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(Technical Report)

Wet Gas Flowmetering Guideline

The American Society of Mechanical Engineers

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Wet Gas Flowmetering Guideline

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THE AMERICAN SOCIETY OF MECHANICAL ENGINEERS

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FOREWORD

This Technical Report is an advisory State-of-the-Art document for wet gas flowmetering applications as understood in 2005. It is based on available wet gas flowmetering research papers, commercial literature, and practical experiences from the oil and gas industry up unto the end of 2005. The operating principles apply to steady-state flows where phase change is not a dominant issue. However, it should be understood that many wet gas flowmetering applications could be unsteady state flows, and phase change could be a dominant issue.

Topics included in this technical report are as follows:

(a) definition of terms
(b) the significance of two-phase flow patterns and the associated flow pattern maps to wet gas meter applications
(c) practical industrial problems that occur when applying the wet gas flowmetering technologies
(d) uncertainty associated with wet gas flowmetering
(e) a comprehensive technical paper reference list
(f) the derivations and limitations of the published wet gas flowmeter correlations

This Report was prepared by Subcommittee 19 (SC 19) of the ASME Standards Committee on Measurement of Fluid Flow in Closed Conduits. At the time of the preparation of this Report, the members of SC 19 considered the subject of wet gas flowmetering not mature enough for a standard to be produced, and that the application of a standard for wet gas metering systems could hinder the continuing development of new technologies.

This document does not endorse any wet gas metering technology or any meter test facility. All information given in this document is derived from available literature. Subcommittee members and contributing authors have attempted to give as fair and precise a description of the known issues, but it should be understood that wet gas flowmetering is a developing science and the committee members and contributing authors are not responsible for the veracity of any referenced material.

Suggestions for improvement of this Standard are welcome. They should be sent to The American Society of Mechanical Engineers; Attn: Secretary, MFC Standards Committee; Three Park Avenue; New York, NY 10016-5990.

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WET GAS FLOWMETERING GUIDELINE

1 INTRODUCTION

This Technical Report discusses the existing definitions of "wet gas flow" and provides suggested definitions for use. Common wet gas flowmetering terminologies, principles, and limitations of the available wet gas meter technologies are also discussed.

Wet gas flowmetering is an important flow measurand in many industries. If a relatively small volume of liquid is present in a gas it is generally said to be "wet." Wet gas flows are not new occurrences in industry (e.g., wet saturated steam flows have been produced since the industrial revolution) but it is only recently that attempts to meter wet gas flows (e.g., by the oil and gas industry) with improved and a perhaps better understood uncertainties have been made. Measurement techniques are being continuously developed but accepted single-phase (dry) gas meter uncertainty is as yet not attainable when a wet gas flow is present. Due to the difficulties involved in wet gas metering it is unlikely that the same level of uncertainty seen with single-phase gas metering will be achieved in the foreseeable future.

There are two distinct wet gas-metering situations:

(a) Where some flow rate knowledge is initially known, for example,

(1) the total mass flow rate is known (such as in a closed cycle system, e.g., a steam power cycle) and either the ratio of liquid-to-gas flow rates or one of the phase flow rates is required to be metered.

(2) one phase flow rate is known (from some other means) and the other phase flow rate is to be metered.

(b) No flow rate information is known (e.g., unprocessed wet natural gas flows) and either or both the liquid and gas phase flow rates are required to be metered. This is a considerably more difficult metering situation as extra information is required and meters being developed for this situation are considered to be at the cutting edge of fluid flowmetering technology.

NOTE: Most of the current technologies ignore the effects of multi-component liquids present in wet gas flows. However, some metering systems are designed to estimate the different quantities of liquid components in a wet gas flow.

2 SYMBOLOGY AND DEFINITIONS

In order to understand the metering techniques available for wet gas flowmetering it is necessary to understand the symbology and definitions that have been used in this Technical Report.

