A METHOD TO CALCULATE PRESSURE DROP FOR GAS-LIQUID FLOW IN LONG HORIZONTAL TRANSMISSION LINES

By

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in partial fulfillment of the requirements for
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DOCTOR OF PHILOSOPHY
December, 1987

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PREFACE

A method was developed to predict the pressure drop for gas-liquid flow in long horizontal transmission lines. The method is based on the assumption that the two phases flow at different velocities within the pipeline and provides a better estimate of the pressure drop than obtained by assuming homogeneous flow.

I wish to express my deep gratitude to my adviser, Dr. R. N. Maddox, for his help, patience, and guidance throughout my work at Oklahoma State University. I would like to thank other members of my advisory committee, Dr. K. J. Bell, Dr. G. L. Foutch, and Dr. C. E. Price for their invaluable suggestions.

My wife, Rajni, son, Gaurav, and my Mother, deserve my appreciation for their constant support and encouragement.

Finally, I would like to thank Pam Hartman for typing this thesis and for all the help she provided.

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NOMENCLATURE

A	pipe cross-section, ft^2
D	pipe diameter, ft
f	friction factor
F_e	Flanigan elevation factor (Equation II-23)
g	acceleration due to gravity, 32.2 ft/sec^2
g_c	gravitational constant
G	superficial gas mass velocity, lb/ft^2 sec
G''	gas mass velocity, lb/ft^2 sec (Figure 5)
G_m	mixture mass velocity, $lb/ft^2/sec$ (Equation II-6)
H_L	liquid holdup, volume percent (defined on page 11)
H_L'	holdup, percent pipe cross-section (defined on page 35) For a given length of the pipe, H_L' also represents the volume percent liquid (also referred to as pseudo holdup in Chapter IV)
h	elevation, ft (Equation II-23)
L	superficial liquid mass velocity, lb/ft^2 sec
L'	line length, ft
L''	liquid mass velocity, lb/ft^2 sec (Figure 5)
N_d	pipe diameter number, dimensionless (defined in Equation II-31, page 22)
N_{gv}	gas velocity number, dimensionless (defined in Equation II-30, page 22)

N_{LV}	liquid velocity number, dimensionless (defined in Equation II-29, page 22)
N_L	liquid viscosity number, dimensionless (defined in Equation II-31, page 22)
N_{FR}	Froude number, dimensionless (defined in Equation II-22, page 14)
P	pressure, psia
ΔP	pressure drop, psia
$(\frac{dP}{dZ})$	pressure gradient, psia/ft
q	volume flow rate, ft^3/sec
Re	two-phase Reynolds number as defined in Equation II-19 for the Beggs and Brill method
Re_{tp}	two-phase Reynolds number as defined in Equation II-37 and presented in Figure 3 for the AGA method
S	correlating parameter for slip velocity
V	velocity, ft/sec
X	correlating parameter in the Lockhart-Martinelli correlation (Equation II-1)
x	ratio of the gas mass flow rate to the total mass flow rate (Equation II-4)
x_L'	mole fraction of liquid in Equation II-41
Z	elevation, ft

Greek Letters

ρ	density, lb/ft^3
Θ	angle of inclination

λ	volume fraction liquid (defined in Equation II-18, page 13). This parameter is expressed as volume percent in Figures 4, 5, 7, 8, and 9
λ'	Baker parameter in Figures 1 and 6 (defined on page 8)
Ψ	Baker parameter in Figures 1 and 6 (defined on page 8)
Ψ'	correlating function for liquid holdup in Equation II-30
σ	surface tension, dynes/cm
μ	viscosity, centipoise, lb/ft sec
ϕ	correlating parameter in the Lockhart-Martinelli method
Σ	summation symbol
τ	shear stress (Equation II-23)

Subscripts

b	base (Equation II-34)
d	subscript for diameter number in Equations II-30 and II-33
g	gas
gv	subscript for gas velocity number in Equations II-30 and II-32
ℓ	liquid
ℓv	subscript for liquid velocity number in Equations II-30 and II-31
m, mc	mixture
ns	no-slip
tp	two-phase
s_g	superficial gas velocity
s_L	superficial liquid velocity

CHAPTER I

INTRODUCTION

Pipelines are commonly used for transporting gas-liquid mixtures. In some cases, a two-phase line can save capital cost by 20 to 25% over two single-phase lines. Properly designing a two-phase line requires accurate knowledge of the two-phase pressure drop. The introduction of a small amount of liquid into a gas stream can increase the pressure drop by an order of magnitude over what might be expected for gas flow alone.

One of the major steps in calculating the pressure drop in two-phase flow is determining the liquid holdup. The liquid holdup in a section of a pipeline is the inventory of liquid within that section. At a given point within the pipeline, the holdup depends on the velocity and the density of the two phases. The holdup is used to calculate the two-phase density and the two-phase friction factor. A number of empirical correlations are available to estimate the liquid holdup.

Although many methods are available for calculating the pressure drop in gas-liquid flow, no single method has proved the best for use under all operating conditions. Akashah et al. (2) developed a method using an equation of state to calculate the thermodynamic properties. In this approach, the thermodynamic properties are calculated from the Soave-Redlich-Kwong (SRK) equation of state, using the algorithm in the computer program UNICALC (32). The liquid and vapor viscosities are calculated using the correlations of Thodos and coworkers (38). The surface tension is calculated from equations given in the GPSA Data Book (23). Akashah (1) calculated the pressure drop for a gas-condensate mixture flowing through a horizontal pipeline with three of the more widely used methods -- the Beggs and Brill, the AGA, and the Lockhart-Martinelli method. For the set of experimental data Akashah considered, only the Lockhart-Martinelli method predicted a pressure drop higher than the observed pressure drop. Many of the more widely used methods for calculating the pressure drop in long transmission lines predict a pressure drop much lower than what is actually encountered.

The existing methods for pressure drop calculations assume homogeneous or no-slip flow. This means that both the gas and the liquid move down the pipeline at the same velocity. The methods use superficial phase velocities and the mixture velocity in the calculation procedure. The superficial velocity of a phase is the velocity the phase would have if it occupied the entire conduit. The mixture velocity is calculated by dividing the total volumetric flowrate of gas and liquid at pipeline conditions by the cross-section area of the pipeline. In reality, a phase occupies only a portion of the pipe cross-section. The gas slips by the liquid and the two-phases generally move at different velocities. Also, the velocity of the gas-liquid mixture controls the two-phase pressure drop. The higher the mixture velocity, the greater the pressure drop.

The present work is an attempt to develop a methodology to artificially increase the mixture velocity so that a higher calculated pressure drop may be obtained by using the more widely used methods for flow in long, horizontal, transmission lines. Such an approach should provide a better estimate of the pressure drop and improve the reliability of the existing methods.

CHAPTER II

LITERATURE REVIEW

Pressure drop in pipelines results from elevation changes, friction losses, and acceleration losses. Any correlation should consider the pressure effects resulting from these losses. So far, the many attempts to manipulate physical data defining these losses to fit actual pressure drops have had limited success. Several investigators have separated experimental data into several groups of observed flow patterns or regimes and then developed a correlation for each flow regime. This is a convenient way of correlating widely scattered data. This chapter briefly reviews some of the widely used pressure drop calculation methods using this approach.

A large number of books, technical papers, and reports have been published on gas-liquid flow based on data collected from laboratory, pilot-plant, or full-scale tests using a limited number of fluids, flow rates, and pipe sizes. For example, Collier (14) has presented a comprehensive analysis of heat transfer and pressure drop in two-phase flow. Much of the information in his book is designed for application in nuclear reactors. Daly and Harlow (15) performed a numerical study to derive a model of countercurrent steam-water flow in large horizontal pipes. The model is useful for designing the emergency core cooling system for a pressurized water reactor. Bell (9) has examined some of the problems associated with heat and mass-transfer in two-phase flow. In this paper, he refers to the Friedel correlation which is widely used for calculating the two-phase pressure drop in the nuclear industry. The literature cited so far deals with flow through relatively short tubes and pipes. The acceleration effect on pressure drop for flow through short pipes is considerably greater than for flow through long lines. The present study deals with the pressure drop encountered in long transmission lines, in which the acceleration effect is negligible. Regardless of the type of application, the publications in gas-liquid flow cover such a variety of methods and approaches that reviewing all of them would be impractical. Those reviewed in this chapter are among the more widely used methods for flow in long horizontal transmission lines. Inclusion or exclusion of an approach has no significance about its worth for a specific application.

The Lockhart-Martinelli Method

Lockhart and Martinelli (30) presented the first method to predict the pressure drop for gas-liquid flow in pipes. They obtained an empirical correlation from experimental data for air and various liquids in pipes ranging from 0.586 inches to 1.017 inches in diameter. Although the correlation is based on data taken in small diameter pipes, it has been used extensively in industry for a wide variety of systems with moderate success (34). One correlating parameter in this method is the square root of the ratio of the pressure drops obtained if each phase occupied the entire pipe. The parameter is

$$X = [(dP/dZ)_{\ell}/(dP/dZ)_{g}]^{1/2} \qquad \text{(II-1)}$$

The pressure loss terms in Equation II-1 are separately calculated assuming that each phase occupies the entire pipe. The parameter X is then used to find a multiplying factor which can be used to calculate the two phase pressure drop from the single phase pressure drop:

$$\left(\frac{dP}{dZ}\right)_{tp} = \left(\frac{dP}{dZ}\right)_{g} \phi_g^2 \qquad \text{(II-2)}$$

$$\left(\frac{dP}{dZ}\right)_{tp} = \left(\frac{dP}{dZ}\right)_{\ell} \phi_\ell^2 \qquad \text{(II-3)}$$

The functions ϕ_g amd ϕ_ℓ were presented in graphical form as shown in Figure 1, page 7. Lockhart and Martinelli identified four flow regimes which they defined as gas turbulent-liquid turbulent, gas turbulent-liquid viscous, gas viscous-liquid turbulent, and gas viscous-liquid viscous flow. They presented a separate curve relating ϕ_g and ϕ_ℓ with X for each flow regime. Degance and Atherton (17) developed equations representing the functional relationships for computer application. Chisholm (13) recommended simplified equations to calculate the Lockhart-Martinelli multiplying factor ϕ_ℓ. The equations are of the form

$$\phi_\ell^2 = 1 + C/X + 1/x^2 \qquad \text{(II-4)}$$

where C = a constant whose value depends upon the flow regime

X = the Lockhart-Martinelli parameter

x = the ratio of the gas mass flowrate to the total mass flowrate.

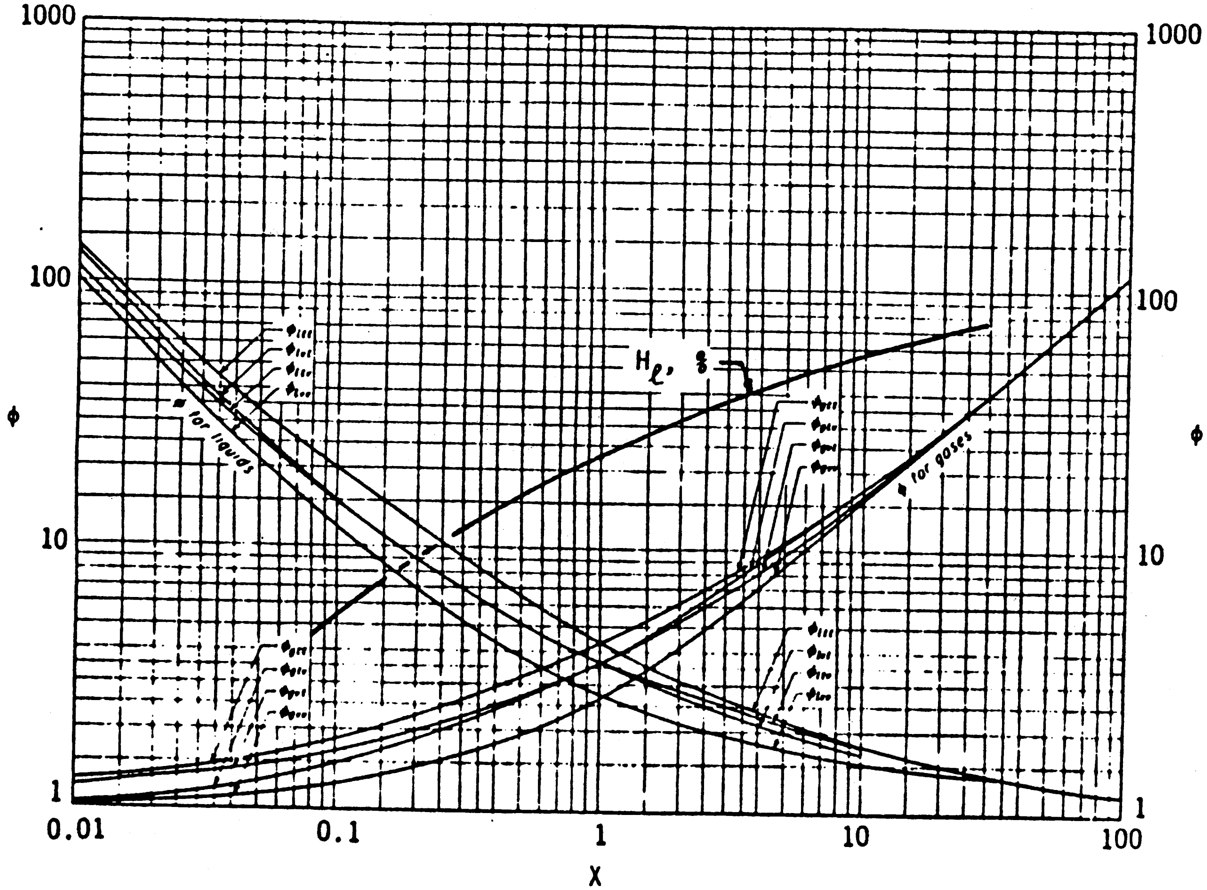

Figure 1. The Relationship Between the Lockhart and Martinelli Correlating Parameters

Source: Boyde (9)

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According to Chisholm (13), the values of ϕ_{ℓ} calculated from Equation II-4 are in good agreement with the empirical curves of Lockhart-Martinelli.

Baker (3) has shown that the Lockhart-Martinelli method over-predicts the pressure drop in large diameter pipes. He developed a flow regime map showing seven different flow patterns. Baker (4) recommended the Flanigan (22) correlation for calculating the pressure drop in inclined lines. Figure 2 shows Baker's flow regime map. The parameters in this map are defined as follows:

L = superficial liquid mass velocity, lb/ft^2 sec

G = superficial gas mass velocity, lb/ft^2 sec

$\lambda' = [(\rho_g/0.075)(\rho_L/62.3)]^{1/2}$

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

ρ_g = gas density, lb/ft^3

ρ_L = liquid density, lb/ft^3

μ_L = liquid viscosity, centipoise

σ = surface tension, dynes/cm

Baker presented a separate equation for each type of flow to calculate the Lockhart-Martinelli multiplier ϕ_g. The equations are of the form

$$\phi_g = \frac{a\ X^b}{(L')^c} \qquad \text{(II-5)}$$

where a, b, and c are empirically-determined constants, X the Lockhart-Martinelli correlating parameter, and L' the line length in feet. The two-phase pressure drop is then calculated from Equation II-2.

