4.8	6.7
1021	3.34	0.147	750	4.5	6.2
1022	6.95	0.142	540	4.0	5.1
1023	12.8	0.133	380	3.5	4.0
1024	18.4	0.125	260	4.0	3.8

G = 150

1000	0.114	0.61	3150	13.4	23.3
1001	0.178	0.61	2900	12.6	21.3
1002	0.271	0.61	2610	11.5	19.5
1003	0.349	0.61	2430	10.8	18.6
1004	0.619	0.61	2000	9.5	15.3
1005	1.02	0.61	1600	8.3	12.8
1006	1.42	0.61	1430	8.0	11.9
1007	1.92	0.61	1240	7.8	11.5
1008	3.05	0.605	1010	7.5	11.2
1009	5.95	0.60	770	8.5	11.5
1010	10.3	0.595	570	9.5	11.7
1011	14.5	0.590	470	10.7	11.9
1012	19.0	0.585	450	12.0	12.6

APPENDIX B

Holdup and Pressure Drop Reproducibility Data

<u>Run No.</u>	<u>Y_L</u>	<u>R_L</u>	<u>ΔP, In. Hg.</u>
<u>Air-Water Runs</u>			
G = 25			
248	0.378	0.58	9.95
251	0.378	0.585	9.85
256	0.378	0.59	9.95
259	0.378	0.592	9.85
262	0.378	0.58	9.85
249	0.053	0.265	4.65
255	0.053	0.248	4.35
258	0.053	0.261	4.35
260	0.053	0.248	4.0
266	0.053	0.265	4.35
245	0.0042	0.0785	3.7
250	0.0042	0.0826	3.8
257	0.0042	0.0775	3.65
261	0.0042	0.087	3.65
268	0.0042	0.0785	3.55
<u>Air-Water Runs</u>			
G = 50			
269	0.378	0.51	8.8
275	0.378	0.495	8.6
281	0.378	0.51	8.6
288	0.378	0.511	8.8
296	0.378	0.499	8.5
<u>Air-Sodium Carbonate Solution Runs</u>			
G = 25			
346	0.378	0.664	12.65
405	0.378	0.627	12.4
408	0.378	0.635	11.85
424	0.378	0.609	11.75
430	0.378	0.62	11.65

Holdup and Pressure Drop Reproducibility Data
(Continued)

<u>Run No.</u>	<u>y_L</u>	<u>R_L</u>	<u>ΔP, In. Hg.</u>
<u>Air-Varsol Runs</u>			
G = 25			
588	0.378	0.511	6.95
591	0.378	0.532	7.1
610	0.378	0.528	7.2
613	0.378	0.539	7.1
616	0.378	0.518	7.0
<u>Air-Oil Blend No. 1 Runs</u>			
G = 25			
805	0.378	0.58	7.9
808	0.378	0.56	7.8
811	0.378	0.565	7.7
814	0.378	0.565	7.9
817	0.378	0.558	7.8
<u>Air-Oil Blend No. 2 Runs</u>			
G = 25			
957	0.363	0.73	10.05
958	0.363	0.725	10.2
959	0.363	0.734	10.05
960	0.363	0.729	10.05
961	0.363	0.711	10.05
<u>Air-Trichloroethylene Runs</u>			
G = 37.5			
1025	0.378	0.53	19.0
1026	0.378	0.525	18.0
1027	0.378	0.525	18.0
1028	0.378	0.521	18.0
1029	0.378	0.515	18.2

APPENDIX C

Summary of Calculated Values

Run No.	x	Holdup RL	Slip Velocity Ft/Sec. VS	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Water Runs</u>				
G = 5				
135	0.054	0.23	10.6	13.0
136	0.039	0.209	14.4	12.5
137	0.028	0.179	17.7	11.0
138	0.0225	0.168	21.2	10.5
139	0.018	0.146	24.6	9.5
140	86.5	0.985	0.42	5.0
141	5.7	0.90	0.45	19.5
142	2.63	0.82	0.63	27.2
143	1.22	0.69	0.87	27.2
144	0.76	0.60	1.15	26.0
145	0.54	0.515	1.36	22.3
146	0.43	0.468	1.66	19.2
147	0.42	0.46	1.67	21.0
148	0.31	0.377	1.94	16.0
149	0.22	0.301	2.67	12.5
150	0.14	0.251	4.1	10.2
151	0.097	0.232	5.82	10.2
152	0.08	0.234	6.83	11.2
153	0.05	0.23	7.8	12.0
194	0.034	0.190	15.9	10.7
195	0.019	0.147	24.0	10.0
196	0.0118	0.104	31.9	7.0
197	0.0068	0.065	48.1	5.2
198	0.0056	0.0405	56.0	3.7
G = 10				
54	0.367	0.312	2.0	12.0
55	0.273	0.297	2.9	10.5
56	0.19	0.255	3.95	10.0
57	0.147	0.249	5.6	10.2
58	0.119	0.234	6.7	10.5
59	0.101	0.231	8.7	11.5
60	0.08	0.224	11.4	13.7