2.1 English Symbols

Symbol	Description	Dimension	Units
A	Area of the meter inlet	L^2	mm^2 (in.^{2})
A_g	Cross-sectional area of gas	L^2	mm^2 (in.^{2})
A_l	Cross-sectional area of liquid	L^2	mm^2 (in.^{2})
A_t	Area of a DP meter at the throat	L^2	mm^2 (in.^{2})
L	Unit length of pipe	L	m (ft)
C_d	Discharge coefficient of a differential pressure (DP) Meter	Dimensionless	Dimensionless
$C_{d_{tp}}$	Discharge coefficient of a differential pressure (DP) meter calculated with use of ΔP_{tp}	Dimensionless	Dimensionless
C_{dg}	Discharge coefficient of a DP meter with wet gas if gas phase flowed alone	Dimensionless	Dimensionless

C_{dl}	Discharge coefficient of a DP meter with wet gas if liquid phase flowed alone	Dimensionless	Dimensionless
C_o	Injected tracer liquid concentration	Various	Various
C_s	Samples tracer liquid concentration	Various	Various
C	Chisholm's parameter	Dimensionless	Dimensionless
D	Meter inlet pipe diameter	L	mm (in.)
M	Murdock's gradient	Dimensionless	Dimensionless
DP	Differential pressure	M/LT^2	Pa (psi)
PD	Positive displacement	N/A	N/A
USM	Ultrasonic flowmeter	N/A	N/A
E	Velocity of approach of a differential pressure meter	Dimensionless	Dimensionless
$f_n(a,b,...)$	Unspecified function with variables, *a*, *b*, etc.	N/A	N/A
f_n^*	Superscript "*" indicates a rearrangement of function f_n	N/A	N/A
f_r	Friction factor	Dimensionless	Dimensionless
f_l	Friction factor for if liquid phase of a two-phase flow flowed alone	Dimensionless	Dimensionless
f_g	Friction factor for if gas phase of a two-phase flow flowed alone	Dimensionless	Dimensionless
Fr	Single-phase flow Froude number	Dimensionless	Dimensionless
Fr_g	Gas densiometric Froude number	Dimensionless	Dimensionless
Fr_l	Liquid densiometric Froude number	Dimensionless	Dimensionless
g	Gravitational constant (9.81 m/s^2)	L/T^2	m/s^2 (ft/sec^2)
GVF	Gas volume fraction at operating conditions	Dimensionless	Dimensionless
GOR	Gas oil volume ratio at standard condition	Dimensionless	MMscf/bbls
h	Enthalpy	$(L/T)^2$	kJ/kg (Btu/lb)
h_l	Liquid enthalpy	$(L/T)^2$	kJ/kg (Btu/lb)
h_g	Gas enthalpy	$(L/T)^2$	kJ/kg (Btu/lb)
h_v	Vapor enthalpy	$(L/T)^2$	kJ/kg (Btu/lb)
LVF	Liquid volume flow rate fraction at operating conditions	Dimensionless	Dimensionless
LMF	Liquid mass flow rate fraction	Dimensionless	Dimensionless
m	Mass	M	kg (lb)
$\dot{m}$	Mass flow rate	M/T	kg/s (lbm/sec)
$\dot{m}_g$	Gas flow rate in mass	M/T	kg/s (lbm/sec)

$\dot{m}_{g,Apparent}$	Gas flow rate in mass predicted by DP meter when using ΔP_{tp}	*M/T*	kg/s (lbm/sec)
$\dot{m}_l$	Liquid mass flow rate	*M/T*	kg/s (lbm/sec)
OR	"Overreading" (i.e., positive bias of gas meter with two-phase flow)	Dimensionless	Dimensionless
P	Line pressure	M/LT^2	Pa (psi)
$\dot{Q}$	Volume flowing	L^3/T	m^3/h (ft^3/hr)
$\dot{Q}_g$	Actual gas volume flowing	L^3/T	m^3/h (ft^3/hr)
$\dot{Q}_{g\,SAC}$	Flowing gas volume if the gas was at standard atmospheric condition	L^3/T	m^3/h (ft^3/hr)
$\dot{Q}_{tp}$	Total volume flow of the two-phase flow	L^3/T	m^3/h (ft^3/hr)
q	Injected tracer volume flowing	L^3/T	m^3/h (ft^3/hr)
$\dot{Q}_l$	Actual liquid volume flowing	L^3/T	m^3/h (ft^3/hr)
Re	Reynolds number	Dimensionless	Dimensionless
Re_{sg}	Superficial gas Reynolds number of a two-phase flow	Dimensionless	Dimensionless
Re_g	Gas Reynolds number	Dimensionless	Dimensionless
Re_l	Liquid Reynolds number	Dimensionless	Dimensionless
s	Slip velocity	*L/T*	m/s (ft/sec)
s_e	Entropy	ML^2/T^2	kJ/K
S_R	Slip ratio	Dimensionless	Dimensionless
t	Time	*T*	s
T	Temperature	N/A	K, R
$\bar{U}$	Average gas velocity	*L/T*	m/s (ft/sec)
$\bar{U}_g$	Average actual gas velocity in two-phase flow	*L/T*	m/s (ft/sec)
$\bar{U}_l$	Average actual liquid velocity in two-phase flow	*L/T*	m/s (ft/sec)
$\bar{U}_{sg}$	Superficial gas velocity	*L/T*	m/s (ft/sec)
$\bar{U}_{sl}$	Superficial liquid velocity	*L/T*	m/s (ft/sec)
V_g	Gas volume	L^3	M^3 (ft^3)
V_l	Liquid volume	L^3	M^3 (ft^3)
We	Single-phase Weber number	Dimensionless	Dimensionless
We_{tp}	Weber number modified for two-phase flow	Dimensionless	Dimensionless
WLR	Water–liquid ratio	Dimensionless	Dimensionless