The constant c in Equation II-5 is a positive number. Since the other terms in the equation are independent of line length, Equation

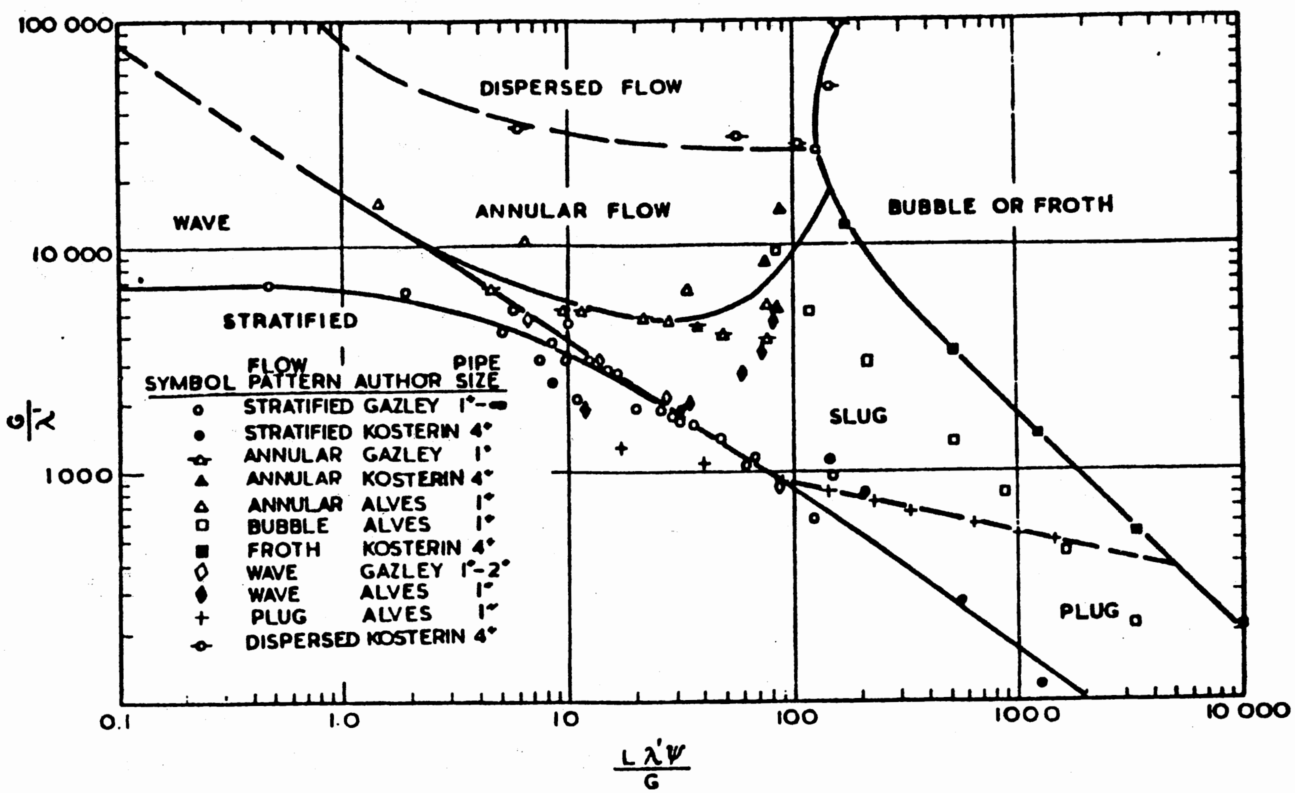

Figure 2. Flow Pattern Regions in Gas-Liquid Flow

Source: Baker (3)

II-5 implies that for very long pipelines, ϕ_g will approach a value close to zero. This does not seem reasonable. Baker based his correlations on experimental data from 8 inch and 10 inch diameter pipelines. According to Husain, Choe, and Wiseman (28), his correlations do not work well for smaller diameter pipes, except for air-water systems.

Besides Baker, Hoogendoorn (25) and Hoogendoorn and Buitelaar (26) also investigated gas-liquid flow patterns in horizontal pipelines. Hoogendoorn (25) measured liquid holdup in smooth pipes with diameters ranging from 24 mm to 140 mm. He used air as the gas and water, spindle oil, and gas oil as liquids in his experiments. Hoogendoorn and Buitelaar (26) then investigated the flow of superheated Refrigerant-11 in horizontal 15 mm pipes. The results of these studies show that the Lockhart-Martinelli method does not work well for wave flow and mist-annular flow. For wave flow, Hoogendoorn proposed the correlation:

$$(\Delta P)_{tp} = \frac{C'}{Z}\,\frac{L'}{D}\,\frac{G_m^2}{\rho_g}\left(\frac{G}{L}\right)^{1.45} \qquad \text{(II-6)}$$

where, C' = a constant which depends on the pipe diameter and roughness

G_m = the mixture mass velocity, kg/m^2 sec

D = the pipe diameter, m

G = superficial gas mass velocity, kg/m^2 sec

L = superficial liquid mass velocity, kg/m^2 sec

$(\Delta P)_{tp}$ = pressure drop, N/m^2

The quantity G_m is evaluated from

$$G_m = \frac{\text{total mass flow rate}}{\text{pipe cross-section}}$$

For mist-annular flow, Hoogendoorn and Buitelaar proposed the correlation:

$$(\Delta P)_{tp} = \frac{1}{2} \frac{L'}{D} \frac{G^2}{\rho_g} = 0.12(G)^{-1/4} \qquad \text{(II-7)}$$

where the terms and the units for the terms are the same as described for Equation II-6.

The Beggs and Brill Method

The Beggs and Brill method (8) is another method based on flow regimes. The general equation for pressure drop is

$$\left(\frac{dP}{dZ}\right)_{total} = \left(\frac{dP}{dZ}\right)_{elev.} + \left(\frac{dP}{dZ}\right)_{fric.} + \left(\frac{dP}{dZ}\right)_{accel.} \qquad \text{(II-8)}$$

The three pressure loss terms in the Beggs and Brill method are:

$$\left(\frac{dP}{dZ}\right)_{elev.} = \frac{g}{g_c} \sin\theta \,[\rho_L H_L + \rho_g (1 - H_L)] \qquad \text{(II-9)}$$

$$\left(\frac{dP}{dZ}\right)_{fric.} = \frac{f_{tp}\, \rho_{tp}\, V_m^2}{2\, g_c\, D} \qquad \text{(II-10)}$$

$$\left(\frac{dP}{dZ}\right)_{accel.} = - \frac{[\rho_L H_L + \rho_g (1 - H_L)]\, V_m\, V_{sg}}{g_c\, P} \frac{dP}{dZ} \qquad \text{(II-11)}$$

where, H_L = the volume of liquid in the line, expressed as percent of total line volume. This parameter is known as the liquid holdup.

V_m = the mixture velocity, ft/sec

V_{sg} = the superficial gas velocity, ft/sec

ρ_{tp} = the two-phase density, lb/ft^3

f_{tp} = the two-phase friction factor

The mixture velocity is defined as the velocity of the total flowing mixture, and is calculated as

$$V_m = \frac{q_g + q_L}{A} \qquad \text{(II-12)}$$

where, q_g = gas flow rate, ft^3/sec

q_L = liquid flow rate, ft^3/sec

A = pipe cross-section, ft^2

The two phase density is calculated from

$$\rho_{tp} = \rho_L H_L + \rho_g (1 - H_L) \qquad \text{(II-13)}$$

The superficial gas velocity is defined as

$$V_{sg} = q_g/A \qquad \text{(II-14)}$$

The two-phase friction factor, f_{tp}, is calculated from the no-slip friction factor f_{ns} using the equation

$$f_{tp}/f_{ns} = e^S \qquad \text{(II-15)}$$

where

$$S = [\ln (Y)]/\{-0.0523 + 3.182 \ln(Y) - 0.8725 [\ln(Y)]^2$$

$$+ 0.01853 [\ln(Y)]^4\} \qquad \text{(II-16)}$$

$$Y = \frac{\lambda}{(H_L)^2} \tag{II-17}$$

$$\lambda = \text{volumetric liquid fraction} = \frac{q_L}{q_L + q_g} \tag{II-18}$$

The no-slip friction factor, f_{ns}, is obtained from the Moody diagram (36) using

$$Re = \frac{V_m \, [\rho_L \lambda + \rho_g \, (1 - \lambda)] \, D}{\mu_\ell \, \lambda + \mu_g \, (1 - \lambda)} \tag{II-19}$$

In Equation II-19, the gas-liquid mixture is treated as a pseudo single phase whose physical properties can be approximated using a volumetric average.

The acceleration pressure loss (Equation II-11) contains a $(\frac{dP}{dZ})$ term. When the three pressure loss terms represented by Equations II-9, II-10, and II-11 are substituted in Equation II-8, and rearranged to solve for the pressure gradient, the final form of the Beggs and Brill equation becomes

$$-\left(\frac{dP}{dZ}\right) = \frac{\frac{g}{g_c} \sin\theta \, [\rho_L \, H_L + \rho_g \, (1 - H_L)] + \frac{f_{tp} \, V_m^2 \, \rho_{tp}}{2 \, g_c \, D}}{1 - \frac{[\rho_L \, H_L + \rho_g \, (1 - H_L)] \, V_m \, V_{sg}}{g_c \, P}} \tag{II-20}$$

The elevation term reduces to zero for horizontal pipes. The friction loss term contains the square of the mixture velocity. The greater the mixture velocity, the greater the pressure drop. The two-phase density, ρ_{tp}, and the two-phase friction factor, f_{tp}, require knowledge of the liquid holdup.

Beggs and Brill separated their experimental data into three different flow regimes and defined them as segregated, intermittent, and distributed flow. For each flow regime, they proposed a separate equation for liquid holdup. The equations are of the form

$$H_L = a\lambda^{\alpha} N_{FR}^{\beta} \qquad \text{(II-21)}$$

where

$$N_{FR} = V_m^2/gd \qquad \text{(II-22)}$$

a, α, and β are empirically-determined constants.

A major drawback with Equation II-15 is that for a given value of λ (from 0.001 to 1.0) the calculated two-phase friction factor decreases as the holdup increases from 0.05 to 0.90. This does not seem reasonable. The denominator in the friction loss term is constant, while the two terms ρ_{tp} and f_{tp} depend upon the holdup. For horizontal pipes, the pressure drop calculated using this method is influenced by the mixture velocity and the liquid holdup.

Danesh (16) criticized the Beggs and Brill holdup correlation, and reported negative values and values greater than unity for data on a gas-condensate pipeline. He concluded that since the correlations were derived from air-water data, the method over-predicts the horizontal holdup for high pressure gas-condensate pipelines. However, the results obtained in the present study do not agree with these conclusions. Table I shows the pressure drop calculated by the Beggs and Brill method using superficial phase velocities in the calculation procedure. With

TABLE I

CALCULATED PRESSURE DROP FOR THE BEGGS AND BRILL PROCEDURE USING SUPERFICIAL PHASE VELOCITIES

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
3	7.75	65	2.16	192	11.68	726.7	0.36	5.8	9.0	4.1	-54.4
	7.75	70	2.01	107	6.67	1081.7	0.51	7.1	1.5	0.5	-66.7
	7.75	70	2.14	236	11.95	1077.7	0.62	7.1	7.5	3.2	-57.3
	7.75	69	2.14	244	9.48	1088.7	0.78	9.8	7.0	2.1	-69.3
	10.14	69	7.83	244	9.48	1082.7	0.82	31.0	10.0	2.1	-79.0
	4.03	79	0.69	627	7.47	1101.7	2.50	13.1	20.0	14.3	-28.5
	5.94	72	2.01	627	7.47	1084.7	2.53	13.6	15.0	5.8	-61.3
	10.14	78	7.83	6592	11.89	974.7	15.25	41.2	24.0	13.6	-43.3
	10.14	69	7.83	4167	12.05	976.7	20.87	53.3	16.0	5.8	-63.7
	7.75	66	2.14	4970	6.47	978.7	21.20	21.0	6.0	2.8	-53.3
	10.14	82	7.83	5420	4.35	930.0	26.91	62.8	18.0	4.4	-75.5
	7.75	82	2.14	5420	4.35	954.7	27.20	11.0	10.0	4.5	-55.0
	7.75	75	2.14	514	25.97	997.7	0.60	6.9	19.0	19.6	+3.1
	7.75	80	2.14	5484	25.52	1021.7	6.55	21.9	32.0	33.6	+5.0
4	12.00	60	25.45	6912	3.19	424.7	34.50	54.9	105.0	15.5	-85.2
5	4.03	60	0.19	40	0.4	50.0	0.20	3.6.	0.8	0.5	-37.5
	4.03	60	0.19	800	8.0	400.0	2.03	8.0	128.7	21.5	-83.3
	4.03	60	0.19	500	5.0	400.0	2.03	10.6	22.3	7.9	-64.6
	4.03	60	0.19	120	1.2	400.0	2.03	13.7	1.5	0.4	-73.3
	4.03	60	0.19	1000	2.0	1000.0	2.03	43.3	4.1	1.3	-68.3
	4.03	60	0.19	800	0.4	50.0	4.03	15.5	5.6	3.7	-33.9
	4.03	60	0.19	400	0.2	50.0	4.03	17.8	3.6	1.0	-72.2
	4.03	60	0.19	200	0.1	50.0	4.03	20.2	2.4	0.2	-91.7
	4.03	60	0.19	100	0.1	50.0	4.03	22.8	0.6	0.1	-88.3
	4.03	60	0.19	2500	5.0	400.0	4.03	22.3	43.2	19.6	-54.6
	4.03	60	0.19	1200	12.0	1000.0	7.01	19.7	118.4	15.5	-86.9
	4.03	60	0.19	800	8.0	1000.0	7.01	19.3	22.8	7.1	-68.8
	4.03	60	0.19	500	5.0	1000.0	7.02	23.7	9.2	2.8	-69.6
	4.03	60	0.19	200	2.0	1000.0	7.02	27.9	1.6	0.4	-75.0
	4.03	60	0.19	1500	3.0	400.0	9.65	22.9	15.7	7.2	-54.1
	4.03	60	0.19	300	3.0	400.0	28.60	11.7	8.4	2.9	-65.5
	4.03	60	0.19	600	1.2	1000.0	28.60	58.6	2.9	0.5	-82.7
	4.03	60	1.19	200	0.4	1000.0	28.60	71.7	1.4	0.1	-92.8
	4.03	60	0.19	4000	8.0	1000.0	28.80	41.2	67.6	18.8	-72.2
	4.03	60	0.19	2500	5.0	1000.0	28.80	42.0	31.9	7.6	-76.2
	4.03	60	0.19	1600	0.8	400.0	34.70	47.7	4.7	2.0	-57.4
	4.03	60	1.19	800	0.4	50.0	34.70	65.1	3.0	0.5	-83.3
	4.03	60	0.19	400	0.2	50.0	34.70	48.1	1.9	0.3	-84.2
	4.03	60	1.19	2400	1.2	400.0	34.80	47.0	20.7	4.2	-79.7
	4.03	60	0.19	2400	1.2	1000.0	79.70	55.0	12.3	2.2	-82.1

TABLE I (Continued)

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
5	4.03	60	0.19	4000	2.0	1000.0	79.80	47.0	27.3	5.6	-79.5
	4.03	60	0.19	1600	0.8	1000.0	79.80	60.4	6.6	1.0	-84.8
	4.03	60	0.19	800	0.4	1000.0	79.80	73.4	1.9	0.1	-94.7
41	3.00	90	0.19	158	1.8	494.7	1.90	--	4.7	--	--
	3.00	90	0.19	169	1.7	472.7	2.60	13.2	30.0	3.5	-88.3
	3.00	90	0.19	169	1.4	468.7	2.70	13.7	8.2	2.6	-68.3
	3.00	90	0.19	283	1.4	473.7	4.30	13.7	8.6	3.5	-59.2
	3.00	90	0.19	375	1.7	466.0	4.60	17.0	10.0	5.5	-45.0
	3.00	90	0.19	734	0.7	473.1	21.30	36.2	8.4	3.1	-63.1
	3.00	90	0.19	788	0.6	471.1	24.70	39.2	6.4	3.0	-53.0
	3.00	91	0.19	157	0.1	466.1	31.40	71.7	1.4	0.1	-92.9
	3.00	90	0.19	655	0.4	468.5	31.90	45.8	3.8	1.6	-57.9
	3.00	90	0.19	186	1.3	466.2	32.90	14.4	4.0	2.7	-32.5
	3.00	90	0.19	143	0.1	466.0	37.80	81.7	1.5	0.1	93.3
	3.00	90	0.19	171	0.1	466.0	32.40	--	1.3	--	--
	3.00	90	0.19	671	0.4	468.3	30.30	--	3.6	--	--
40	2.00	65	0.02	10	0.1	50.0	0.12	2.4	0.9	0.3	-66.7
	2.00	65	0.02	24	0.1	50.0	0.30	2.7	1.2	0.3	-75.0
	2.00	65	0.02	65	0.1	50.0	1.49	9.4	0.4	0.2	-50.0

* The calculated pressure drops have been rounded off.