Summary of Calculated Values
(Continued)

Run No.	x	Holdup R _L	Slip Velocity Ft/Sec. V _S	Pressure Drop Lb/Sq Ft Per Ft of Pipe ΔP _{TP} /L
<u>Air-Water Runs (Continued)</u>				
61	0.065	0.227	14.7	13.7
78	119.0	0.985	0.33	3.0
79	14.5	0.941	0.40	6.2
80	5.8	0.889	0.61	16.0
81	3.6	0.843	0.64	18.7
82	1.7	0.737	0.93	23.0
83	1.05	0.633	1.15	22.2
84	0.77	0.57	1.43	20.5
85	0.62	0.49	1.54	19.0
187	0.053	0.205	15.9	10.0
188	0.033	0.16	23.6	8.2
189	0.0227	0.118	32.1	6.7
190	0.0163	0.090	40.3	5.0
191	0.0124	0.067	46.7	4.5
192	0.009	0.048	56.7	4.0
193	0.0064	0.039	66.4	4.5
224	0.415	0.415	1.6	15.5
225	0.352	0.352	2.15	17.5
226	0.289	0.289	3.0	10.7
227	0.266	0.266	3.95	10.7
228	0.232	0.232	4.8	10.5
229	0.235	0.235	5.8	10.7
230	0.232	0.232	7.4	10.0
278	0.0048	0.035	74.6	4.0
279	0.0031	0.028	85.1	4.0
280	0.0016	0.020	99.4	4.0

G = 25

33	0.418	0.349	2.36	10.2
34	0.225	0.266	4.8	10.5
35	0.16	0.277	8.66	13.0
36	0.125	0.255	10.7	15.5
37	0.103	0.241	13.4	15.2
86	12.1	0.985	0.57	2.2
87	5.5	0.938	0.45	4.2
88	2.6	0.878	0.6	7.5
89	2.6	0.78	0.81	12.2
90	1.62	0.692	1.07	14.7
91	1.18	0.605	1.18	15.0
92	0.95	0.555	1.38	14.2
158	0.74	0.532	1.65	15.0
180	0.087	0.232	14.75	15.5
181	0.059	0.204	23.7	14.2

Summary of Calculated Values
(Continued)

Run No.	x	Holdup RL	Slip Velocity Ft/Sec. V _S	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Water Runs (Continued)</u>				
182	0.042	0.167	31.4	13.2
183	0.032	0.143	39.6	13.0
184	0.026	0.116	47.5	13.0
185	0.0211	0.10	56.0	12.5
186	0.0175	0.087	65.5	12.5
218	0.56	0.42	1.85	12.0
219	0.407	0.342	2.5	10.5
220	0.31	0.315	3.5	10.7
221	0.246	0.286	4.4	10.5
222	0.212	0.268	5.5	16.7
223	0.175	0.258	6.8	12.5
245	0.0151	0.0785	74.5	12.2
246	0.0127	0.066	82.5	12.2
247	0.010	0.056	95.4	13.5

G = 50

27	0.595	0.41	1.75	9.0
28	0.318	0.336	4.35	10.5
29	0.22	0.306	7.5	13.5
30	0.174	0.289	9.5	15.5
31	0.144	0.278	12.3	16.7
47	0.595	0.415	1.8	8.2
99	80.0	0.985	0.2	1.0
100	16.9	0.948	0.2	1.7
101	7.7	0.90	0.4	3.5
102	3.74	0.827	0.7	6.2
103	2.3	0.74	0.8	8.0
104	1.67	0.693	1.1	9.5
105	1.34	0.643	1.65	11.2
120	0.194	0.303	8.85	21.7
121	0.145	0.286	12.6	20.7
122	0.115	0.261	16.6	21.0
123	0.098	0.238	19.7	21.0
124	0.085	0.24	23.8	20.2
157	1.12	0.607	1.45	9.7
173	0.126	0.248	13.3	18.5
174	0.084	0.232	22.2	18.7
175	0.062	0.196	29.9	18.2
176	0.05	0.175	36.9	18.5
177	0.040	0.16	45.6	19.5
178	0.0335	0.151	54.7	21.0
179	0.0282	0.136	63.3	19.2
212	0.82	0.503	1.6	10.0

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup R_L</u>	<u>Slip Velocity V_S</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$</u>
<u>Air-Water Runs (Continued)</u>				
213	0.59	0.41	2.0	9.7
214	0.443	0.36	2.7	10.7
215	0.36	0.36	4.0	11.0
216	0.305	0.311	4.4	12.5
217	0.23	0.311	6.7	13.5
252	0.025	0.126	70.3	20.5
253	0.0211	0.115	78.3	21.7
254	0.0185	0.110	91.5	22.7