x	Flow quality (ratio of gas to total mass flow)	Dimensionless	Dimensionless
x_m	James modified flow quality	Dimensionless	Dimensionless
x_l	Multi-component liquid mix flow quality	Dimensionless	Dimensionless
X_{LM}	Modern Lockhart–Martinelli parameter	Dimensionless	Dimensionless
$X_{Murdock}$	Murdock parameter	Dimensionless	Dimensionless
$X_{Chisholm}$	Chisholm parameter (which is the same as X_{LM})	Dimensionless	Dimensionless
X	Original Lockhart–Martinelli parameter	Dimensionless	Dimensionless
Y or Y_g	Expansibility coefficient for a differential pressure meter	Dimensionless	Dimensionless
Y_{tp} or $Y_{g\,tp}$	Expansibility coefficient for a differential pressure meter calculated with use of ΔP_{tp}	Dimensionless	Dimensionless
K	Meter factor for linear single-phase meters	$1/L^3$	$1/m^3$
PVT	Pressure, volume, and temperature equation of state calculations	N/A	N/A
R	Gas volume flow rate at standard conditions per barrels of hydrocarbon liquid at separator conditions	Dimensionless	Dimensionless
S_h	Ratio of volumes between the standard U.S. barrel volume at set flow conditions and that same masses volume at standard conditions	Dimensionless	Dimensionless
C_A	Armand coefficient	Dimensionless	Dimensionless
DR	Gas-to-liquid density ratio at flowing conditions	Dimensionless	Dimensionless
Water Cut	Water volume to total liquid volume flow rate at standard conditions	Dimensionless	Dimensionless
V_g	Volume of gas in a unit length of pipe	L^3	m^3 (ft^3)
V_l	Volume of liquid in a unit length of pipe	L^3	m^3 (ft^3)

2.2 Greek Symbols

ΔP_t	Nonphase specific gas differential pressure	M/LT^2	Pa (psi)
ΔP_g	Superficial gas differential pressure	M/LT^2	Pa (psi)
ΔP_l	Superficial liquid differential pressure	M/LT^2	Pa (psi)
ΔP_{tp}	Actual wet gas/two-phase differential pressure	M/LT^2	Pa (psi)
ΔP_{hl}	Head loss/permanent pressure loss	M/LT^2	Pa (psi)
ΔP_f	Friction pressure loss in straight pipe of single- phase flow	M/LT^2	Pa (psi)