**The errors have not been rounded off. A small difference is likely to occur between the reported error and the error calculated from reported values.

the exception of the last two cases from Baker (3), the calculated pressure drop is lower than the observed pressure drop for all cases.

The AGA Method

The American Gas Association (19) and the American Petroleum Institute (6) made a major contribution to the study of gas-liquid flow by funding a research project in which a wide range of experimental data was collected from the literature, and evaluated for accuracy and reliability. Existing correlations were then tested against evaluated data. The correlations which fit the data provided a starting point for developing an improved method (now known as the AGA method) for predicting the two-phase pressure drop. The AGA method represents a different approach in that it uses no flow regime map. The liquid holdup is a function of the volume fraction input liquid and the two-phase Reynolds number. Figure 3, as reproduced from Akashah (1), shows this relationship. The three pressure loss terms in the AGA pressure drop equation are

$$(\Delta P)_{elev.} = \frac{\Sigma h \; F_e \; \rho_L}{144} \tag{II-23}$$

where, Σh = algebraic sum of elevation changes, ft

F_e = the Flanigan elevation factor

$$(\Delta P)_{fric.} = \frac{2 \; f_{tp} \; L' \; V_m^2 \; \rho_{tp}}{144 \; \rho_g \; D} \tag{II-24}$$

$$(\Delta P)_{accel.} = \frac{1}{144 \; g_c}\{[\frac{\rho_g \; V_{sg}^2}{1 - H_L} + \frac{\rho_L \; V_{sL}^2}{H_L}]_{downstream}$$

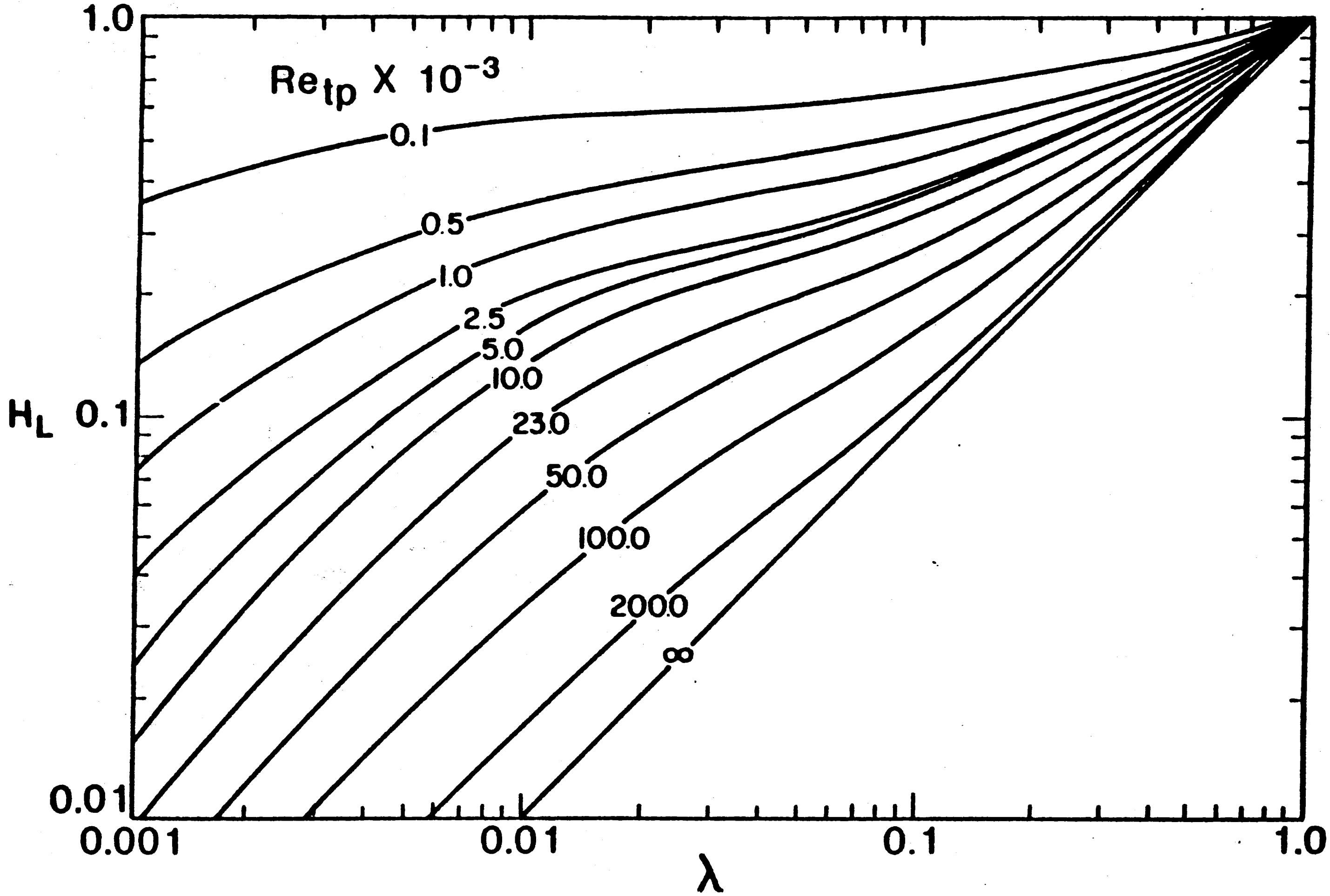

Figure 3. The AGA Holdup Correlation

Source: Akashah (1)

$$- \left[\frac{\rho_g V_{sg}^2}{1 - H_L} + \frac{\rho_L V_{sL}^2}{H_L}\right]_{upstream} \cos\theta\} \qquad \text{(II-25)}$$

The two-phase density is calculated from the equation

$$\rho'_{tp} = \frac{\rho_L \lambda^2}{H_L} + \frac{\rho_g (1 - \lambda)^2}{(1 - H_L)} \qquad \text{(II-26)}$$

The terms V_{sg}, V_m, λ, and H_L refer to the same variables defined during the discussion on the Beggs and Brill method. The superficial liquid velocity V_{sL} is defined as:

$$V_{sL} = \frac{q_L}{A} \qquad \text{(II-27)}$$

The total pressure drop is the algebraic sum of the three pressure loss terms represented by Equations II-23, II-24, and II-25. As before, the elevation and the acceleration terms may be neglected for long lines. The denominator in Equation II-24 is constant. So, the friction loss term depends only on the holdup and the mixture velocity. Table II shows the pressure drops calculated with the AGA method using superficial phase velocities. With the exception of the two cases from Baker (3), all the calculated pressure drops are lower than the observed pressure drops.

Eaton et al. (20) conducted an experimental investigation of gas-liquid flow in 2 and 4 inch diameter lines. They used natural gas as the gas and water, crude oil, and distillate as the liquid. Eaton et al. attempted to study the effect of changes in flow patterns on the pressure drop and develop separate correlations for each pattern if needed. They concluded that many of the variables which controlled the

TABLE II

CALCULATED PRESSURE DROP FOR THE AGA METHOD USING SUPERFICIAL PHASE VELOCITIES

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
3	7.75	65	2.16	192	11.68	726.7	0.36	1.6	9.0	7.1	-21.1
	7.75	70	2.01	107	6.67	1081.7	0.51	--	1.5	--	--
	7.75	70	2.14	236	11.95	1077.7	0.62	1.6	7.5	5.1	-32.0
	7.75	69	2.14	244	9.48	1088.7	0.78	1.4	7.0	3.5	-50.0
	10.14	69	7.83	244	9.48	1082.7	0.82	29.8	10.0	6.3	-37.0
	4.03	79	0.69	627	7.47	1101.7	2.50	2.5	20.0	20.0	0.0
	5.94	72	2.01	627	7.47	1084.7	2.53	2.6	15.0	12.6	-16.1
	10.14	78	7.83	6592	11.89	974.7	14.96	16.0	24.0	21.5	-10.4
	10.14	69	7.83	4167	12.05	976.7	20.87	21.5	16.0	9.0	-43.4
	7.75	66	2.14	4970	6.47	978.7	21.20	23.0	6.0	4.2	-25.8
	10.14	82	7.83	5420	4.35	930.0	26.91	29.4	18.0	6.4	-64.3
	7.75	82	2.14	5420	4.35	954.7	27.20	1.4	10.0	3.5	-65.0
	7.75	75	2.14	514	25.97	997.7	0.60	1.5	19.0	27.1	+42.6
	7.75	80	2.14	5484	25.52	1021.7	6.55	6.4	32.0	49.1	+53.4
4	12.00	60	25.45	69.12	3.19	424.7	34.50	41.6	105.0	19.8	-81.1
5	4.03	60	0.19	40	0.40	50.0	0.20	0.1	0.8	0.5	-37.5
	4.03	60	0.19	800	8.00	400.0	2.03	7.2	128.7	26.2	-79.6
	4.03	60	0.19	500	5.00	400.0	2.03	2.0	22.3	11.0	-50.7
	4.03	60	0.19	120	1.20	400.0	2.03	2.0	1.5	0.8	-46.7
	4.03	60	0.19	1000	2.00	1000.0	2.03	32.5	4.1	1.6	-61.0
	4.03	60	0.19	800	0.40	50.0	4.03	3.9	5.6	5.5	-1.8
	4.03	60	0.19	400	0.20	50.0	4.03	3.6	3.6	0.7	-86.1
	4.03	60	0.19	200	0.10	50.0	4.03	13.5	2.4	0.2	-91.7
	4.03	60	0.19	100	0.05	50.0	4.03	20.0	0.6	0.0	-91.7
	4.03	60	0.19	2500	5.00	400.0	4.03	9.6	43.2	25.6	-40.7
	4.03	60	0.19	1200	12.00	1000.0	7.01	--	118.4	--	--
	4.03	60	0.19	800	8.00	1000.0	7.01	2.0	22.8	9.7	-57.4
	4.03	60	0.19	500	5.00	1000.0	7.02	7.3	9.2	4.1	-55.4
	4.03	60	0.19	200	2.00	1000.0	7.02	8.4	1.6	0.7	-57.4
	4.03	60	0.19	1500	3.00	400.0	9.65	9.8	15.7	9.9	-36.9
	4.03	60	0.19	300	3.00	400.0	28.60	2.0	8.4	2.9	-65.5
	4.03	60	0.19	600	1.20	1000.0	28.60	37.6	2.9	0.6	-79.3
	4.03	60	0.19	200	0.40	1000.0	28.60	46.8	1.4	0.1	-94.3
	4.03	60	0.19	4000	8.00	1000.0	28.80	31.2	67.6	20.9	-69.1
	4.03	60	0.19	2500	5.00	1000.0	28.80	31.4	31.9	8.8	-72.4
	4.03	60	0.19	1600	0.80	400.0	34.70	38.5	4.7	2.2	-53.2
	4.03	60	0.19	800	0.40	50.0	34.70	45.2	3.0	0.6	-80.0
	4.03	60	0.19	400	0.20	50.0	34.70	50.8	1.9	0.2	-89.5
	4.03	60	0.19	2400	1.20	400.0	34.80	37.7	20.7	4.7	-77.3

TABLE II (Continued)

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
5	4.03	60	0.19	2400	1.2	1000.0	79.7	84.8	12.3	1.7	-86.2
	4.03	60	0.19	4000	2.0	1000.0	79.8	81.7	27.3	4.2	-84.6
	4.03	60	0.19	1600	0.8	1000.0	79.8	84.2	6.6	0.8	-87.9
	4.03	60	0.19	800	0.4	1000.0	79.8	86.3	1.9	0.2	-89.5
41	3.00	90	0.19	157	1.8	494.7	1.9	2.0	4.7	3.7	-21.3
	3.00	90	0.19	169	1.7	494.7	2.6	2.6	30.0	5.1	-83.0
	3.00	90	0.19	169	1.4	472.7	2.7	2.7	8.2	3.9	-52.4
	3.00	90	0.19	283	1.4	468.7	4.3	4.3	8.6	5.2	-39.5
	3.00	90	0.19	375	1.7	473.7	4.6	4.6	10.0	7.9	-21.2
	3.00	90	0.19	734	0.7	466.0	21.3	23.6	8.4	3.9	-53.6
	3.00	90	0.19	788	0.6	473.1	24.7	27.8	6.4	3.7	-42.2
	3.00	90	0.19	157	0.1	471.1	31.4	51.9	1.4	0.1	-92.2
	3.00	90	0.19	655	0.4	466.1	31.9	40.4	3.8	1.8	-53.1
	3.00	90	0.19	186	1.3	468.5	32.9	3.0	4.0	4.0	-1.3
	3.00	90	0.19	143	0.1	466.2	37.8	58.8	1.5	0.1	-93.3
	3.00	90	0.19	171	0.1	466.0	32.4	71.4	1.3	0.2	-88.1
	3.00	90	0.19	671	0.4	468.3	30.3	--	3.6	--	--
40	2.00	65	0.02	10	0.1	0.16	50.0	1.5	0.9	0.3	-71.1
	2.00	65	0.02	24	0.1	0.17	50.0	1.4	1.2	0.3	-72.7
	2.00	65	0.02	65	0.1	0.12	50.0	6.0	0.4	0.2	-42.8

* The calculated pressure drops have been rounded off.
**The errors have not been rounded off. A small difference is likely to occur between the reported error and the error calculated from reported values.

flow pattern also controlled the pressure drop in horizontal lines. They proposed that a single correlation for the liquid holdup should be adequate for all flow regimes. The correlating function for the holdup is:

$$H_L = \Psi' \left[\frac{N_{LV}^{0.575}}{N_{gv}\ N_d^{0.0277}} \left(\frac{P}{P_b}\right)^{0.05} \left(\frac{N_L}{N_{LB}}\right)^{0.10}\right] \tag{II-28}$$

where

$$N_{LV} = 1.938\ V_{sL}\ (\rho_L/\sigma)^{0.25} \tag{II-29}$$

$$N_{gv} = 1.938\ V_{sg}\ (\rho_g/\sigma)^{0.25} \tag{II-30}$$

$$N_d = 120.872\ d\ (\rho_L/\sigma)^{0.25} \tag{II-31}$$

$$P/P_b = P/14.65 \tag{II-32}$$

$$N_L = 0.15726\ \mu_L\ (1/\rho_L\ \sigma^3)^{0.25}$$

$$N_{LB} = 0.00226 \text{ (based on water)} \tag{II-33}$$

In the above equations, the units of the individual quantities are:

σ = surface tension, dynes/cm

μ_L = liquid viscosity, centipoise

d = pipe diameter, ft

ρ_L, ρ_g = density, lb/ft^3

V_{SL}, V_{sg} = superficial velocity, ft/sec

P = pressure psia

The last term in Equation II-28 includes the viscous effects on the holdup function. The value of $(\frac{N_L}{N_{Lb}})^{0.1}$ is always greater than unity. According to Eaton et al., this term makes the liquid holdup estimated from Equation II-28 generally greater than the holdup calculated by other methods.