G = 100

93	56.5	0.964		1.0
94	16.8	0.935	0.25	2.5
95	7.05	0.878	0.55	3.2
96	4.15	0.831	0.96	5.7
97	3.2	0.788	1.24	6.2
98	2.4	0.738	1.26	7.0
115	0.376	0.35	6.5	18.5
116	0.278	0.311	9.6	21.7
117	0.228	0.296	12.9	23.2
118	0.19	0.272	15.5	24.5
119	0.166	0.259	18.6	26.0
156	2.2	0.71	1.28	7.5
166	0.248	0.312	11.9	22.5
167	0.172	0.269	18.6	26.7
168	0.125	0.24	26.2	29.0
169	0.10	0.221	33.2	30.7
170	0.083	0.202	40.6	32.5
171	0.070	0.19	48.0	34.7
172	0.060	0.182	55.7	35.7
206	1.56	0.63	1.7	7.5
207	1.12	0.53	1.7	9.2
208	0.86	0.47	2.2	11.5
209	0.70	0.443	3.2	12.7
210	0.60	0.41	3.9	13.7
211	0.455	0.363	5.2	17.2
263	0.054	0.16	61.8	37.2
264	0.048	0.154	68.5	38.2
265	0.042	0.147	77.8	39.0

G = 150

40	3.4	0.77	1.15	6.7
41	1.78	0.65	2.5	9.7

Summary of Calculated Values
(Continued)

Run No.	x	Holdup R _L	Slip Velocity Ft/Sec. V _S	Pressure Drop Lb/Sq Ft Per Ft of Pipe ΔPTP/L
<u>Air-Water Runs (Continued)</u>				
42	0.92	0.489	3.0	15.2
43	0.65	0.43	4.9	18.0
44	0.51	0.381	6.35	21.2
62	2.3	0.692	1.4	9.0
63	1.22	0.565	2.85	12.5
64	0.77	0.455	4.0	14.7
72	82.5	0.972	-	2.2
73	21.0	0.955	0.4	3.0
74	10.2	0.913	0.55	4.0
75	6.4	0.876	1.0	5.0
76	4.75	0.832	1.0	6.0
77	3.75	0.80	1.2	6.2
106	0.57	0.42	6.6	21.2
107	0.43	0.378	9.1	25.0
108	0.35	0.333	10.6	27.5
109	0.29	0.314	13.9	31.5
110	0.26	0.31	17.1	33.0
155	4.0	0.77	1.2	11.7
159	0.38	0.332	9.0	28.2
160	0.255	0.297	15.9	34.7
161	0.192	0.272	22.4	38.2
162	0.163	0.249	28.4	40.7
163	0.132	0.235	34.3	45.0
164	0.111	0.218	40.9	47.0
165	0.098	0.199	45.9	51.0
200	2.32	0.702	1.45	6.5
201	1.67	0.63	1.85	9.5
202	1.29	0.58	2.63	11.7
203	1.05	0.521	2.9	16.2
204	0.89	0.485	3.36	15.0
205	0.665	0.425	4.3	17.5
272	0.087	0.196	53.3	53.2
273	0.078	0.181	58.9	55.2
274	0.068	0.168	67.1	60.0

Summary of Calculated Values
(Continued)

Run No.	x	Holdup R_L	Slip Velocity Ft/Sec. V_S	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Sodium Carbonate Solution Runs</u>				
G = 5				
371	12.1	0.958	0.56	17.7
372	3.1	0.844	0.70	35.0
373	1.18	0.605	0.80	27.2
374	0.78	0.483	0.99	23.5
375	0.53	0.39	1.31	20.0
376	0.42	0.377	1.68	17.5
377	0.246	0.312	2.76	14.2
378	0.157	0.266	4.28	12.0
379	0.114	0.256	5.93	11.7
380	0.092	0.235	6.85	12.0
381	0.082	0.228	7.85	11.2
382	0.0445	0.215	12.1	12.2
383	0.0284	0.175	19.8	11.2
384	0.0128	0.127	32.9	8.0
402	0.056	0.102	41.6	7.0
403	0.0305	0.073	49.0	5.2
404	0.0215	0.0615	53.2	3.7
G = 10				
357	16.5	0.965	0.64	12.0
358	4.3	0.852	0.66	23.7
359	1.63	0.638	0.80	22.2
360	1.09	0.55	1.02	24.5
361	0.74	0.443	1.27	17.2
362	0.60	0.423	1.63	17.5
363	0.35	0.317	2.53	12.7
364	0.227	0.283	4.1	12.0
365	0.166	0.249	5.5	11.7
366	0.135	0.263	6.8	12.0
367	0.123	0.255	7.9	12.0
368	0.080	0.23	12.0	13.7
369	0.048	0.192	19.4	12.7
370	0.0264	0.143	33.0	10.0
395	0.0264	0.147	32.3	10.2
396	0.0184	0.118	41.7	8.2
397	0.014	0.094	49.4	7.5
398	0.0107	0.0785	57.4	7.5
399	0.0076	0.0675	69.1	6.2
400	0.0056	0.0575	77.6	6.7
401	0.0024	0.0475	95.9	6.7

Summary of Calculated Values
(Continued)