ΔP_{l_f}	Friction pressure loss in straight pipe if liquid phase of two-phase flow flowed alone	M/LT^2	Pa (psi)
ΔP_{g_f}	Friction pressure loss in straight pipe if gas phase of two-phase flow flowed alone	M/LT^2	Pa (psi)
ξ	Smith & Leang blockage factor	Dimensionless	Dimensionless
θ	Linn modified Murdock gradient	Dimensionless	Dimensionless
α_g	Void fraction	Dimensionless	Dimensionless
α_l	Liquid hold up	Dimensionless	Dimensionless
δ	Fluctuation	Various	Various
β	"Beta" ratio, i.e., the square root of the ratio of minimum cross sectional area to inlet area of a DP meter	Dimensionless	Dimensionless
κ	Isentropic exponent	Dimensionless	Dimensionless
μ	Absolute viscosity	ML/T	Pa.s (lbf.s/ft^2)
μ_g	Absolute viscosity of gas	ML/T	Pa.s (lbf.s/ft^2)
μ_l	Absolute viscosity of liquid	ML/T	Pa.s (lbf.s/ft^2)
ρ	Density	M/L^3	kg/m^3 (lbm/ft^3)
ρ_g	Gas density	M/L^3	kg/m^3 (lbm/ft^3)
ρ_l	Liquid density	M/L^3	kg/m^3 (lbm/ft^3)
$\rho_{homogenous}$	Density of a perfectly mixed two-phase flow	M/L^3	kg/m^3 (lbm/ft^3)
$v_{homogeneous}$	Specific volume of a perfectly mixed two-phase flow	L^3/M	m^3/kg (ft^3/lbm)
$\rho_{l_{homogeneous}}$	Density of a multi-component liquid homogeneous mix	M/L^3	kg/m^3 (lbm/ft^3)
σ_l	Liquid interfacial tension	M/T^2	N/m (lbf/ft)
ω	Frequency	$1/T$	Hz

2.3 Definition of Terms

2.3.1 Common Terms Used when Describing Wet Gas Flow. Many common terms used when describing wet gas flow are common to the general industry and single-phase flowmetering. It is therefore not seen as appropriate or necessary to list them all here. The Norwegian Society for Oil and Gas Measurement has produced a "Handbook of Multiphase Flowmetering" [1] has a good definition of terms section for wet gas and multiphase flowmetering to which we refer the interested reader. However, the following several terms need to be discussed in this Report directly:

Reynolds number: for single-phase flowmeters the Reynolds number is often important. The Reynolds number is the ratio of the inertia to viscous forces and is shown in eq. (1).

$$Re = \frac{gas\ inertia\ \ forces}{gas\ viscous\ \ forces} = \frac{\rho \bar{U} D}{\mu} = \frac{4\dot{m}}{\pi \mu D} \quad (1)$$

There are many cases in industry where a dry gas meter is used to meter the flow when the flow is wet gas. Many meter designs are calibrated by relating the meter factor (e.g., a DP meter discharge coefficient, C_d, or a turbine meter K-factor) to the flows Reynolds number. With low liquid loading wet gas flow, it is often assumed that the liquid effect on the average gas velocity and the difference between gas mass flow and total mass flow is negligible and the Reynolds number is calculated using the "superficial" gas velocity or the gas mass flow rate. The "superficial" gas velocity is defined as the average gas velocity of the flow if that gas flow component of the wet gas flowed alone in the pipe. The associated Reynolds number is denoted here by Re_{sg} to indicate the value was obtained by assuming the liquid present in the gas has no influence. In other words, Re_{sg} is the Reynolds number that would exist if the gas flowed alone. Equation (2) shows this Reynolds number.

$$\mathrm{Re}_{sg} = \frac{\rho_g \bar{U}_{sg} D}{\mu_g} = \frac{4\dot{m}_g}{\pi \mu_g D} \tag{2}$$

where $\bar{U}_{sg}$ is the superficial gas flow average velocity that is calculated by eq. (3).

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} \tag{3}$$

The superficial gas velocity is always less than the actual average gas velocity, i.e., $U_g > U_{sg}$, due to the blocking effect of the liquid phase causing a gas velocity increase. For low liquid loading, dry gas meters are often used to predict the gas flow rate. These meters are often sized using an expected Reynolds number range based on eq. (2). It should be noted that as the liquid loading increases for a given gas flow rate, the assumption that single-phase flowmetering methods and eq. (2) can be utilized becomes increasingly invalid.

Lockhart–Martinelli parameter: a dimensionless number used to express the liquid fraction of a wet gas stream, and is the square root of the ratio of the liquid inertia if the liquid flowed alone in the conduit to the gas inertia if the gas flowed alone in the conduit. It is denoted here by the symbol X_{LM} and it is calculated by eq. (4).

$$\mathrm{X}_{\mathrm{LM}} = \sqrt{\frac{\text{Inertia of Liquid Flowing Alone}}{\text{Inertia of Gas Flowing Alone}}} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} \tag{4}$$

There can be considerable confusion over the origins and the physical meaning of this parameter. This is discussed in detail in Nonmandatory Appendix A.