Some investigators (7,17) have suggested that a reasonably good estimate of the pressure drop can be obtained when the Eaton et al. correlation is used to calculate the liquid holdup. Tables III and IV show the results with the Eaton et al. correlation. The calculated pressure drops are in some cases higher than those reported in Tables I and II. However, they are lower than the observed pressure drops in all but the last two cases from Baker (3). This is because the mixture velocity controls the calculated two-phase pressure drop. The calculation procedure assuming no-slip flow does not include a logical way of increasing the mixture velocity.

Battara et al. (7), Mandhane et al. (33,34), and Degance and Atherton (17) provide a critical evaluation of some of the widely used holdup correlations and pressure drop calculation methods in two-phase flow. Battara et al. and Mandhane et al. concluded that no single pressure drop calculation method is suitable for all operating conditions. Mandhane et al. recommended different methods for different flow regimes. The order of preference among the recommended methods varied with the type of flow regime map used. Degance and Atherton concluded that the constant slip method of Dukler et al. (18) was the most accurate among the currently available methods.

TABLE III

CALCULATED PRESSURE DROP FOR THE BEGGS AND BRILL PROCEDURE USING SUPERFICIAL PHASE VELOCITIES AND THE EATON ET AL. HOLDUP CORRELATION

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
3	7.75	65	2.16	192	11.68	726.7	0.36	2.8	9.0	6.3	-29.9
	7.75	70	2.01	107	6.67	1081.7	0.51	6.8	1.5	1.0	-33.3
	7.75	70	2.14	236	11.95	1077.7	0.62	5.6	7.5	3.2	-57.3
	7.75	69	2.14	244	9.48	1088.7	0.78	7.9	7.0	2.4	-65.7
	10.14	69	7.83	244	9.48	1082.7	0.82	9.9	10.0	2.1	-79.0
	4.03	79	0.69	627	7.47	1101.7	2.50	10.1	20.0	15.2	-24.0
	5.94	72	2.01	627	7.47	1084.7	2.53	14.0	15.0	8.1	-49.4
	10.14	78	7.83	6592	11.89	974.7	14.96	34.3	24.0	14.9	-37.7
	10.14	69	7.83	4167	12.05	976.7	20.87	45.0	16.0	4.5	-71.9
	7.75	66	2.14	4970	6.47	978.7	21.20	40.6	6.0	2.8	-48.3
	10.14	82	7.83	5420	4.35	930.0	26.91	52.5	18.0	4.4	-75.5
	7.75	82	2.14	5420	4.35	954.7	27.20	47.8	10.0	4.8	-52.4
	7.75	75	2.14	514	26.97	997.7	0.60	3.7	19.0	21.0	+10.5
	7.75	80	2.14	5484	25.52	1021.7	6.55	15.9	32.0	34.5	+7.8
4	12.00	60	25.45	6912	3.19	424.7	34.50	61.3	105.0	12.4	-88.1
5	4.03	60	0.19	40	0.40	50.0	0.20	0.2	0.8	0.6	-25.0
	4.03	60	0.19	800	8.00	400.0	2.03	4.6	128.7	28.0	-78.2
	4.03	60	0.19	500	5.00	400.0	2.03	14.8	22.3	14.8	-33.6
	4.03	60	0.19	120	1.20	400.0	2.03	11.9	1.5	0.9	-40.0
	4.03	60	0.49	1000	2.00	1000.0	2.03	25.8	4.1	1.2	-70.2
	4.03	60	0.19	800	0.40	50.0	4.03	10.0	5.6	4.2	-25.0
	4.03	60	0.19	400	0.20	50.0	4.03	13.7	3.6	1.0	-72.2
	4.03	60	0.49	200	0.10	50.0	4.03	17.7	2.4	0.3	-87.5
	4.03	60	0.19	100	0.05	50.0	4.03	22.3	0.6	0.1	-88.3
	4.03	60	0.19	2500	5.00	400.0	4.03	16.1	43.2	21.6	-49.9
	4.03	60	0.19	1200	12.00	1000.0	7.01	15.1	118.4	30.7	-74.1
	4.03	60	0.19	800	8.00	1000.0	7.01	17.7	22.8	7.3	-68.0
	4.03	60	0.19	500	5.00	1000.0	7.02	27.7	9.2	5.4	-40.9
	4.03	60	0.19	200	2.00	1000.0	7.02	27.1	1.6	0.4	-75.0
	4.03	60	0.19	1500	3.00	400.0	9.65	19.5	15.7	7.5	-52.3
	4.03	60	0.19	300	3.00	400.0	28.60	11.9	8.4	5.2	-38.8
	4.03	60	0.19	600	1.20	1000.0	28.60	53.3	2.9	0.4	-53.3
	4.03	60	0.19	200	0.40	1000.0	28.60	63.5	1.4	0.1	-92.8
	4.03	60	0.19	4000	8.00	1000.0	28.80	35.4	67.6	19.5	-71.1
	4.03	60	0.19	2500	5.00	1000.0	28.80	39.6	31.9	7.7	-76.7
	4.03	60	0.19	1600	0.80	400.0	34.70	51.0	4.7	2.0	-57.4
	4.03	60	0.19	800	0.40	50.0	34.70	57.7	3.0	0.5	-86.7
	4.03	60	0.19	400	0.20	50.0	34.70	63.9	1.9	0.1	-94.7

TABLE III (Continued)

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
5	4.03	60	0.19	2400	1.2	400.0	34.8	46.9	20.7	4.2	-79.7
	4.03	60	0.19	2400	1.2	1000.0	79.7	84.8	12.3	1.9	-84.5
	4.03	60	0.19	4000	2.0	1000.0	79.8	81.5	27.3	5.5	-79.8
	4.03	60	0.19	1600	0.8	1000.0	79.8	86.8	6.6	1.0	-87.8
	4.03	60	0.19	800	0.4	1000.0	79.8	90.2	1.9	0.2	-89.5
41	3.00	90	0.19	158	1.8	494.7	1.9	7.7	4.7	3.9	-16.2
	3.00	90	0.19	169	1.7	472.7	2.6	10.0	30.0	5.4	-81.9
	3.00	90	0.19	169	1.4	468.7	2.7	10.8	8.2	2.8	-66.3
	3.00	90	0.19	283	1.4	473.7	4.3	13.8	8.6	3.7	-59.2
	3.00	90	0.19	375	1.7	466.0	4.6	13.4	10.0	5.5	-45.5
	3.00	90	0.19	734	0.7	473.1	21.3	37.8	8.4	3.1	-63.1
	3.00	90	0.19	788	0.6	471.1	24.7	41.2	6.4	3.0	-53.0
	3.00	90	0.19	157	0.1	466.1	31.4	64.4	1.4	0.1	-92.9
	3.00	90	0.19	655	0.4	468.5	31.9	51.3	3.8	1.6	-57.9
	3.00	90	0.19	186	1.3	466.2	32.9	11.5	4.0	2.4	-32.1
	3.00	90	0.19	143	0.1	466.0	37.8	70.3	1.5	0.1	-93.3
	3.00	90	0.19	171	0.1	466.0	32.4	64.1	1.3	0.1	-92.3
	3.00	90	0.19	671	0.4	468.3	30.3	49.0	3.6	1.9	-47.8
40	2.00	65	0.02	10	0.1	50.0	0.12	0.1	0.9	0.3	-66.7
	2.00	65	0.02	24	0.1	50.0	0.30	0.8	1.25	0.7	-41.3
	2.00	65	0.02	65	0.1	50.0	1.49	5.8	0.4	0.3	-25.0

* The calculated pressure drops have been rounded off.
**The errors have not been rounded off. A small difference is likely to occur between the reported error and the error calculated from reported values.

TABLE IV

CALCULATED PRESSURE DROP FOR THE AGA METHOD USING SUPERFICIAL PHASE VELOCITIES AND USING THE EATON ET AL. HOLDUP CORRELATION

Data Source Ref. #	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
3	7.75	65	2.16	192	11.68	726.7	0.36	2.8	9.0	7.1	-22.0
	7.75	70	2.01	107	6.67	1081.7	0.51	--	1.5	--	--
	7.75	70	2.14	236	11.95	1077.7	0.62	5.6	7.5	5.4	-28.0
	7.75	69	2.14	244	9.48	1088.7	0.78	7.9	7.0	3.5	-50.0
	10.14	69	7.83	244	9.48	1082.7	0.82	47.8	10.0	5.0	-50.0
	4.03	79	0.69	627	7.47	1101.7	2.50	10.1	20.0	18.6	-7.0
	5.94	72	2.01	627	7.47	1084.7	2.53	14.0	15.0	11.7	-24.6
	10.14	78	7.83	6592	11.89	974.7	14.96	34.3	24.0	9.3	-61.2
	10.14	69	7.83	4167	12.05	976.7	20.87	45.0	16.0	6.6	-58.7
	7.75	66	2.14	4970	6.47	978.7	21.20	40.5	6.0	3.3	-44.3
	10.14	82	7.83	5420	4.35	930.0	26.91	52.5	18.0	6.8	-57.5
	7.75	82	2.14	5420	4.35	954.7	27.20	9.8	10.0	3.6	-64.0
	7.75	75	2.14	514	26.97	997.7	0.60	3.7	19.0	27.0	+42.1
	7.75	80	2.14	5484	25.52	1021.7	6.55	15.8	32.0	42.0	+31.5
4	12.00	60	25.45	6912	3.19	424.7	34.50	61.2	105.0	14.8	-85.9
5	4.03	60	0.19	40	0.40	50.0	0.20	0.1	0.8	0.7	-12.5
	4.03	60	0.19	800	8.00	400.0	2.03	15.1	128.7	21.2	-83.5
	4.03	60	0.19	500	5.00	400.0	2.03	6.1	22.3	8.5	-61.5
	4.03	60	0.19	120	1.20	400.0	2.03	11.9	1.5	0.8	-46.7
	4.03	60	0.19	1000	2.00	1000.0	2.03	48.3	4.1	1.4	-65.8
	4.03	60	0.19	800	0.40	50.0	4.03	10.2	5.6	2.8	-50.0
	4.03	60	0.19	400	0.20	50.0	4.03	13.8	3.6	1.5	-61.1
	4.03	60	0.19	200	0.10	50.0	4.03	17.7	2.4	0.2	-91.7
	4.03	60	0.19	100	0.05	50.0	4.03	22.3	0.6	0.0	-93.3
	4.03	60	0.19	2500	5.00	400.0	4.03	16.2	43.2	18.9	-56.2
	4.03	60	0.19	1200	12.00	1000.0	7.01	--	118.4	--	--
	4.03	60	0.19	800	8.00	1000.0	7.01	6.1	22.8	8.5	-61.5
	4.03	60	0.19	500	5.00	1000.0	7.02	17.7	9.2	5.5	-40.1
	4.03	60	0.19	200	2.00	1000.0	7.02	27.1	1.6	0.7	-58.1
	4.03	60	0.19	1500	3.00	400.0	9.65	19.6	15.7	6.8	-56.7
	4.03	60	0.19	300	3.00	400.0	28.60	7.9	8.4	3.3	-61.2
	4.03	60	0.19	600	1.20	1000.0	28.60	53.3	2.9	0.5	-81.4
	4.03	60	0.19	200	0.40	1000.0	28.60	63.5	1.4	0.1	-94.3
	4.03	60	0.19	4000	8.00	1000.0	28.80	35.3	67.6	19.4	-71.3
	4.03	60	0.19	2500	5.00	1000.0	28.80	39.7	31.9	7.8	-75.5
	4.03	60	0.19	1600	0.80	400.0	34.70	51.0	4.7	1.9	-59.6
	4.03	60	0.19	800	0.40	50.0	34.70	57.7	3.0	0.5	-83.3

TABLE IV (Continued)

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
5	4.03	60	0.19	2400	1.2	1000.0	79.7	83.0	12.3	1.6	-87.0
	4.03	60	0.19	4000	2.0	1000.0	79.8	81.6	27.3	4.2	-84.6
	4.03	60	0.19	1600	0.8	1000.0	79.8	86.8	6.6	0.8	-87.9
	4.03	60	0.19	800	0.4	1000.0	79.8	90.2	1.9	0.2	-89.5
41	3.00	90	0.19	159	1.8	494.7	1.9	7.7	4.7	4.5	-4.2
	3.00	90	0.19	169	1.7	472.7	2.6	10.0	30.0	3.9	-86.9
	3.00	90	0.19	169	1.4	468.7	2.7	10.7	8.2	3.0	-63.4
	3.00	90	0.19	283	1.4	473.7	4.3	13.9	8.6	3.0	-65.7
	3.00	90	0.19	375	1.7	466.0	4.6	13.5	10.0	5.5	-45.0
	3.00	90	0.19	734	0.7	473.1	21.3	10.7	8.4	3.0	-63.4
	3.00	90	0.19	788	0.6	471.1	24.7	41.2	6.4	2.9	-55.3
	3.00	90	0.19	157	0.1	466.1	31.4	64.4	1.4	0.1	-92.9
	3.00	90	0.19	655	0.4	468.5	31.9	51.3	3.8	1.5	-60.5
	3.00	90	0.19	186	1.3	466.2	32.9	11.5	4.0	1.6	-60.0
	3.00	90	0.19	143	0.1	466.0	37.8	70.3	1.5	0.1	-93.3
	3.00	90	0.19	171	0.1	466.0	32.4	64.1	1.3	0.1	-89.7
	3.00	90	0.19	694	0.41	468.3	30.15	49.0	3.6	0.8	-77.8
40	2.00	65	0.02	10	0.1	50.0	0.12	0.1	0.9	0.3	-72.2
	2.00	65	0.02	24	0.1	50.0	0.30	0.8	1.2	0.4	-71.1
	2.00	65	0.02	65	0.1	50.0	1.49	5.8	0.4	0.3	-25.0

* The calculated pressure drops have been rounded off.
**The errors have not been rounded off. A small difference is likely to occur between the reported error and the error calculated from reported values.

Akashah (1) first applied the idea of combining an equation of state with two-phase predictive methods for calculating pressure drop in gas-liquid flow. He developed a computer program containing some widely used two-phase flow pressure drop calculating methods. The program is capable of predicting the pressure drop for horizontal, inclined, and vertical flow under adiabatic as well as non-adiabatic conditions. The algorithm for his program provided a starting point for the present work.

One of the major drawbacks in the methods using flow regime maps is that the transition between the various flow regimes is not well defined. According to Eaton et al. (20), there is considerable overlap between the various flow regimes. One of the problems encountered when using methods based on flow regime maps is determining which flow pattern exists for a given set of flow conditions, and selecting the correct correlation for that pattern. Several correlations are available for any particular flow regime. The number of possible flow patterns also varies, depending upon the method by which the regimes are identified.

Another drawback with methods like the Lockhart-Martinelli, the Beggs and Brill, and the AGA is that these methods (as originally developed) do not account for the changing compositon of the gas and the liquid. Akashah et al. (2) incorporated an equation of state in the calculation procedure to account for the changing composition.