Run No.	x	Holdup RL	Slip Velocity Ft/Sec. VS	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Sodium Carbonate Solution Runs (Continued)</u>				
G = 25				
343	25.4	0.97	0.55	6.5
344	6.6	0.885	0.65	10.5
345	2.44	0.725	0.80	16.0
346	1.64	0.628	1.0	16.5
347	1.15	0.545	1.2	15.5
348	0.93	0.50	1.5	15.0
349	0.57	0.395	2.3	13.2
350	0.35	0.348	3.9	12.2
351	0.266	0.301	5.25	13.5
352	0.216	0.289	6.2	13.0
353	0.201	0.27	7.0	13.0
354	0.129	0.227	10.8	16.5
355	0.0785	0.221	19.0	15.5
356	0.047	0.168	31.1	13.5
391	0.037	0.148	39.4	13.2
392	0.030	0.126	46.8	12.2
393	0.024	0.109	55.3	13.2
394	0.020	0.098	66.4	12.5
417	0.0166	0.085	73.8	13.2
418	0.0132	0.071	88.3	14.2
G = 50				
329	35.2	0.978	0.1	1.5
330	9.2	0.915	0.45	4.5
331	3.55	0.795	0.7	10.0
332	2.38	0.712	0.8	11.7
333	1.66	0.63	1.1	10.7
334	1.32	0.60	1.3	11.7
335	0.79	0.481	2.1	12.0
336	0.52	0.384	3.0	11.5
337	0.38	0.335	4.2	15.2
338	0.31	0.307	5.0	14.7
339	0.284	0.303	6.0	15.2
340	0.189	0.284	10.4	19.2
341	0.116	0.225	16.7	21.0
342	0.070	0.204	28.1	18.7
389	0.046	0.155	43.5	19.7
390	0.032	0.134	61.9	21.2
415	0.0266	0.118	73.7	22.7
416	0.0226	0.101	81.7	23.7

Summary of Calculated Values
(Continued)

Run No.	x	Holdup R_L	Slip Velocity Ft/Sec. V_S	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Sodium Carbonate Solution Runs (Continued)</u>				
G = 100				
316	16.8	0.948	0.9	3.0
317	6.8	0.86	0.8	5.0
318	4.6	0.801	0.9	5.7
319	3.2	0.73	1.0	4.0
320	2.62	0.70	1.4	8.0
321	1.56	0.59	2.2	10.5
322	1.23	0.575	4.5	17.0
323	0.74	0.43	3.8	15.2
324	0.61	0.406	4.6	16.7
325	0.55	0.367	4.8	17.2
326	0.51	0.367	5.7	18.2
327	0.225	0.272	15.0	26.7
328	0.14	0.213	24.8	30.0
387	0.092	0.182	38.1	31.0
388	0.068	0.171	56.3	34.7
413	0.055	0.153	64.4	35.0
414	0.049	0.144	74.5	39.2

G = 150

300	97.0	0.98	-	1.75
301	23.5	0.966	-	3.0
302	12.3	0.935	1.9	4.0
303	7.5	0.871	0.9	5.5
304	6.85	0.84	1.2	6.0
305	4.7	0.793	1.1	7.5
306	3.9	0.767	1.5	8.5
307	2.3	0.666	2.2	11.5
308	1.5	0.579	3.0	14.5
309	1.1	0.497	3.6	16.7
310	0.89	0.462	4.2	18.2
311	0.83	0.435	4.7	19.0
312	0.77	0.433	5.5	20.5
313	0.344	0.311	13.5	33.0
314	0.218	0.27	23.5	40.0
385	0.146	0.225	37.2	45.5
386	0.108	0.186	48.1	48.5
411	0.091	0.180	58.0	51.0
412	0.079	0.160	64.8	53.0

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup R_L</u>	<u>Slip Velocity Ft/Sec. V_S</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$</u>
<u>Air-Varsol Runs</u>				
G = 5				
594	9.8	0.941	0.4	8.2
595	2.52	0.803	0.55	16.0
596	0.95	0.485	0.57	14.0
597	0.64	0.43	0.88	11.2
598	0.433	0.35	1.2	9.2
599	0.368	0.295	1.4	7.5
600	0.209	0.285	2.7	7.0
601	0.134	0.25	4.2	7.5
602	0.098	0.278	6.14	8.2
603	0.08	0.254	7.11	8.2
604	0.073	0.258	8.35	8.7
605	0.046	0.225	12.5	8.7
606	0.025	0.168	20.1	6.7
607	0.0118	0.103	32.0	3.7
608	0.005	0.064	44.5	2.5
609	0.0044	0.061	46.5	2.5
G = 10				
570	13.4	0.94	0.28	4.2
571	3.5	0.84	0.61	7.0
572	1.31	0.591	0.63	10.7
573	0.89	0.46	0.71	9.5
574	0.61	0.392	1.04	7.7
575	0.49	0.356	1.32	7.5
576	0.29	0.311	2.5	6.2
577	0.188	0.283	4.0	7.5
578	0.14	0.292	6.0	8.2
579	0.115	0.286	7.1	8.2
580	0.104	0.28	8.1	9.7
581	0.069	0.241	12.3	9.0
582	0.039	0.188	20.2	7.5
583	0.022	0.132	31.9	5.7
584	0.0143	0.090	43.6	4.7
585	0.0094	0.070	57.4	4.5
586	0.00495	0.052	77.6	4.5
587	0.00155	0.039	103.3	4.5
G = 25				
516	20.6	0.985	1.4	1.7
517	5.35	0.907	0.9	5.5