The natural gas production industry tends to use the Lockhart–Martinelli parameter to describe the relative amount of liquid in a gas flow. The Lockhart–Martinelli parameter is often denoted in wet gas metering papers by the upper case letter "*X*." It is also occasionally denoted as "*LM*." Due to the similarity that the upper case "*X*" has to steam "quality" (or "dryness fraction"), which is symbolized by the lower case "*x*," in this Report, the Lockhart–Martinelli parameter is denoted by "X_{LM}."

Note that in eq. (4) the volume flow rates are at actual flowing conditions and not at any reference condition. In eq. (4) the gas mass or volume flow rate terms indicates the total gaseous phase (i.e., it includes liquid vapor) mass or volume flow rate. The gas density is the density of the overall gas and liquid vapor phase mix. That is, it includes the effect of any liquid component mass saturated in the gas.

Froude number and the densiometric Froude number: the gas densiometric Froude number (Fr_g) is a wet gas flow modification of the standard Froude number (Fr). The standard Froude number is defined as the square root of the inertial force to the gravitational force ratio and is calculated by eq. (5).

$$\mathrm{Fr} = \sqrt{\frac{\text{Inertia Force}}{\text{Gravity Force}}} \quad (5)$$

The gas densiometric Froude number is defined as the square root of the gas inertial force if the gas phase flowed alone to the liquid gravity force ratio. The gas densiometric Froude number is calculated by eq. (6).

$$Fr_g = \sqrt{\frac{\text{Superficial Gas Inertia Force}}{\text{Liquid Gravity Force}}} = \frac{\bar{U}_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} \quad (6)$$

Where the term $\bar{U}_{sg}$ is the superficial gas velocity as found by eq. (3). Equation (6) is derived from first principles in Nonmandatory Appendix A.

The liquid densiometric Froude number is defined as the square root of the ratio of the liquid inertial force if the liquid flowed alone to the liquid gravity force. It is calculated by eq. (7).

$$Fr_l = \sqrt{\frac{\text{Superficial Liquid Inertia Force}}{\text{Liquid Gravity Force}}} = \frac{\bar{U}_{sl}}{\sqrt{gD}}\sqrt{\frac{\rho_l}{\rho_l - \rho_g}} \quad (7)$$

where U_{sl} is the superficial liquid flow average velocity, which is calculated by eq. (8).

$$\bar{U}_{sl} = \frac{\dot{m}_l}{\rho_l A} \quad (8)$$

Occasionally, the Lockhart–Martinelli parameter (X_{LM}) may be described as the ratio of the liquid densiometric Froude number and the gas densiometric Froude number as it will be noted that the liquid gravity forces cancel out in this case. That is:

$$X_{LM} = \frac{Fr_l}{Fr_g} = \frac{\bar{U}_{sl}}{\bar{U}_{sg}}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} \quad (9)$$

Weber number: there are only a few technical papers that discuss the effect liquid properties have on flowmeters being used to meter wet gas flow. These generally discuss gross differences in response when changing liquid types. No technical paper is known to "us" that give details of the effect on meters of changing individual liquid properties (i.e., viscosity and surface/interfacial tension). Many researchers suspect that the interfacial tension may have an effect. Fluid mechanics literature defines the Weber number as the ratio of the liquid inertial force to the liquids surface tension force [eq. (10)]. With wet gas flow, this Report defines the Weber number to be the gas inertial force if the gas flowed alone in the conduit to the liquid surface tension force [eq. (11)]. That is:

$$We = \frac{Inertia\ Forces}{Surface\ Tension\ Forces} \quad (10)$$

$$We_{tp} = \frac{\dot{m}_g^2}{\sigma_l \rho_g D^3} \quad (11)$$

density ratios: many wet gas meters have outputs dependent on pressure. When correction factors are required to correct liquid-induced gas metering errors, these factors sometimes include pressure effects. To keep correction factors dimensionless (as well as for other theoretical reasons) phase density ratios are often used instead of the pressure. That is, gas-to-liquid density ratio $\left(\frac{\rho_g}{\rho_l}\right)$ or liquid-to-gas density ratio $\left(\frac{\rho_l}{\rho_g}\right)$. In this Report the term "DR" denotes the gas-to-liquid density ratio.

gas volume fraction: the fraction of gas volume flow rate compared to the total volume flow rate (i.e., the sum of the gas and liquid volume flow rates). It is calculated by eq. (12).

$$\mathrm{GVF} = \frac{\dot{Q}_g}{\dot{Q}_g + \dot{Q}_l} \quad (12)$$

Equation (12) is at actual flowing conditions. The GVF is often expressed as a percentage [see eq. (13)].

$$\mathrm{GVF}(\%) = \frac{\dot{Q}_g}{\dot{Q}_g + \dot{Q}_l} * 100\% \quad (13)$$

Note that in eqs. (12) and (13) the gas volume flowing is the volume of the humid gas at flow conditions. That is, the gas volume flow rate is the gas phase and the liquid vapor component. The liquid volume flow rate is the volume of the "free liquid" flow rate.