The major drawback in the methods mentioned so far is the assumption that both phases move at the same velocity, so that the velocity of the homogeneous fluid is the sum of the superficial velocities of the two phases. While all investigators recognize the existence of slip in two-phase flow, the efforts to account for it have been mostly empirical. The Beggs and Brill method, for example, uses Equation II-15 to account for slip in two-phase flow. The AGA method uses a similar equation given by

$$f_{tp} = (1 - \frac{\ln \lambda}{S})\ f_{ns} \tag{II-34}$$

where

$$S = 1.281 - 0.478\ (-\ln \lambda) + (0.444)\ (-\ln \lambda)^2$$

$$- 0.094\ (-\ln \lambda)^3 + (0.00843)\ (-\ln \lambda)^4 \tag{II-35}$$

$$f_{ns} = 0.0014 + \frac{0.125}{Re_{tp}^{0.32}} \tag{II-36}$$

In Equation II-38, Re_{tp} is evaluated as

$$Re_{tp} = \frac{D\ V_m\ \rho'_{tp}}{\mu_{tp}} \tag{II-37}$$

where ρ'_{tp} is defined as in Equation II-19, V_m as in Equation II-12, and

$$\mu_{tp} = \mu_L\ \lambda + (1 - \lambda)\ \mu_g \tag{II-38}$$

OTHER METHODS

Lagierie et al. (29) carried out experiments using natural gas, condensate, and heavy gas oil in 6 inch diameter lines to measure the pressure drop and study the flow pattern. Lagierie et al. calculated the pressure drop for plug flow by solving the momentum balance equations for each phase. The equations are

$$\left(\frac{\partial P}{\partial X}\right)_g = \left(\frac{\tau_L\ P_L + \tau_g\ P_g}{A}\right) + [\rho_L\ H_L + \rho_g\ (1 - H_L)]g\ \sin\theta \qquad \text{(II-39)}$$

$$\left(\frac{\partial P}{\partial X}\right)_\ell = \frac{2f\ \rho_{tp}\ V_{SL}^2}{D} + \rho_{tp}\ g\ \sin\theta \qquad \text{(II-40)}$$

where, τ_L = shear stress in the liquid phase

τ_g = shear stress in the gas phase

A = pipe cross-section

V_{SL} = superficial liquid velocity

P_L = pressure in the liquid phase

P_g = pressure in the gas phase

f = friction factor

ρ_{tp} = two-phase density given by Equation II-19

Dukler et al. (18) performed a similarity analysis, and proposed a constant slip method in which the two-phase friction factor is correlated with the no-slip friction factor and the flowing volume fraction liquid. Dukler et al. recommended Equation II-34, which was developed from a similarity analysis. Degance and Atherton (17) showed that Dukler's constant slip method is more accurate than the Lockhart-Martinelli method over a wide range of data. The calculated pressure drops reported in Tables II and IV were obtained by using Equation II-34 to determine the friction factor. Safti (38) developed an algorithm to incorporate slip in the calculations, based on mathematical manipulation of Dukler's method. However, the procedure did not significantly improve pressure drop prediction. Gould (24) has accounted for slip between phases by determining the mole fraction of liquid using the equation

$$x_L' = \frac{H_L V_g}{H_L V_g + (1 - H_L) V_L} \qquad \text{II-41}$$

He then adjusted the composition of the mixture and calculated the pressure drop using the Beggs and Brill method. Gould presented a comparison of calculated and observed pressure drops for only one set of experimental data for which he obtained good agreement.

In Equation II-41, V_g and V_L stand for the gas and liquid velocities, respectively. They are defined as

$$V_g = \frac{q_g}{A(1 - H_L)} \qquad \text{(II-42)}$$

$$V_L = \frac{q_L}{A H_L} \qquad \text{(II-43)}$$

Gould used the Beggs and Brill correlation to calculate the holdup and the phase velocities for Equation II-41.

In addition to the methods mentioned above, other approaches to account for varying phase velocities are mentioned in some of the literature for transient analysis in two-phase flow. For instance, Scoggins (39) used the steady state correlation of Beggs and Brill to account for phase slippage in his transient flow model. He took this approach recognizing that there is only limited experimental data to support its validity. Fisher (21) used a variable slip correlation based on air-water flooding data. His correlation for vertical flow assumes that gravity forces dominate slip between phases and is, therefore, not applicable to horizontal flow. Lyczkowski (31) has discussed some of the computational difficulties in considering unequal phase velocities for transient flow models.

The following conclusions can be drawn from this literature review:

1. A wide variety of methods is available for calculating the pressure drop in two-phase flow. No single method works well for all flow conditions.

2. All investigators recognize that gas and liquid do not move with the same velocity in the pipeline. Yet most use superficial phase velocities in the calculation procedure, and account for slip between phases using empirical correlations.

3. The calculated pressure drop using supercritical phase velocities is generally lower than the measrued pressure drop. A higher calculated pressure drop is obtained in some cases, when the Eaton et al. holdup correlation is used to estimate liquid holdup. However, the calculated pressure drop is much lower than the observed pressure drop. The calculation procedure should be modified to obtain a higher calculated pressure drop.

4. The liquid holdup and the mixture velocity are the key parameters which influence the calculated pressure drop for long horizontal lines.

5. Using phase velocities instead of superficial phase velocities in the calculation procedure may provide a better estimate of the pressure drop.

The present work attempts to develop a methodology to systematically increase the mixture velocity so that a higher calculated pressure drop is obtained. This can be done if the two phases are assumed to move at different velocities, and by using phase velocities instead of superficial phase velocities in the calculation procedure. Such an approach, when tested with field data may provide a better method to estimate the pressure drop for gas-liquid flow in long, horizontal transmission lines.

CHAPTER III

THE PROPOSED METHOD FOR STEADY STATE CALCULATIONS

When a mixture of gas and liquid enters a pipeline the two phases tend to separate. The gas flows rapidly down the pipe leaving the liquid behind. To determine the pressure drop, the pipeline is divided into an appropriate number of segments. After the pressure at the end of the first segment is determined, the pressure at the end of the next segment is calculated. This procedure is continued until the pressure at the end of the line is determined. The key assumptions in the proposed method for pressure drop calculation are:

1. The liquid and the gas move at different velocities.
2. The gas and liquid are in equilibrium.
3. The composition of the slip liquid within a given segment of the pipeline is the same as the composition of the liquid at average segment conditions.
4. The liquid is uniformly distributed throughout the segment.
5. The velocity of the gas-liquid mixture always increases in proportion to the increase in gas velocity.

Akashah (1) found that the number of segments did not affect the calculated pressure drop for pipelines up to 100 miles long. For lines less than 100 miles long, the stated assumptions should be valid for the entire pipeline.

As mentioned in Chapters I and II, the liquid holdup in a line is the inventory of liquid within the line. For a conduit of constant cross-section, assumption 4 allows holdup to be also defined as the fraction of the pipe cross-section occupied by the liquid. In equation form, the holdup can be defined as

$$H_L' = \frac{\text{cross-section area occupied by the liquid}}{\text{total pipe cross section}} = \frac{A_L}{A} \qquad \text{(III-1)}$$

where A_L = cross-section area occupied by the liquid.

Some amount of liquid always exists in a line. When the liquid moves slower than the gas, the holdup increases. For example, if the liquid velocity reduces to 80 percent of its original velocity at pipeline conditions, the holdup will increase by 25 percent. The increase in holdup reduces the area available for gas flow, but not the gross throughput through the line. The two phases are assumed to be in equilibrium and their compositions are determined from an equilibrium flash calculation. At steady state conditions, a material balance shows that the holdup and the gross throughput through the line are not dependent on each other. Also, the composition of the individual phases and the overall composition of the mixture will change from point to point in the pipeline.

When the liquid velocity is lower than the gas velocity, the holdup increases without increasing the gross throughput. If the liquid holdup is artificially increased and the homogeneous mixture assumed to flow through the area available for gas flow, a higher calculated pressure drop will be obtained. The artificially increased holdup is not the true holdup in the line and is designated as "pseudo" holdup.

In the present work, assumptions 1 through 5 are applied in the pressure drop calculation procedure, using the "psuedo" holdup. The procedure outlined does not necessarily represent what actually happens when two phases flow simultaneously through a pipeline, but is a methodology to obtain a more accurate estimate of the two-phase pressure drop.

CHAPTER IV

APPLICATION OF THE PROPOSED METHOD TO FIELD DATA

The experimental data reported in Tables I through IV in Chapter II were obtained from different sources (3,4,5,40,41). In most cases, detailed compositional data were not available. The gas density and the liquid API gravity were reported in all sources. In cases where compositional data were not reported, the gas was assumed to consist of methane, ethane, propane, and n-butane. The gas composition was then adjusted until the gas gravity matched the reported value. The liquid was characterized as a heavy component with a characterization factor of 11.2. The normal boiling point and the molecular weight were determined from the charts of Maxwell (35). These data are sufficient to estimate thermodynamic properties using the SRK equation of state (33). In the case of the data from Baker (3), only one liquid and gas composition was reported. This liquid and gas composition was used for all cases

Tables IX through XVI in the Appendix show the gas composition and liquid characterization data derived from each source. Table XIV in the Appendix shows the liquid and gas flowrates for the data from each source.

The proposed method was introduced into the Beggs and Brill calculation procedure. A pseudo holdup was assumed and the phase velocities calculated. The superficial gas velocity and the mixture velocity were calculated using Equations II-14 and II-12, respectively. The mixture velocity to be used in the pressure drop equation was determined from the area available for gas flow. The gas velocity was used in Equation II-20 instead of the superficial gas velocity to calculate the acceleration pressure drop. The assumed pseudo holdup was used to compute the two phase friction factor and the two phase density using Equations II-15 and II-13, respectively. The pressure drop calculated this way was higher than the pressure drop calculated using superficial phase velocities. By adjusting the value of the pseudo holdup, the calculated pressure drop could be matched with the observed pressure drop.

Table V shows the calculated and observed pressure drops using this approach in the Beggs and Brill calculation procedure. In most cases, the pseudo holdup necessary to match the observed pressure drop is higher than the holdup estimated from the Beggs and Brill correlation or the Eaton et al. correlation. In four cases, shown at the bottom of the table, the calculated pressure drop was higher than the observed pressure drop even after reducing the pseudo holdup to practically zero. The observed pressure drops in these and a few other cases reported in the table are small. Considering the line lengths and the pressures involved, the accuracy of these data is questionable.

Equation for Pseudo Holdup

The pseudo holdup values are not claimed to be the true holdup in the pipeline, but rather the values necessary to match calculated pressure drops with observed pressure drops. The pseudo values were plotted individually against a number of variables like the gas velocity number, the liquid velocity number, the diameter number, the gas-liquid ratio, the Froude number, and the input liquid content λ (expressed as

TABLE V

CALCULATED PRESSURE DROPS FOR THE BEGGS AND BRILL METHOD

Data Source Ref. No.	Observed Pressure Drop psia	Calculated Pressure Drop psia	Pseudo Holdup vol. %	Calculated Liquid Velocity ft/sec	Calculated Gas Velocity ft/sec	Mixture Velocity ft/sec	Input Liquid Content vol. %
3	9.0	9.0	30.4	0.09	11.02	11.06	0.36
	1.5	1.5	48.8	0.06	3.65	3.67	0.51
	7.5	7.5	39.3	0.05	4.88	4.92	0.62
	7.0	7.0	48.3	0.07	7.65	7.71	0.78
	10.0	10.1	58.5	0.03	5.62	5.67	0.82
	20.0	20.0	19.8	1.40	13.75	14.10	2.50
	15.0	15.0	32.4	0.40	7.45	7.64	2.53
	24.0	24.0	32.8	21.10	3.42	4.03	14.96
	16.0	16.0	39.6	1.12	2.84	3.58	20.87
	6.0	6.0	32.1	2.90	3.89	4.92	21.20
	18.0	18.0	53.4	0.84	2.69	3.70	26.91
	10.0	10.0	27.5	2.81	2.89	3.96	27.20
	19.0	19.0	7.8	0.58	8.26	8.31	0.60
	32.0	32.0	0.8	104.34	11.99	12.83	6.55
4	105.0	105.0	62.8	0.97	3.10	4.74	34.50
5	0.8	0.8	27.8	0.11	21.15	21.19	0.20
	128.7	128.7	67.7	1.11	105.99	107.78	2.03
	22.3	22.3	46.1	0.94	40.30	41.13	2.03
	1.5	1.5	48.8	0.21	9.90	10.11	2.03
	4.1	4.1	53.5	1.69	4.88	6.83	2.03
	5.6	5.6	22.6	2.64	19.54	20.31	4.03
	3.6	3.6	53.5	0.56	15.90	16.54	4.03

TABLE V (Continued)

Data Source Ref. No.	Observed Pressure Drop psia	Calculated Pressure Drop psia	Pseudo Holdup vol. %	Calculated Liquid Velocity ft/sec	Calculated Gas Velocity ft/sec	Mixture Velocity ft/sec	Input Liquid Content vol. %
5	2.4	2.4	71.7	0.21	12.88	13.40	4.03
	0.6	0.6	70.8	0.10	6.08	6.33	4.03
	43.2	43.2	36.1	5.67	32.08	35.28	4.03
	118.4	118.4	66.6	1.88	53.02	56.71	7.01
	22.8	22.8	49.3	1.73	22.05	23.73	7.01
	9.2	9.2	49.7	1.07	13.78	14.85	7.02
	1.6	1.6	49.6	0.43	5.47	5.90	7.02
	15.7	15.7	36.8	3.35	18.66	20.62	9.65
	8.4	8.4	47.2	0.55	24.20	24.70	28.60
	2.9	2.9	62.2	0.87	3.61	5.04	28.60
	1.4	1.4	81.6	0.22	2.47	3.46	28.60
	67.6	67.6	53.2	6.85	20.25	28.02	28.80
	31.9	31.9	55.3	4.13	12.93	18.04	28.80
	4.7	4.7	54.3	3.66	3.62	5.53	34.70
	3.0	3.0	62.2	1.01	3.14	4.80	34.70
	1.9	1.9	75.8	0.41	2.45	3.74	34.70
	20.7	20.7	59.5	3.16	9.08	13.73	34.80
	12.3	12.3	79.6	3.37	1.28	6.26	79.70
	27.3	27.3	81.5	6.05	1.99	9.47	79.80
	6.6	6.6	83.7	2.13	0.92	4.51	79.80
	1.9	1.9	84.7	1.05	0.47	2.30	79.80
41	4.7	4.7	0.7	34.97	12.80	13.05	1.90
	30.0	30.0	69.4	0.42	36.82	37.78	2.60
	8.2	8.2	49.0	0.53	18.81	19.33	2.70
	8.6	8.6	41.4	1.03	16.56	17.29	4.30
	10.0	10.0	31.2	1.81	17.24	18.06	4.60

TABLE V (Continued)

Data Source Ref. No.	Observed Pressure Drop psia	Calculated Pressure Drop psia	Pseudo Holdup vol. %	Calculated Liquid Velocity ft/sec	Calculated Gas Velocity ft/sec	Mixture Velocity ft/sec	Input Liquid Content vol. %
41	8.4	8.4	42.3	2.48	6.79	8.61	21.30
	6.4	6.4	30.6	3.67	4.99	6.61	24.70
	1.4	1.4	76.3	0.28	1.97	2.86	31.40
	3.8	3.8	33.5	2.77	2.99	4.38	31.90
	4.0	4.0	22.8	1.26	12.16	12.53	32.90
	1.5	1.5	80.0	0.25	1.67	2.68	37.80
	1.3	1.3	81.2	0.34	1.78	2.64	32.40
	3.6	3.6	4.7	20.38	2.35	3.37	30.30
40	0.9	0.9	53.9	0.05	52.75	52.81	0.12
	1.2	1.2	54.3	0.13	53.74	53.90	0.30
	0.4	0.4	31.2	0.91	16.35	16.59	1.48
3	2.0*	3.0	0.10	30.9	5.10	5.16	0.61
5	1.9*	2.5	0.10	89.2	48.36	48.84	0.18
	1.7*	3.3	0.10	148.4	91.37	92.29	0.16
	3.8*	5.0	0.10	225.0	130.00	131.30	0.16

*Cases where observed and calculated pressure drops could not be matched.

volume percent). Beggs and Brill (8) have explained these variables in their paper. With the exception of λ, and to some extent the Froude number, none of the other variables had any significant relationship to the pseudo holdup. Beggs and Brill concluded that the Froude number and λ were the two variables which influenced the liquid holdup for their experimental data, and obtained equations of the form of Equation II-21 to estimate the holdup.