Summary of Calculated Values
(Continued)

Run No.	x	Holdup R _L	Slip Velocity Ft/Sec. V _S	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Varsol Runs (Continued)</u>				
518	2.08	0.72	0.7	7.2
519	1.38	0.585	0.1	6.5
520	0.97	0.51	0.88	5.5
521	0.78	0.46	1.0	6.7
522	0.475	0.353	1.8	6.0
523	0.305	0.311	3.0	6.2
524	0.23	0.28	4.6	7.5
525	0.187	0.299	6.1	9.0
526	0.172	0.272	6.8	9.7
527	0.158	0.264	7.1	10.0
528	0.068	0.202	18.5	9.2
529	0.041	0.151	30.0	8.7
530	0.028	0.126	43.2	9.0
531	0.0206	0.115	57.4	9.5
532	0.0148	0.087	73.6	10.0
533	0.0103	0.070	95.2	10.7

G = 50

534	7.4	0.945	1.3	3.0
535	2.94	0.815	0.8	4.5
536	1.95	0.723	0.7	4.5
537	1.35	0.617	0.75	5.0
538	1.1	0.575	0.8	5.0
539	0.66	0.448	1.4	7.2
540	0.43	0.37	2.1	6.2
541	0.32	0.345	3.9	7.5
542	0.26	0.32	4.6	8.5
543	0.243	0.308	5.6	9.5
544	0.16	0.266	9.3	11.7
545	0.099	0.218	16.2	13.0
546	0.0603	0.174	26.4	13.2
547	0.0435	0.148	38.6	14.2
548	0.0325	0.131	52.3	15.7
549	0.0245	0.112	64.5	16.7
550	0.0185	0.106	88.0	18.2

G = 100

500	5.55	0.876	0.6	3.2
501	3.7	0.795	0.2	4.2
502	2.6	0.75	0.7	4.5
503	2.1	0.71	0.8	3.2
504	1.27	0.641	2.1	5.5
505	0.83	0.508	1.8	5.5

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup RL</u>	<u>Slip Velocity Ft/Sec. Vs</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$</u>
<u>Air-Varsol Runs (Continued)</u>				
506	0.61	0.447	3.1	8.0
507	0.51	0.41	3.5	8.2
508	0.475	0.389	4.2	9.7
509	0.445	0.375	4.4	8.2
510	0.192	0.255	12.5	16.5
511	0.122	0.211	19.0	18.7
512	0.09	0.188	32.5	24.0
513	0.07	0.166	42.4	26.0
514	0.054	0.153	58.5	28.0
515	0.042	0.133	72.5	30.7

G = 150

554	8.25	0.878	-	2.7
555	5.5	0.855	0.1	4.5
556	3.8	0.807	0.4	4.7
557	3.2	0.772	0.6	6.2
558	1.91	0.667	0.9	6.5
559	1.28	0.583	1.2	7.7
560	0.94	0.513	2.1	9.7
561	0.78	0.49	2.9	11.0
562	0.71	0.468	3.6	12.5
563	0.47	0.39	6.1	15.7
564	0.25	0.286	12.5	24.5
565	0.173	0.235	19.7	30.2
566	0.122	0.199	31.8	34.2

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup R_L</u>	<u>Slip Velocity Ft/Sec. V_S</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$</u>
<u>Air-Oil Blend No. 1 Runs</u>				
G = 5				
785	13.2	0.958	0.6	9.5
786	3.37	0.795	0.55	18.2
787	1.29	0.577	0.77	16.0
788	0.87	0.482	1.0	13.7
789	0.605	0.423	1.4	12.2
790	0.48	0.409	1.8	12.2
791	0.283	0.358	3.29	11.2
792	0.193	0.322	4.42	9.5
793	0.132	0.28	6.5	9.2
794	0.097	0.272	8.9	9.5
795	0.060	0.235	13.1	9.0
796	0.025	0.173	25.4	7.2
797	0.0119	0.118	37.1	5.0
798	0.0068	0.098	45.5	5.0
799	0.00285	0.0755	54.7	4.2
800	0.0012	0.056	59.0	3.5
G = 10				
766	18.4	0.965	0.6	5.2
767	4.65	0.84	0.6	11.7
768	1.84	0.625	0.7	12.7
769	1.2	0.532	0.9	11.7
770	0.84	0.465	1.3	10.5
771	0.74	0.400	1.5	10.7
772	0.4	0.373	2.8	10.5
773	0.277	0.334	4.2	10.2
774	0.194	0.300	6.0	9.2
775	0.144	0.269	8.0	9.7
776	0.094	0.224	11.9	10.0
777	0.045	0.193	24.6	9.0
778	0.022	0.140	42.3	6.7
779	0.0122	0.103	60.0	7.0
780	0.0062	0.073	79.1	7.5
781	0.0021	0.0475	101.5	8.0
G = 25				
715	7.0	0.905	0.9	6.5
716	2.8	0.76	0.9	8.2
717	1.89	0.667	1.1	9.7
718	1.31	0.589	1.4	9.7