There is often confusion caused by the fact that the parameter called the "gas volume fraction" (or "GVF") is actually the gas Volume flow rate fraction. It is therefore possible (and common) for engineers to mistake the GVF to be defined as the gas to total unit pipe volume ratio for a steady wet gas flow. It is not. It is the gas volume flow rate to the total volume flow rate ratio of a steady wet gas flow. These two different parameters are only the same value when there is no "slip" (i.e., no average velocity difference; see para. 2.3.2) between the gas and liquid phases. This condition rarely exists in practice and usually the slip value is unknown. These statements are discussed in further detail in Nonmandatory Appendix B.

The liquid flow rate term is the "free liquid" flow rate. The term "free liquid" indicates a flowing component that is in liquid form and is distinct from any liquid vapor phase commingling with the gas phase. A gas flow with a finite relative humidity below saturation is not considered to be wet gas. Saturated gas flow (i.e., a relative humidity of 100%) is also not considered to be a wet gas flow as long as there is no free liquid. In cases of single component flows (e.g., steam, refrigerants, etc.) there is no difference between "free liquid" quantity and the total liquid component quantity. From here on, unless otherwise stated, this document drops the term "free liquid" and uses "liquid" to describe the liquid components flowing in excess to that saturated in the gas phase.

For the ASME wet gas flow definition GVF values need to be converted to Lockhart–Martinelli parameter values when evaluating whether a flow is wet gas flow or general two-phase/multiphase flow. This is discussed in para. 2.3.3.

liquid volume fraction: the term liquid volume flow (or "LVF") is sometimes used. This is the liquid flowing volume to the total flowing volume ratio. It is calculated by eq. (14).

$$LVF = \frac{\dot{Q}_l}{\dot{Q}_g + \dot{Q}_l} = \frac{\dot{Q}_l}{\dot{Q}_g} * GVF \qquad (14)$$

Equation (14) is at actual flowing conditions. The LVF is often expressed as a percentage [see eq. (15)].

$$\mathrm{LVF}\ (\%) = \frac{\dot{Q}_l}{Q_g + Q_l} * 100\% \qquad (15)$$

For the ASME wet gas definition, LVF values need to be converted to Lockhart–Martinelli parameter values when evaluating whether a flow is wet gas flow or general two-phase/multiphase flow. This is discussed in para. 2.3.3.

flow quality/dryness fraction: industries dealing with steam flows tend to use "steam quality" (often called the "dryness fraction" or the "gas mass fraction") to describe the liquid content of the flow. Steam quality is denoted as lowercase "*x*." The definition of quality is the vapor mass flow rate to the total mass flow rate ratio [see eq. (16)].

$$x = \frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g} \qquad (16)$$

This is often expressed in terms of percentage as shown by eq. (17):

$$x = \left(\frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g} \right) * 100\% \qquad (17)$$

In steam-based industries, steam is sometimes called wet steam if it is not superheated regardless of the steam quality. That is any quality value greater than zero is sometimes called wet steam flow. In other instances steam is considered wet if the quality is greater than 50%. These wet gas definitions do not agree with the ASME wet gas flow definition. To compare with the ASME definition steam quality values should be converted to Lockhart–Martinelli parameter values when evaluating whether a flow is wet or general two-phase/multiphase flow. This is discussed in para. 2.3.3.

gas-to-liquid flow rate ratio: the liquid content in a gas flow can be described directly as the liquid mass flow rate to gas mass flow rate ratio (or vice versa). A liquid-to-gas ratio can be described by mass or volume ratio. If a volume ratio is used, then it must be stated if the gas volume is at flow conditions or at standard conditions. Metering engineers rarely use this method of describing liquid content in a two-phase flow but oil industry reservoir engineers often describe flows as a number of barrels of liquid per million standard cubic feet of gas. That is, a liquid volume to a gas volume if the gas flowed at standard conditions. As the actual flow conditions are usually at nonstandard conditions, a conversion is required to get the gas-to-liquid ratio in actual conditions. Liquid-gas mass or volume ratios should be converted to Lockhart–Martinelli parameter values when evaluating whether a flow is wet or general two-phase/multiphase flow. This is discussed in para. 2.3.3.