Generally, the lower the value of λ, the greater the change in the pseudo holdup needed to match the observed pressure drop. However, there was no clear cut relationship between the two variables. Some investigators (3,8,20) have obtained good correlations by including the ratio of two quantities or two similar quantities as a variable in their analyses. Baker (3) used the ratio of $(\frac{L}{G})$, Beggs and Brill used $(\frac{f_{tp}}{f_{ns}})$, and Eaton et al. used $(\frac{\mu_L}{\mu_W})$ in their correlation. Plotting H_L'/λ against λ showed a definite relationship. Figure 4 shows the plot. H_L' represents the pseudo holdup.

The pseudo holdup values in Figure 4 are the values required to match observed pressure drops with pressure drops calculated with the Beggs and Brill calculation procedure. The data was curve fitted to obtain an equation of the form

$$\ln Y = A + B(\ln \lambda) + C(\ln \lambda)^2 + D(\ln \lambda)^3 \qquad \text{(IV-1)}$$

where, $Y = \dfrac{\text{liquid holdup}}{\text{input liquid content}}$

λ = input liquid content, volume percent

A,B,C,D = constants

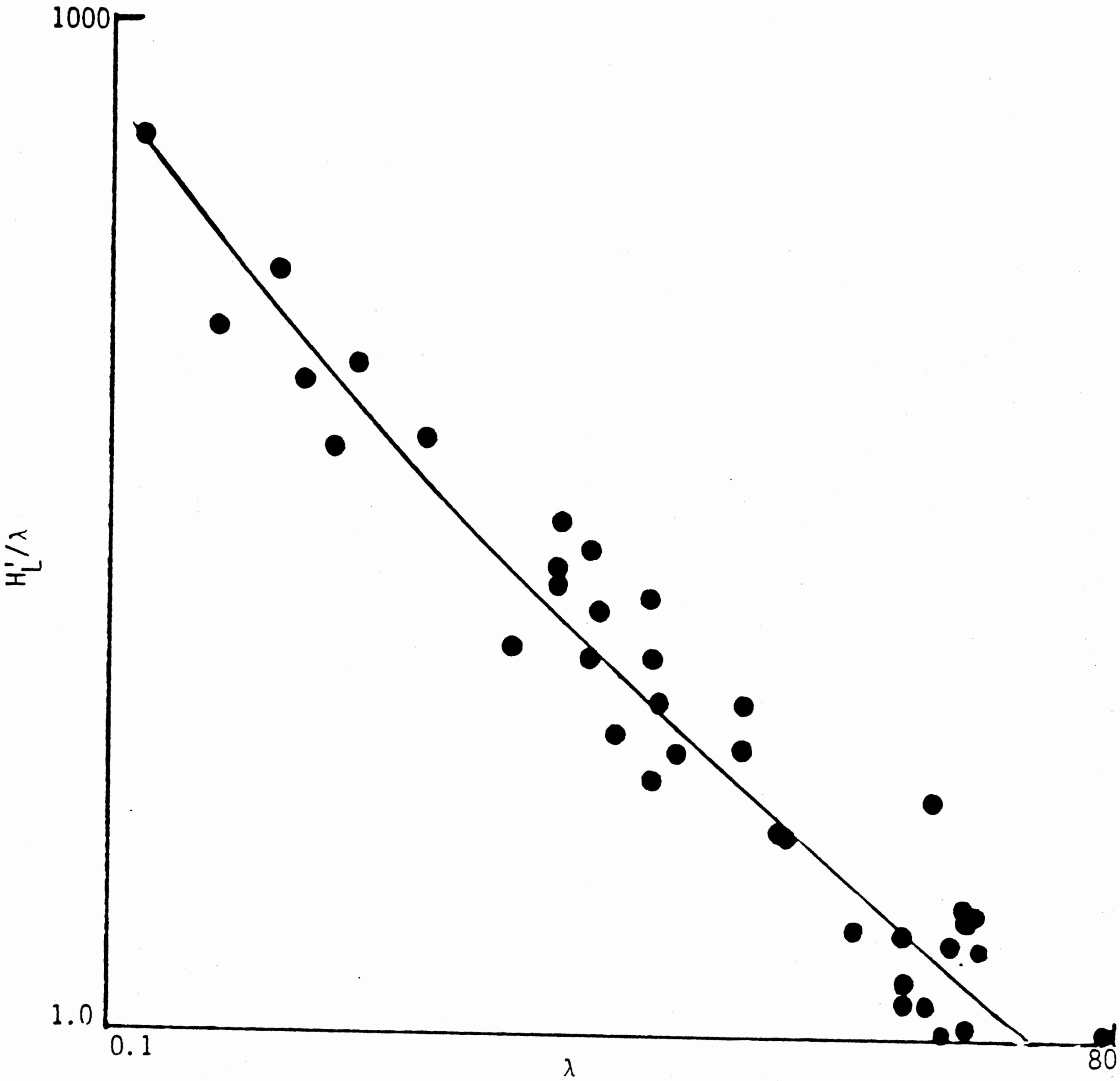

Figure 4. Pseudo Liquid Holdup as a Function of Volume Percent Input Liquid With the Beggs and Brill Calculation Procedure

The values of the constants in Equation IV-1 were determined from a regression analysis, using the MARQ computer program (13). Including the Froude number as a variable did not significantly improve the pseudo holdup values estimated by an equation similar to Equation IV-1.

Equation IV-1 was used to estimate the pseudo holdup and calculate the phase velocities. The superficial gas velocity and the mixture velocity were calculated from Equations II-14 and II-12, respectively. The mixture velocity to be used in the pressure drop equation was computed from the area available for gas flow. The gas velocity based on the area available for gas flow was used instead of superficial gas velocity to calculate the acceleration pressure drop. Table VI shows the results using the Beggs and Brill pressure drop equation.

Discussion of Results

The deviations from observed pressure drop values were calculated from:

$$\text{Percent Error} = \frac{(\text{Calculated Value} - \text{Observed Value})}{\text{Observed Value}} \qquad \text{(IV-2)}$$

The average absolute percent error is calculated from:

$$\text{Average Absolute Error, \%} = \sum_{0}^{N} \frac{|\text{Percent Error}|}{N} \qquad \text{(IV-3)}$$

where N = number of observations.

Mandhane et al. (34) have discussed other types of error parameters and their relative merits. In the present work, the average absolute percent error was chosen as the parameter for comparison, because it is

TABLE VI

CALCULATED PRESSURE DROPS USING EQUATION IV-1 AND PHASE VELOCITIES IN THE BEGGS AND BRILL METHOD

Data Source Ref. No.	Observed Pressure Drop psia	Calculated Pressure Drop psia	Pseudo Holdup Volume Percent	Percent Absolute Error
3	9.0	7.6	23.5	-15.5
	1.5	1.7	24.0	+13.3
	7.5	5.0	24.3	-33.3
	7.0	3.6	24.7	-48.6
	10.0	3.3	24.8	-67.0
	20.0	24.0	28.1	+20.0
	15.0	13.5	28.1	-10.0
	24.0	30.2	40.4	+25.8
	16.0	14.8	44.1	-7.5
	6.0	7.5	44.4	+25.0
	18.0	15.6	47.6	-13.3
	10.0	16.2	47.8	+62.0
	19.0	--	--	--
	32.0	--	--	--
4	105.0	70.3	52.5	-33.0
5	0.8	0.7	23.1	-12.5
	128.7	32.5	27.2	-74.7
	22.3	12.6	27.2	-43.5
	1.5	0.8	27.3	-46.7
	4.1	4.6	27.3	+12.2
	5.6	6.7	30.0	+19.6
	3.6	1.7	30.2	-52.8
	2.4	0.5	30.3	-79.2
	0.6	0.2	30.3	-66.7
	43.2	42.7	35.7	-1.1
	118.4	30.9	33.7	-73.9
	22.8	14.0	33.8	-38.6
	9.2	5.6	33.9	-39.1
	1.6	1.0	33.9	-37.5
	15.7	15.4	36.0	-1.9
	8.4	4.6	48.6	-45.2
	2.9	1.8	48.7	-37.9
	1.4	0.2	48.7	-85.7
	67.6	67.1	48.2	-0.7
	31.9	27.1	48.6	-15.0
	4.7	2.6	36.0	-36.6
	3.0	3.0	45.0	0.0
	1.9	2.7	51.8	+42.1
	20.7	--	--	--

TABLE VI (Continued)

Data Source Ref. No.	Observed Pressure Drop psia	Calculated Pressure Drop psia	Pseudo Holdup Volume Percent	Percent Absolute Error
5	12.3	20.9	70.2	+69.9
	27.3	54.7	69.7	+100.4
	6.6	9.8	70.2	+48.5
	1.9	0.6	48.7	-66.7
41	4.7	6.3	27.1	+34.0
	30.0	5.8	28.3	-80.5
	8.2	4.4	28.4	-46.3
	8.6	6.3	30.6	-26.7
	10.0	10.0	31.0	0.0
	8.4	7.4	44.4	-11.9
	6.4	8.7	46.4	+35.9
	1.4	0.4	50.2	-92.8
	3.8	6.0	50.3	+36.7
	4.0	4.5	28.8	+12.5
	1.5	0.4	53.3	-73.3
	1.3	0.5	50.9	-61.5
	3.6	6.7	49.5	+86.1
40	0.9	0.5	23.0	-44.4
	1.2	0.6	23.4	-50.0
	0.4	0.3	26.3	-25.0

sensitive to errors associated with small measured values of the pressure drop.

The calculated gas velocity is sensitive to the area available for gas flow. The mixture velocity increases in proportion to the estimated value of the pseudo holdup. The results in Table VI show that Equation IV-1 is not adequate to accurately estimate the pseudo holdup. Although the calculated pressure drops shown in Table VI are higher than those reported in Tables I and III, the percent errors are large.

The error in predicting the pressure drop can be reduced by improving the estimate of the liquid holdup. A flow regime map was developed for this purpose. Figure 1 shows Baker's flow regime map. The parameters plotted in this figure are $\frac{G}{\lambda'}$ and $\frac{L\lambda'\Psi}{G}$, where G and L are superficial gas and liquid mass velocities, respectively. The variables are defined in Chapter II. In Figure 2, the quantities G and L are calculated by assuming that each phase occupies the entire conduit. The quantities $\frac{G''}{\lambda'}$ and $\frac{L''}{G''}\lambda'\Psi$ were calculated using mass velocities instead of superficial mass velocities. Figure 5 shows a plot of these two quantities. The quantity G" in this figure is the gas mass velocity and L" the liquid mass velocity. They are calculated as follows:

$$G'' = \rho_g V_g \tag{IV-4}$$

and

$$L'' = \rho_L V_L \tag{IV-5}$$

The area occupied by each phase was computed from the pseudo holdup values reported in Table V. These were used to determine the phase

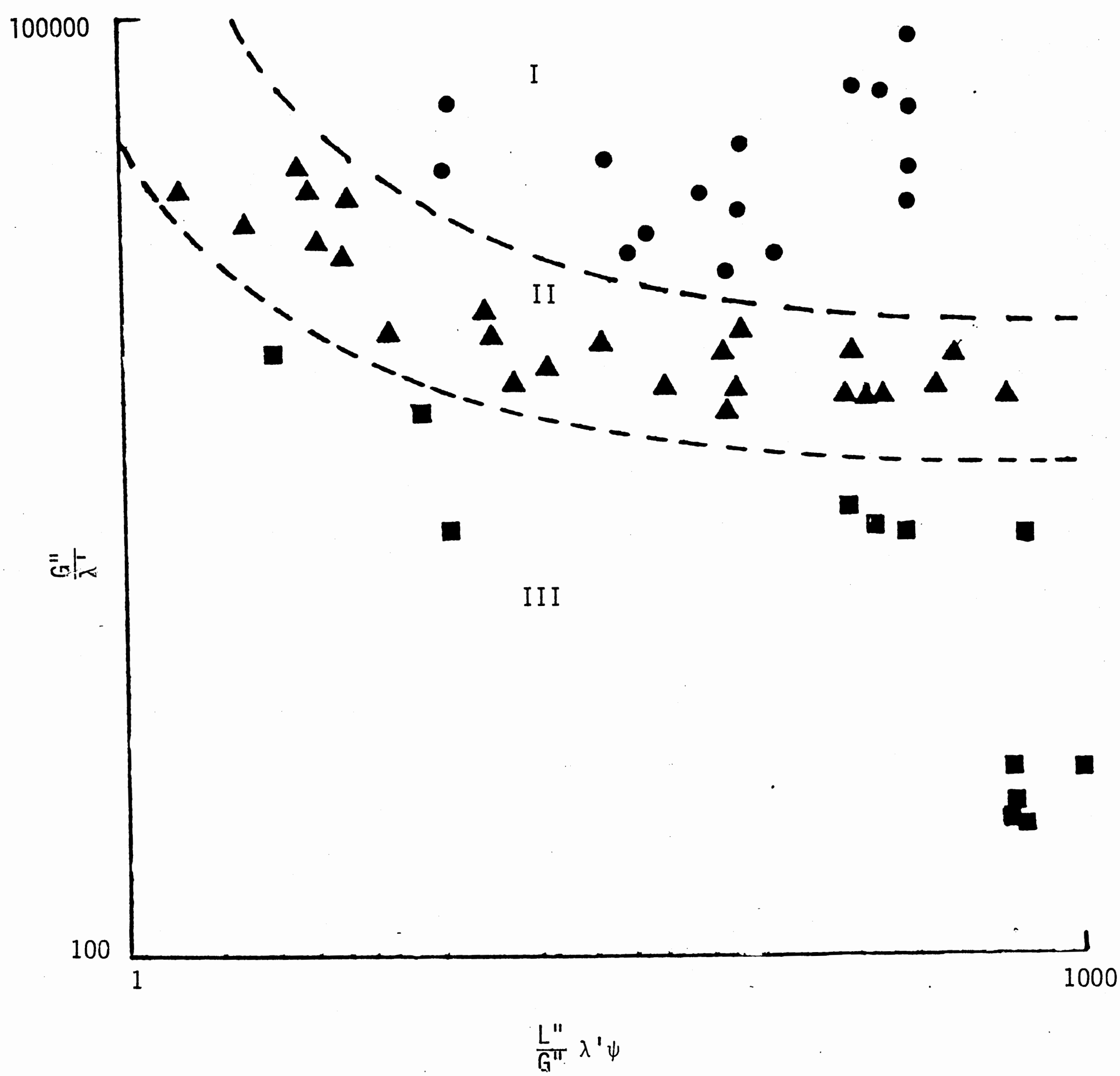

Figure 5. Flow Pattern Regions for the Proposed Method

velocities for Equations IV-4 and IV-5. These parameters may be termed "modified" Baker parameters. The data were lumped into three groups designated as Regions I, II, and III. For each region, the quantity $\frac{H_L'}{\lambda}$ was plotted as a function of the volume percent input liquid. Figures 6 through 8 show the plots. The equation for estimating the pseudo holdup corresponding to each region was obtained by curve-fitting the data using the MARQ computer program (12). The equations are:

Region I: $$Y = A_I - B_I \ln \lambda - C_I (\ln \lambda)^2 \qquad \text{(IV-6)}$$

$$Y = \ln \left(\frac{H_L'}{\lambda}\right)$$

$$A_I,\ B_I,\ C_I = \text{constant}$$

Region II: $$Y = A_{II} - B_{II} \ln \lambda + C_{II} (\ln \lambda)^2 \qquad \text{(IV-7)}$$

$$Y = \ln \left(\frac{H_L'}{\lambda}\right)$$

$$A_{II},\ B_{II},\ C_{II} = \text{constant}$$

Region III: $$Y = A_{III} (\lambda)^{B_{III}} \qquad \text{(IV-8)}$$

$$Y = \frac{H_L'}{\lambda}$$

$$A_{III},\ B_{III} = \text{constant}$$

A trial and error procedure is required to calculate the holdup values using Equations IV-6 through IV-8. The pipeline is divided into an appropriate number of segments. For each segment, the steps involved are:

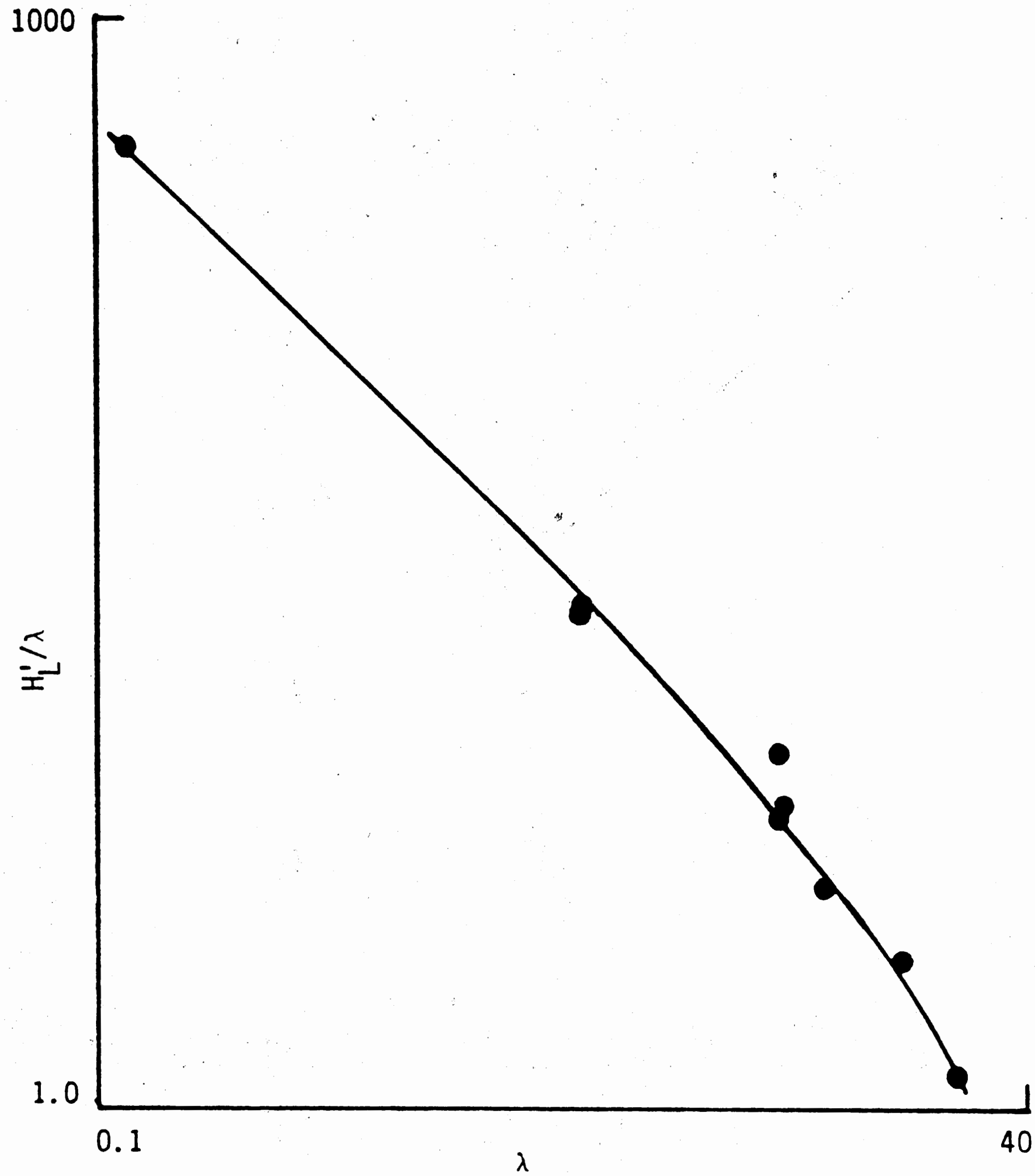

Figure 6. Pseudo Liquid Holdup as a Function of Volume Percent Input Liquid for Region I

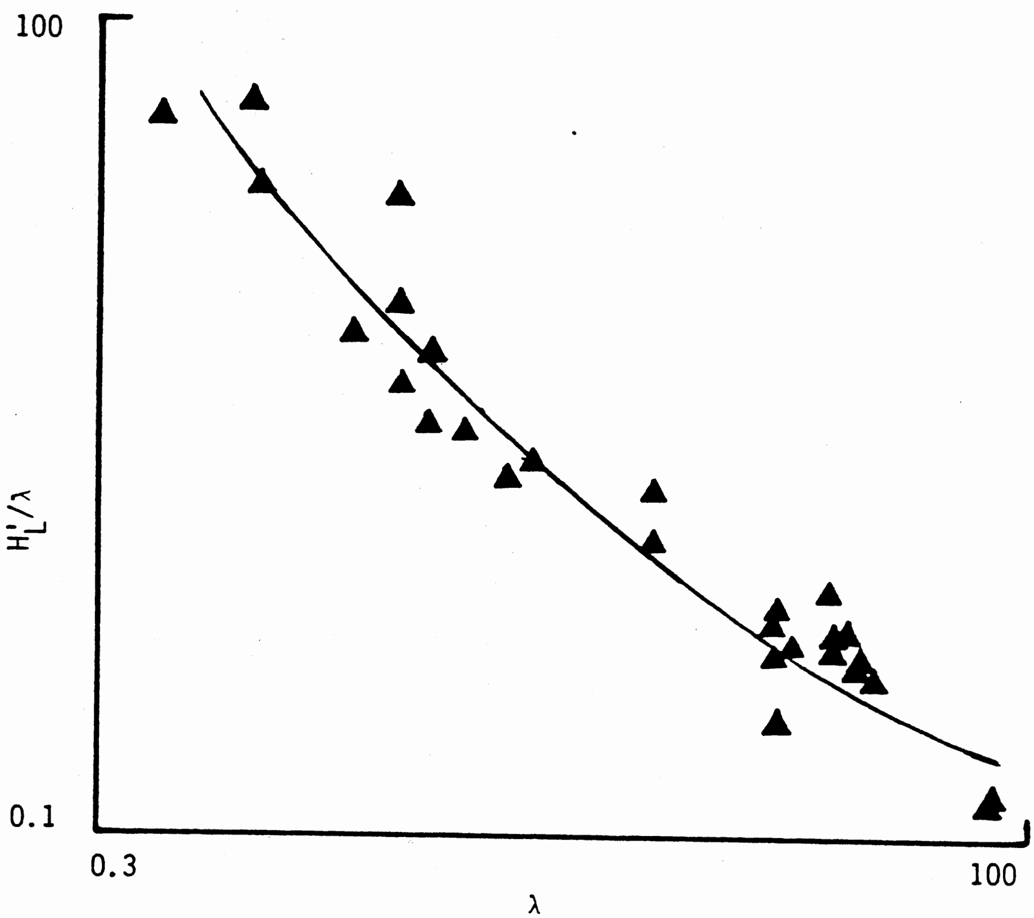

Figure 7. Pseudo Liquid Holdup as a Function of Volume Percent Input Liquid Content For Region II

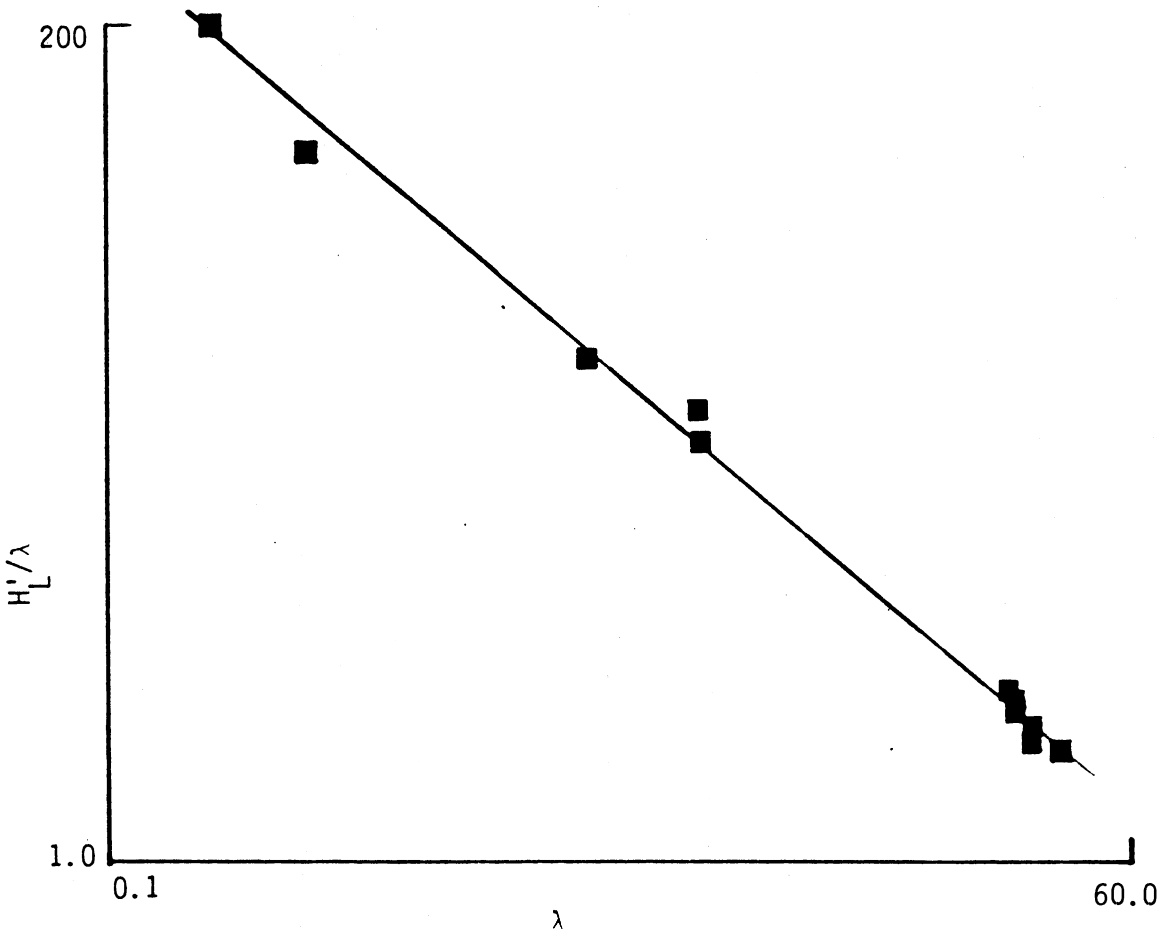

Figure 8. Pseudo Liquid Holdup as a Function of Volume Percent Input Liquid Content for Region III

1. Estimate a holdup using Equation IV-1 and calculate the phase velocities.

2. Calculate the quantities $\frac{G'''}{\lambda'}$ and $\frac{L''}{G''}\lambda'\Psi$, and locate the flow regime.

3. Calculate the holdup using the equation which corresponds to the flow regime located in step 2. If the point corresponding to the calculated values of $\frac{G''}{\lambda'}$ and $\frac{L''}{G''}\lambda\Psi$ falls within Region I, use Equation IV-6. If the point falls within Region II, use Equation IV-7. If the point falls within Region III, use Equation IV-8.

4. Using the pseudo holdup values from step 3, calculate the quantities $\frac{G''}{\lambda'}$ and $\frac{L''}{G''}\lambda\Psi$, and locate the flow regime again.

If the flow regime from step 4 is the same as the one from step 2, then use the holdup value from step 3 and complete the pressure drop calculation using the gas velocity, and the mixture velocity based on the area available for gas flow in the pressure drop equation. If the flow regime from step 4 is not the same as the one from step 2, use the holdup value from step 3 as the next estimate, and repeat steps 2 through 4 until the flow regimes match.

Table VII shows the pressure drops calculated by the Beggs and Brill equations using Figures 5 through 8 to estimate the pseudo holdup. The percent errors reported in Table VII is in many cases much lower than those reported in Table VI. This shows that using the flow regime map provides a better estimate of the pseudo holdup, and leads to a better prediction of the pressure drop.

The line of demarkation between the regions in Figure 5 is arbitrary. The lines were drawn as shown using Figure 2 as a guideline. The regions can also be defined differently. For example,

TABLE VII

CALCULATED PRESSURE DROP USING FIGURE 5 AND PHASE VELOCITIES FOR THE BEGGS AND BRILL METHOD

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Region	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
3	7.75	65	2.16	192	11.68	726.7	0.36	39.9	III	9.0	7.6	-15.5
	7.75	70	2.01	107	6.67	1081.7	0.44	39.2	II	1.5	1.7	+13.3
	7.75	70	2.14	236	11.95	1077.7	0.61	43.3	III	7.5	5.0	-33.3
	7.75	69	2.14	244	9.48	1088.7	0.78	66.1	II	7.0	3.6	-48.6
	10.14	69	7.83	244	9.48	1082.7	0.82	32.8	II	10.0	3.3	-67.0
	4.03	79	0.69	627	7.47	1101.7	2.46	28.8	II	20.0	24.0	+20.0
	5.94	72	2.01	627	7.47	1084.7	2.53	28.7	II	15.0	13.5	-10.0
	10.14	78	7.83	6592	11.89	974.7	14.96	35.7	I	24.0	28.9	+20.4
	10.14	69	7.83	4167	12.05	976.7	20.58	41.4	II	16.0	14.8	-7.5
	7.75	66	2.14	4970	6.47	978.7	21.01	32.1	I	6.0	6.0	0.0
	10.14	82	7.83	5420	4.35	930.0	20.51	41.7	II	18.0	15.6	-13.3
	7.75	82	2.14	5420	4.35	954.7	26.91	74.5	III	10.0	16.2	+62.0
	7.75	75	2.14	514	25.97	997.7	0.60	--	--	19.0	--	--
	7.75	80	2.14	5484	25.52	1021.7	6.55	--	--	32.0	--	--
4	12.00	60	25.45	6912	3.19	424.7	34.50	52.5	II	105.0	70.3	-33.0
5	4.03	60	0.19	40	0.40	50.0	0.20	37.2	III	0.8	0.7	-12.5
	4.03	60	0.19	800	8.00	400.0	2.03	28.9	II	128.7	32.5	-74.7
	4.03	60	0.19	500	5.00	400.0	2.03	27.2	II	22.3	12.6	-43.5
	4.03	60	0.19	120	1.20	400.0	2.03	51.2	III	1.5	1.4	-6.7
	4.03	60	0.19	1000	2.00	1000.0	2.03	75.0	III	4.1	4.6	+12.2
	4.03	60	0.19	800	0.40	50.0	4.03	29.0	II	5.6	6.3	+12.5
	4.03	60	0.19	400	0.20	50.0	4.03	56.3	III	3.6	4.1	+13.9
	4.03	60	0.19	200	0.10	50.0	4.03	56.3	III	2.4	1.0	-58.3
	4.03	60	0.19	100	0.05	50.0	4.03	56.4	III	0.6	0.3	-50.0
	4.03	60	0.19	2500	5.00	400.0	4.03	40.8	I	43.2	42.7	+1.1
	4.03	60	0.19	1200	12.00	1000.0	7.01	43.3	I	118.4	67.3	-43.1
	4.03	60	0.19	800	8.00	1000.0	7.01	43.2	I	22.8	23.2	+1.7
	4.03	60	0.19	500	5.00	1000.0	7.02	43.2	I	9.2	9.3	+1.1
	4.03	60	0.19	200	2.00	1000.0	7.02	42.6	I	1.6	1.6	0.0
	4.03	60	0.19	1500	3.00	400.0	9.65	40.9	I	15.7	15.4	-1.9
	4.03	60	0.19	300	3.00	400.0	28.60	53.5	I	8.4	8.6	+2.4
	4.03	60	0.19	600	1.20	1000.0	28.60	47.9	II	2.9	1.8	-37.9
	4.03	60	0.19	100	0.20	1000.0	28.60	75.1	III	1.4	1.4	0.0
	4.03	60	0.19	4000	8.00	1000.0	28.80	47.9	II	67.6	67.1	-0.7
	4.03	60	0.19	2500	5.00	1000.0	28.80	47.6	II	31.9	27.1	-15.0