Summary of Calculated Values
(Continued)

Run No.	x	Holdup R_L	Slip Velocity Ft/Sec. V_S	Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$
<u>Air-Oil Blend No. 1 Runs (Continued)</u>				
719	1.06	0.535	1.6	10.2
720	0.64	0.454	2.7	11.7
721	0.44	0.40	4.0	10.5
722	0.31	0.353	5.5	11.0
723	0.23	0.331	7.5	11.0
724	0.153	0.294	12.1	11.7
725	0.075	0.213	36.4	12.0
726	0.0435	0.172	40.5	12.0
727	0.0285	0.148	58.6	12.0
728	0.0201	0.118	76.0	12.5
729	0.0139	0.101	99.0	14.2

G = 50

732	10.1	0.93	0.85	2.7
733	4.0	0.84	1.15	6.5
734	2.7	0.775	1.3	9.0
735	1.9	0.69	1.3	7.7
736	1.51	0.653	1.7	8.5
737	0.91	0.533	2.4	9.5
738	0.62	0.461	3.4	9.7
739	0.435	0.411	5.15	11.5
740	0.325	0.35	6.5	12.2
741	0.218	0.286	10.0	14.0
742	0.111	0.233	21.4	16.7
743	0.0715	0.175	36.0	18.7
744	0.045	0.147	52.5	19.2
745	0.0335	0.130	70.5	20.0
746	0.0252	0.122	90.4	21.2

G = 100

700	18.6	0.922	-	3.2
701	7.6	0.88	0.8	4.7
702	5.1	0.85	1.4	6.0
703	3.64	0.785	1.3	6.5
704	3.0	0.749	1.5	7.2
705	1.73	0.647	2.3	10.0
706	1.22	0.563	2.9	10.5
707	0.85	0.485	4.0	13.0
708	0.64	0.42	5.1	14.5
709	0.43	0.345	7.9	17.5
710	0.222	0.274	17.5	25.2
711	0.134	0.218	30.2	30.0

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup R_L</u>	<u>Slip Velocity Ft/Sec. V_S</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe ΔPTP/L</u>
<u>Air-Oil Blend No. 1 Runs (Continued)</u>				
712	0.095	0.193	45.5	31.7
713	0.074	0.168	57.2	34.0
714	0.057	0.155	75.0	36.0
G = 150				
750	11.4	0.909	0.6	4.5
751	7.5	0.89	1.6	5.7
752	5.3	0.846	1.8	7.0
753	4.35	0.827	2.4	8.2
754	2.61	0.73	2.6	10.5
755	1.7	0.65	3.7	11.7
756	1.25	0.575	4.1	14.7
757	0.97	0.513	5.2	16.7
758	0.645	0.407	6.75	20.7
759	0.332	0.297	13.6	30.2
760	0.202	0.246	25.8	34.7
761	0.155	0.21	34.5	43.7
762	0.118	0.193	47.0	49.0

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup R_L</u>	<u>Slip Velocity Ft/Sec. V_S</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$</u>
<u>Air-Oil Blend No. 2 Runs</u>				
G = 25				
942	9.9	0.913	1.0	6.7
943	3.9	0.805	1.3	12.2
944	2.61	0.745	1.65	12.5
945	1.84	0.672	1.95	13.7
946	1.48	0.635	2.3	14.0
947	0.89	0.56	3.7	15.2
948	0.575	0.485	5.2	14.7
949	0.435	0.435	6.9	14.7
950	0.33	0.387	8.7	15.7
951	0.211	0.336	13.1	16.5
952	0.109	0.289	25.8	18.5
953	0.062	0.249	43.9	19.7
954	0.0415	0.196	60.6	21.0
955	0.0293	0.165	79.2	22.0
956	0.0207	0.129	97.7	24.0
G = 50				
927	13.5	0.943	1.3	6.7
928	5.35	0.868	1.7	12.2
929	3.65	0.813	1.9	12.5
930	2.56	0.748	2.1	13.7
931	2.07	0.708	2.3	14.0
932	1.22	0.627	3.6	15.2
933	0.805	0.535	4.9	14.7
934	0.6	0.485	6.5	14.7
935	0.46	0.431	8.15	15.7
936	0.3	0.372	12.0	16.5
937	0.153	0.288	22.9	18.5
938	0.092	0.263	39.2	19.7
939	0.0635	0.232	56.4	21.0
940	0.0465	0.218	76.8	22.0
941	0.0352	0.201	95.2	24.0
G = 100				
912	25.0	0.965	1.8	6.5
913	10.2	0.92	2.3	8.7
914	6.95	0.88	2.4	10.5
915	4.91	0.832	2.6	11.7
916	4.0	0.795	2.6	13.0
917	2.35	0.715	3.7	16.5