liquid loading: it should be noted that the term "liquid loading" is commonly used in industry and is utilized often in this Report. This term is a nonquantitative expression that relates relative amounts of liquid flowing with a gas flow. That is, a light liquid loading indicates that there is a relatively small amount of liquid flowing with the gas, and a heavy liquid loading indicates that there is a relatively large amount of liquid flowing with the gas.

multiphase flow: there is some ambiguity in industry over the meaning of the phrase "multiphase flow." Technically there are three phases. These are the solid, liquid, and gaseous states of matter. Therefore, technically a multiphase flow is a flow with all three phases, and a two-phase flow is a flow with a combination of any two of these three phases. However, it has become standard practice in the oil and gas industry to refer to two-phase flows of gas and liquid as multiphase flows when the liquid phase consists of more than one component (e.g., water, hydrocarbon liquids, and/or injected fluids). Although this term is technically incorrect, this Report recognizes that the term is in universal use in industry. It should therefore be understood that the term "multiphase" means a flow of gas and liquid where there is more than one liquid component. A "multiphase meter" nearly always means the metering device meters the gas flow rate, the total liquid flow rate, and the liquid phases "water cut." ("Water cut" is a term used to describe the ratio of water to the total hydrocarbon liquid and water volume flow when both liquid components are converted to a standard pressure and temperature. It is usually expressed as a percentage.) Most multiphase meters do not meter any entrained solids such as sand production.

2.3.2 Other Two-Phase Flow Liquid to Gas Ratio Phrases. The "Void Fraction" is defined as the ratio of the gas phase to pipe cross-sectional areas [see eq. (18)]. It is usually denoted by the symbol "α_g."

$$\alpha_g = \frac{A_g}{A} \tag{18}$$

Note that the "liquid hold up" (sometimes called simply "hold up") is defined as the ratio of the liquid phase to pipe cross-sectional area [see eq. (19)].

$$\alpha_l = 1 - \alpha_g = \frac{A_l}{A} \tag{19}$$

The void fraction and hold up are not descriptive of the liquid-to-gas flow rate ratio as they do not account for slip between the phases. Nevertheless, this term is occasionally mentioned in wet gas metering literature and it is therefore noted here.

Note that the terms "slip" (denoted by the letter "s") and the "slip ratio" (denoted by the letter "S_R") are commonly used in the two-phase flow texts. Slip is the condition when the phases travel at different velocities. The "slip velocity" is the difference in the actual average velocities of the phases (which is different to the superficial phase velocities). Equation (20) shows the slip velocity. Also note that the slip ratio is the ratio of the actual average gas velocity to the actual average liquid velocity. Equation (21) shows the slip ratio.

$$s = \bar{U}_g - \bar{U}_l \tag{20}$$

$$S_R = \frac{\bar{U}_g}{\bar{U}_l} \tag{21}$$

Nonmandatory Appendix B discusses the difference between GVF and void fraction in more detail.

2.3.2 Relations Between Different Wet Gas Flow Parameters. There is no universally agreed method of describing the relative quantity of liquid and gas in a two-phase flow. With no standard method of describing the liquid content of a gas flow, engineers are free to use any method they wish to describe the flow. There can be and occasionally there are other ways of describing the liquid content of a wet gas flow but the methods discussed here are the most common methods in industry. With the use of these different methods it is often necessary to relate one parameter to the other. Equations (22) through (27) indicate how to interrelate the six most common terms: the Lockhart–Martinelli parameter, the gas volume fraction, the quality (or "dryness fraction"), the liquid volume fraction, the liquid-to-gas mass flow rate ratio, and the liquid-to-gas volume flow rate ratio. They also show how the system pressure (i.e., the liquid-to-gas density ratio for a given fluid combination) dictates the relationship between these parameters.