TABLE VII (Continued)

Data Source Ref. No.	Pipe Diameter Inches	Line Temp. °F	Line Length Miles	Liquid Flow Rate bbl/day	Gas Flow Rate MMSCFD	Inlet Pressure psia	Input Liquid Content Vol. %	Liquid Holdup Vol. %	Region	Observed Pressure Drop psia	Calculated* Pressure Drop psia	Percent ** Error
5	4.03	60	0.19	1600	0.8	400.0	34.70	51.8	II	4.7	4.1	-12.8
	4.03	60	0.19	800	0.4	50.0	34.70	51.8	II	3.0	2.4	-25.0
	4.03	60	0.19	400	0.2	50.0	34.70	27.0	I	1.9	1.1	-42.1
	4.03	60	0.49	2400	1.2	400.0	34.80	51.6	II	20.7	15.3	-26.1
	4.03	60	0.19	2400	1.2	1000.0	79.70	82.1	II	12.3	10.8	-12.2
	4.03	60	0.19	4000	2.0	1000.0	79.80	82.2	II	27.3	28.7	+5.1
	4.03	60	0.19	1600	0.8	1000.0	79.80	82.2	II	6.6	5.2	-21.2
	4.03	60	0.19	800	0.4	1000.0	79.80	82.2	II	1.9	1.5	-21.0
41	3.0	60	0.19	158	1.8	494.7	1.90	29.0	II	4.7	6.3	+34.0
	3.0	60	0.19	169	1.7	472.7	2.60	52.0	I	30.0	21.3	-29.0
	3.0	60	0.19	169	1.4	468.7	2.70	53.2	III	8.2	11.1	-35.4
	3.0	90	0.19	283	1.4	473.7	4.30	--	--	8.6	--	--
	3.0	60	0.19	375	1.7	466.0	4.60	29.2	II	10.0	10.0	0.0
	3.0	60	0.19	734	0.7	473.1	21.30	63.2	II	8.4	9.4	+11.9
	3.0	60	0.19	788	0.6	471.1	24.70	44.6	II	6.4	8.7	+44.6
	3.0	60	0.19	157	0.1	466.1	31.40	75.1	III	1.4	1.4	0.0
	3.0	60	0.19	655	0.4	468.5	31.90	49.9	II	3.8	6.0	+36.7
	3.0	60	0.19	186	1.3	466.2	32.90	28.6	II	4.0	4.9	+22.5
	3.0	60	0.19	143	0.1	466.0	37.80	79.2	III	1.5	1.5	0.0
	3.0	90	0.19	171	0.1	466.0	32.40	--	--	1.3	--	--
	3.0	60	0.19	671	0.41	468.3	30.30	48.7	II	3.6	6.7	+86.1
40	2.0	65	0.02	10	0.1	50.0	0.12	51.3	I	0.9	0.7	-22.2
	2.0	65	0.02	24	0.1	50.0	0.30	45.4	II	1.2	0.6	-50.4
	2.0	65	0.02	65	0.1	50.0	1.49	29.7	II	0.4	0.3	-14.3

* The calculated pressure drops have been rounded off.
**The errors have not been rounded off. A small difference is likely to occur between the reported error and the error calculated from reported values.

Region I may be considered to include points with a value of less than 10 for the horizontal axis in Figure 5. Region II may be considered to include all points with a horizontal axis value between 10 and 150. Region III may be considered as all points with a value greater than 150 for the horizontal axis. However, defining the three regions in this way does not provide a better estimate of the pressure drop than is obtained by using Figure 5 along with Equations IV-6 through IV-9. The results in Table VII show that, although the distinction between the different flow regimes may not be well defined, the concept of using a flow regime map is useful for improving the predictive methods for two-phase pressure drop.

Table VIII gives a comparison of the average absolute error for different types of pressure drop calculation procedures. The results show that the methodology based on the assumption that the two phase flow at different velocities substantially cuts down the average absolute error. The absolute error reduces to 40% of the error associated with the Beggs and Brill method assuming homogeneous or no-slip flow.

Pressure Drop Calculation Procedure

The following procedure is recommended for calculating the two-phase pressure drop based on the results shown in Table VII:

1. Divide the pipeline into an appropriate number of segments. For each segment, assume a pressure drop and compute the required phase properties at average segment conditions.

2. Estimate a holdup from Equation IV-1. Use the trial and error procedure outlined earlier to calculated the holdup. The procedure

TABLE VIII

COMPARISON OF AVERAGE ABSOLUTE ERROR IN CALCULATING PRESSURE DROP USING SUPERFICIAL AND PHASE VELOCITIES

Method of Calculation	Holdup Correlation Used	Type of Flow Assumed	Average Absolute Error, %
Beggs and Brill	Beggs and Brill	No-Slip	66.7
Beggs and Brill	Eaton	No-Slip	64.5
Beggs and Brill	Equations II-5, IV-6, and IV-7	Slip	26.2
AGA	AGA	No-Slip	56.1
AGA	Eaton	No-Slip	63.0
AGA	Equations IV-5, IV-6, and IV-7	Slip	29.4

involves using the flow regime map (Figure 5) along with Equations IV-6 through IV-8.

3. Calculate the phase velocities.

4. Calculate the superficial phase velocities using Equations II-14, and II-12.

5. Compute the mixture velocity to be used in the pressure drop equation based on the area available for gas flow.

6. Compute the pressure drop using the mixture velocity determined from step 5 and the gas velocity (instead of the superficial gas velocity) in the pressure drop equation.

If the pressure drop calculated from step 6 agrees with the assumed pressure drop within a specified tolerance, proceed with calculations for the next segment. If not, repeat steps 1 through 6 with the calculated pressure as the next estimate. The total pressure drop is obtained by algebraically summing the calculated pressure drop in all the segments. The pressure drop calculated with this approach provides a better estimate of the true pressure drop than that calculated by using superficial phase velocities.

CHAPTER V

CONCLUSIONS AND RECOMMENDATIONS

An improved method has been developed for calculating the pressure drop in gas-liquid flow in long horizontal transmission lines. The procedure is programmed into the pressure drop calculation routines originally developed by Akashah (1). Several cases involving experimental data from different sources were calculated using the program. The following are the conclusions of the study:

1. In most cases, the pressure drop calculated using superficial phase velocities is lower than the observed pressure drop.

2. In some cases, using the Eaton et al. holdup correlation results in a higher calculated pressure drop.

3. The liquid holdup and the mixture velocity are the key parameters which influence the calculated pressure drop.

4. In most cases, the pseudo liquid holdup required to match calculated pressure drops with observed pressure drops is greater than the holdup estimated by the Beggs and Brill, the AGA, and the Eaton et al. correlations.

5. The pressure drop estimated using phase velocities instead of superficial phase velocities, and increasing the mixture velocity in proportion to the pseudo holdup is higher than the pressure drop estimated by assuming no-slip flow.

6. The absolute error in the estimated pressure drop can be substantially reduced by using a flow regime map (Figure 5) along with

appropriate equations (Equations IV-6 through IV-9) to estimate the liquid holdup.

7. The methodology based on the assumption that the two phases flow at different velocities provides a better estimate of the pressure drop.

8. The Beggs and Brill method provides the best estimate of the pressure drop for flow in long horizontal lines when the method recommended in this study is incorporated into the calculation procedure.

The following recommendations are made for calculating the pressure drop in two-phase flow, and for future work:

1. Calculations should first be made with the assumption of no-slip flow to obtain an estimate of the minimum pressure drop to expect in a line.

2. The concept of allowing the two phases to move at different velocities should be applied to study the change in pressure, temperature, and holdup with time during unsteady flow conditions.

3. An investigation should be conducted to study the effect of slip on the holdup in inclined lines.

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APPENDIX

SPECIFICATION OF DATA FOR TEST CASES

TABLE IX

SPECIFICATIONS FOR DATA FROM BAKER (3)

Gas Composition:

Component	Mole%
C_1	95.12
C_2	3.26
C_3	0.84
$i\text{-}C_4$	0.22
$n\text{-}C_4$	0.26
$i\text{-}C_5$	0.06
CO_2	0.10
C_{6+}	0.14
	100.00

Liquid Composition:

Component	Mole%
C_1	24.28
C_2	4.57
C_3	2.86
$i\text{-}C_4$	1.58
$n\text{-}C_4$	1.87
$i\text{-}C_5$	1.58
C_{6+}	63.26
	100.00

TABLE IX (Continued)

C_6+ Specification:

Fraction #	NBP, °F	API Gravity	Molecular Weight
1	128.6	73.7	71.0
2	287.0	58.0	116.1
3	351.7	52.8	137.0
4	484.4	43.7	185.5
5	594.3	37.4	243.1
6	687.7	32.7	281.0

TABLE X

SPECIFICATIONS FOR DATA FROM BAKER (4)

Gas Composition:

Component	Mole%
C_1	84.2
C_2	5.8
C_3	5.0
n-C_4	5.0
	100.00

C_{6+} Specification:

NBP (Normal Boiling Point)	390°F
API Gravity	37
Molecular Weight	154

TABLE XI

SPECIFICATIONS FOR DATA FROM BAKER (5)

Gas Composition:

Component	Mole%
C_1	77.0
C_2	16.0
C_3	6.0
$n\text{-}C_4$	1.0
	100.00

C_{6+} Specification:

NBP (Normal Boiling Point)	575°F
API Gravity	30
Molecular Weight	226

TABLE XII

SPECIFICATIONS FOR DATA FROM VAN WINGEN (41)

Gas Composition:

Component	Mole%
C_1	88.0
C_2	5.0
C_3	5.5
$n\text{-}C_4$	1.5
	100.00

C_{6+} Specification:

NBP (Normal Boiling Point)	365°F
API Gravity	36
Molecular Weight	144

TABLE XIII

SPECIFICATIONS FOR DATA FROM SCHNEIDER ET AL. (40)

Pressure Drop, psia	0.90	1.21	0.35
Component	Composite Feed Composition, Mole%		
C_1	74.61	71.18	54.78
C_2	15.48	14.78	11.38
C_3	4.04	3.86	2.97
$n\text{-}C_4$	2.73	2.62	1.97
$n\text{-}C_{13}$	3.14	7.56	28.90
	100.00	100.00	100.00

TABLE XIV

SPECIFICATION OF FLOW RATES

Data Source Ref. No.	Line Temp. °F	Pipe Diameter inches	Pipeline Length Miles	Observed Pressure Drop psia	Liquid Flow Rate bbl/d	Gas Flow Rate MMSCFD	Inlet Pressure psia	Volume Percent Liquid at Inlet
3	65	7.75	2.16	9.0	192	11.68	726.7	0.36
	70	7.75	2.01	1.5	107	6.67	1081.7	0.51
	70	7.75	2.14	7.5	236	11.95	1077.7	0.62
	69	7.75	2.14	7.0	244	9.48	1088.7	0.78
	69	10.14	7.83	10.0	244	9.48	1082.7	0.82
	79	4.03	0.69	20.0	627	7.47	1101.7	2.50
	72	5.94	2.01	15.0	627	7.47	1084.7	2.53
	78	10.14	7.83	24.0	6592	11.89	974.7	15.25
	69	10.14	7.83	16.0	4167	12.05	976.7	20.87
	66	7.75	2.14	6.0	4970	6.47	978.7	21.20
	82	10.14	7.83	18.0	5420	4.35	930.0	26.91
	82	7.75	2.14	10.0	5420	4.35	954.7	27.20
	75	7.75	2.14	19.0	514	25.97	997.7	0.60
	80	7.75	2.14	32.0	5484	25.52	1021.7	6.55
4	60	12.0	25.45	105.0	6912	3.19	424.7	34.50
5	60	4.03	0.19	0.8	40	0.40	50.0	0.20
	60	4.03	0.19	128.7	800	8.00	400.0	2.03
	60	4.03	0.19	22.3	500	5.00	400.0	2.03
	60	4.03	0.19	1.5	1200	1.20	400.0	2.03
	60	4.03	0.19	4.1	1000	2.00	1000.0	2.03
	60	4.03	0.19	5.6	800	0.40	50.0	4.03
	60	4.03	0.19	3.6	400	0.20	50.0	4.03

TABLE XIV (Continued)

Data Source Ref. No.	Line Temp. °F	Pipe Diameter inches	Pipeline Length Miles	Observed Pressure Drop psia	Liquid Flow Rate bbl/d	Gas Flow Rate MMSCFD	Inlet Pressure psia	Volume Percent Liquid at Inlet
5	60	4.03	0.19	2.4	200	0.1	50.0	4.03
	60	4.03	0.19	0.6	100	0.1	50.0	4.03
	60	4.03	0.19	43.2	2500	5.0	400.0	4.03
	60	4.03	0.19	118.4	1200	12.0	1000.0	7.01
	60	4.03	0.19	22.8	800	8.0	1000.0	7.01
	60	4.03	0.19	9.2	500	5.0	1000.0	7.02
	60	4.03	0.19	1.6	200	2.0	1000.0	7.02
	60	4.03	0.19	15.7	1500	3.0	400.0	9.65
	60	4.03	0.19	8.4	300	3.00	400.0	28.60
	60	4.03	0.19	2.9	600	1.2	1000.0	28.60
	60	4.03	0.19	1.4	200	0.4	1000.0	28.60
	60	4.03	0.19	67.6	4000	8.0	1000.0	28.80
	60	4.03	0.19	31.9	2500	5.0	1000.0	28.80
	60	4.03	0.19	4.7	1600	0.8	400.0	34.70
	60	4.03	0.19	3.0	800	0.4	50.0	34.70
	60	4.03	0.19	1.9	400	0.2	50.0	34.70
	60	4.03	0.19	20.7	2400	1.2	400.0	34.80
	60	4.03	0.19	12.3	2400	1.2	1000.0	79.70
	60	4.03	0.19	27.3	4000	2.0	1000.0	79.80
	60	4.03	0.19	6.6	1600	0.8	1000.0	79.80
	60	4.03	0.19	1.9	800	0.4	1000.0	79.80

TABLE XIV (Continued)

Data Source Ref. #	Line Temp. °F	Pipe Diameter inches	Pipeline Length Miles	Observed Pressure Drop psia	Liquid Flow Rate bbl/d	Gas Flow Rate MMSCFD	Inlet Pressure psia	Volume Percent Liquid at Inlet
41	90	3.00	0.19	4.7	169	1.7	494.7	1.9
	90	3.00	0.19	30.0	169	1.4	472.7	2.6
	90	3.00	0.19	8.2	186	1.3	468.7	2.7
	90	3.00	0.19	8.6	283	1.4	473.7	4.3
	90	3.00	0.19	10.0	375	1.7	466.0	4.6
	90	3.00	0.19	8.4	734	0.7	473.1	21.3
	90	3.00	0.19	6.4	788	0.6	471.1	24.7
	90	3.00	0.19	1.4	157	0.1	466.1	31.4
	90	3.00	0.19	3.8	655	0.4	468.5	31.9
	90	3.00	0.19	4.0	143	0.1	466.2	32.8
	90	3.0	0.19	1.5	143	0.1	466.0	37.8
	90	3.0	0.19	1.3	171	0.1	466.0	32.4
	90	3.0	0.19	3.6	671	0.4	468.3	30.3
40	65	2.00	0.02	0.9	10	0.1	50.0	0.12
	65	2.00	0.02	1.2	24	0.1	50.0	0.30
	65	2.00	0.02	0.4	65	0.1	50.0	1.49

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VITA

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