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup RL</u>	<u>Slip Velocity Ft/Sec. Vs</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe $\Delta P_{TP}/L$</u>
<u>Air-Oil Blend No. 2 Runs (Continued)</u>				
918	1.54	0.653	5.5	19.7
919	1.15	0.585	6.6	21.7
920	0.88	0.493	7.3	23.7
921	0.60	0.400	9.7	28.5
922	0.31	0.325	19.7	37.2
923	0.184	0.277	34.1	47.0
924	0.133	0.221	45.1	50.5
925	0.101	0.207	59.8	53.2
926	0.0765	0.21	89.5	55.7

G = 150

900	15.0	0.945	3.4	10.7
901	10.2	0.895	1.9	12.2
902	7.25	0.875	3.1	13.7
903	5.95	0.86	4.2	14.5
904	3.55	0.781	4.3	18.5
905	2.35	0.72	6.0	20.0
906	1.75	0.642	6.2	24.5
907	1.32	0.58	7.5	28.0
908	0.9	0.49	9.8	33.0
909	0.475	0.367	17.2	45.0
910	0.30	0.32	28.5	56.0
911	0.208	0.277	42.3	61.2

Summary of Calculated Values
(Continued)

<u>Run No.</u>	<u>x</u>	<u>Holdup R_L</u>	<u>Slip Velocity Ft/Sec. V_S</u>	<u>Pressure Drop Lb/Sq Ft Per Ft of Pipe ΔP_{TP}/L</u>
<u>Air-Trichloroethylene Runs</u>				
G = 37.5				
1013	1.62	0.70	0.5	14.0
1014	1.09	0.57	0.5	12.2
1015	0.74	0.487	0.75	10.2
1016	0.60	0.437	0.95	11.5
1017	0.35	0.336	1.7	11.7
1018	0.23	0.314	3.2	12.5
1019	0.172	0.266	4.3	16.0
1020	0.132	0.244	6.1	17.5
1021	0.075	0.216	10.0	18.2
1022	0.043	0.157	20.4	16.2
1023	0.0234	0.112	36.8	13.0
1024	0.016	0.078	50.9	13.0
G = 150				
1000	5.35	0.888	0.9	7.0
1001	3.53	0.818	0.7	6.2
1002	2.46	0.736	0.5	7.2
1003	2.0	0.685	0.6	9.0
1004	1.17	0.565	0.9	10.0
1005	0.77	0.455	1.5	11.7
1006	0.565	0.405	2.1	15.2
1007	0.423	0.352	3.4	19.0
1008	0.270	0.288	6.6	24.7
1009	0.151	0.221	13.4	32.5
1010	0.0915	0.165	24.3	36.2
1011	0.066	0.137	44.6	43.5
1012	0.050	0.132	47.5	42.0

APPENDIX D

Sample Calculations

Data for Run 307, Appendix A

air rate = 0.61 CFM at average pressure of test section

liquid rate = 0.73 CFM

holdup = 2340 cc

base pressure = 9.2 psig

pressure differential = 14.8 inches Hg

total mass velocity = 150 lb/sq ft sec.

Liquid Holdup, R_L

Experimental Holdup

holdup + liquid between valve and drain + liquid adhering to

wall = total holdup

$$2340 + 20 + 25 = 2385$$

total volume of section = 3570 cc

$$R_L = \frac{2385}{3570} = 0.666$$

Calculated Holdup

$$x = \left(\frac{W_L}{W_G} \right)^{0.9} \frac{\mu_L^{0.19} \sigma^{0.205} \rho_G^{0.70}}{G^{0.435} \rho_L^{0.72}} \quad (\text{Equation 10})$$

$$\rho_L = 74.0 \text{ lb/cu ft}$$

$$\mu_L = 3.15 \text{ centipoises}$$

$$\sigma = 75.9 \text{ dynes/cm}$$

$$G = 150 \text{ lb/sq ft sec.}$$

As G is greater than 50, use 50

For ρ_G (air density at average test section conditions)

pressure at base of test section = 9.2 psig

pressure differential for test section = 14.8 inches Hg or
7.25 psi

pressure at top of test section = $9.2 - 7.25 = 1.95$ psig

average pressure in test section = $\frac{9.2 + 1.95}{2} = 5.58$ psig

air density at 75°F and 14.7 psig = $\frac{(28.8)(520)}{(379)(535)} = 0.074$ lb/cu ft

28.8 = molecular weight of air

379 = volume of one mol of gas at 520°R and 14.7 psig

535 = air temperature °F

$$\rho_G = \frac{(0.074)(14.7 + 5.58)}{14.7} = 0.102 \text{ lb/cu ft}$$

$$W_G = 0.61 \rho_G = (0.61)(0.102) = 0.0622 \text{ lb/min.}$$

$$W_L = 0.73 \rho_L = (0.73)(74) = 54 \text{ lb/min.}$$

$$x = \left(\frac{54}{0.0622} \right)^{0.9} \frac{(3.15)^{0.19} (75.9)^{0.205} (0.102)^{0.70}}{(50)^{0.435} (74)^{0.72}}$$

$$x = 2.3$$

From Figure 12 for $x = 2.3$, calculated $R_L = 0.71$

$$\text{Actual } R_L - \text{calculated } R_L = 0.666 - 0.71 = -0.044$$

Pressure Drop, ΔP_{TP}

Experimental Pressure Drop

$$(Z_2 - Z_1) \left(\frac{W_L + W_G}{W_L v_L + W_G v_G} \right) + p_2 - p_1 + \Delta P_{TP} = 0 \quad (\text{Equation 11})$$

$$Z_2 - Z_1 = 20 \text{ ft}$$

$$v_L = \frac{1}{\rho_L} = 0.0135 \text{ cu ft/lb}$$

$$v_G = \frac{1}{\rho_G} = \frac{1}{0.102} = 9.8 \text{ cu ft/lb}$$

$$p_2 - p_1 = -7.25 \text{ psi or } -1040 \text{ psf}$$

$$(20) \left[\frac{54 + 0.062}{(54)(0.0135) + (0.062)(9.8)} \right] - 1040 + \Delta P_{TP} = 0$$

$$\Delta P_{TP} = 1040 - 810 = 230 \text{ lb/sq ft}$$

Calculated Pressure Drop

$$\psi = \frac{1}{\mu_L^{0.149} \sigma^{0.194} G^{0.70}} \quad (\text{Equation 13})$$

$$\psi = \frac{1}{(3.15)^{0.149} (75.9)^{0.194} (150)^{0.70}}$$

$$\psi = 0.011$$

Liquid velocity in pipe

$$\text{flow rate} = \frac{54}{60 \times 74} = 0.0122 \text{ cu ft/sec.}$$

$$\text{cross sectional area of pipe} = 0.006 \text{ sq ft}$$

$$\text{area for liquid flow} = 0.006 R_L = 0.004 \text{ sq ft}$$

$$\text{liquid velocity} = \frac{0.0122}{0.004} = 3.05 \text{ ft/sec.}$$

Gas velocity in pipe

$$\text{flow rate} = \frac{0.0622}{60 \times 0.102} = 0.01015 \text{ cu ft/sec.}$$

$$\text{area for gas flow} = 0.006 (1 - R_L) = 0.002 \text{ sq ft}$$

$$\text{gas velocity} = \frac{0.0105}{0.002} = 5.25 \text{ ft/sec.}$$

$$\text{Slip velocity, } V_S = 5.25 - 3.05 = 2.2 \text{ ft/sec.}$$

$$\text{From Figure 28 for } V_S = 2.2, \psi = 0.011$$

$$\frac{\Delta P_{TP} - \Delta P_L}{L} = 8 \text{ lb/sq ft per ft of pipe}$$

For total liquid flow at $G = 150 \text{ lb/sq ft sec.}$

$$Re = \frac{D G}{\mu} = \frac{(0.0875)(150)(3600)}{(3.15)(2.42)} = 6200$$

From Figure 3, at $Re = 6200$ $f = 0.0375$

$$\Delta P_L = \frac{f L G^2}{2 D_g \rho} = \frac{(0.0375)(1)(150)^2}{(2)(0.0875)(32.2)(74)} = 2.0 \text{ lb/sq ft}$$

$$\Delta P_{TP} = 8 + 2 = 10 \text{ lb/sq ft per ft for 20 ft of}$$

$$\text{pipe} = 200 \text{ lb/sq ft}$$

$$\text{Total pressure drop} = 810 + 200 = 1010 \text{ lb/sq ft}$$

$$\% \text{ error between observed and calculated} = \frac{1040 - 1010}{1040} = 3\%$$

AUTOBIOGRAPHY

Gordon A. Hughmark was born in Cleveland, Ohio on July 15, 1924. Grade school and high school educations were obtained in Cincinnati, Ohio. He graduated from the University of Cincinnati in 1947 with the degree of Chemical Engineer. Graduate work at Louisiana State University from 1951 to 1955 led to the M.S. degree in Chemical Engineering in 1955.

He was employed as a Chemical Engineer by the Vulcan Copper and Supply Company, Cincinnati, Ohio from 1947 to 1951 and by the Ethyl Corporation, Baton Rouge, Louisiana from February 1951 to the present.

EXAMINATION AND THESIS REPORT

Candidate: Gordon A. Hughmark

Major Field: Chemical Engineering

Title of Thesis: Holdup and Pressure Drop with Gas-Liquid Flow in
A Vertical Pipe.

Approved:

Major Professor and Chairman

Dean of the Graduate School

EXAMINING COMMITTEE:

Date of Examination:

May 12, 1959