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{1-(GVF)}{(GVF)}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{(LVF)}{1-(LVF)}\sqrt{\frac{\rho_l}{\rho_g}} \quad (22)$$

$$GVF = \frac{1}{1+\left(\frac{m_l}{m_g}*\frac{\rho_g}{\rho_l}\right)} = \frac{1}{\left(\frac{\dot{Q}_l}{\dot{Q}_g}\right)+1} = 1-(LVF) = \frac{\sqrt{\frac{\rho_l}{\rho_g}}}{X_{LM}+\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{\left(\frac{\rho_l}{\rho_g}\right)}{\left(\frac{\rho_l}{\rho_g}\right)+\left(\frac{1-x}{x}\right)} \quad (23)$$

$$x = \frac{1}{1+\left(\frac{\dot{m}_l}{\dot{m}_g}\right)} = \frac{1}{1+\left(\frac{\rho_l}{\rho_g}\frac{\dot{Q}_l}{\dot{Q}_g}\right)} = \frac{1}{1+X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1+\left(\left(\frac{1-(GVF)}{(GVF)}\right)*\frac{\rho_l}{\rho_g}\right)} = \frac{1}{1+\left(\left(\frac{(LVF)}{1-(LVF)}\right)*\frac{\rho_l}{\rho_g}\right)} \quad (24)$$

$$\frac{\dot{m}_l}{\dot{m}_g} = X_{LM}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{\rho_l}{\rho_g}\frac{\dot{Q}_l}{\dot{Q}_g} = \frac{1-(GVF)}{GVF}\frac{\rho_l}{\rho_g} = \frac{LVF}{1-LVF}\frac{\rho_l}{\rho_g} = \frac{1-x}{x} \quad (25)$$

$$\frac{\dot{Q}_l}{\dot{Q}_g} = X_{LM}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\rho_g}{\rho_l}\frac{\dot{m}_l}{\dot{m}_g} = \frac{1-(GVF)}{GVF} = \frac{LVF}{1-(LVF)} = \left(\frac{1-x}{x}\right)\frac{\rho_g}{\rho_l} \quad (26)$$

$$LVF = 1-(GVF) = \frac{X_{LM}\sqrt{\frac{\rho_g}{\rho_l}}}{1+X_{LM}\sqrt{\frac{\rho_g}{\rho_l}}} = \frac{\left(\frac{1-x}{x}\right)}{\left(\frac{1-x}{x}\right)+\frac{\rho_l}{\rho_g}} = \frac{\frac{\rho_g}{\rho_l}}{\frac{\rho_g}{\rho_l}+\frac{\dot{m}_g}{\dot{m}_l}} = \frac{\frac{\dot{Q}_l}{\dot{Q}_g}}{1+\frac{\dot{Q}_l}{\dot{Q}_g}} \quad (27)$$

NOTES:

(1) All gas volume flows are at actual flowing conditions (and not standard conditions).

(2) It is not possible to include the barrels of liquid/standard volume gas parameter here (typically "bbl/MMSCFD" — barrels of liquid/million standard cubic feet) as the conversion of standard volume to actual volume for nonperfect gases is different for individual gases. For given gas and liquid types, standard volume flow rates and actual flow conditions the gas mass flow rate or actual volume flow rate must be calculated along with the gas and liquid densities, and these values are then input with the liquid mass or actual volume flow rates to the relevant equation above.

Nonmandatory Appendices C and D discuss the relationships of the different wet gas flow defining parameters in detail, and Nonmandatory Appendix D shows graphical representations of their relationships.

3 TYPES OF WET GAS FLOWS

This Report defines wet gas flow as any gas/liquid two-phase flow where

$$X_{LM} = \sqrt{\frac{Inertia\ of\ Liquid\ Flowing\ Alone}{Inertia\ of\ Gas\ Flowing\ Alone}} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} \leq 0.3 \quad (28)$$

This Report defines any gas and liquid flow combination as a two-phase flow. The gas and liquid phase components are not relevant to the definition of wet gas flow. Wet gas flow is therefore a special sub-set of two-phase flow. $X_{LM} \leq 0.3$ is a two-phase flow that is a wet gas flow. $X_{LM} > 0.3$ is a general two-phase flow and not a wet gas flow.

As the wet gas definition does not differentiate between phase components, a wet gas flow can be a flow with $X_{LM} \leq 0.3$ that has one or more liquid component(s). In the oil and gas industry it is common to call a flow of natural gas/hydrocarbon liquid/water/other injected fluids a multiphase flow. Hence, the definition allows a wet gas flow to be a single-component two-phase flow (e.g., water/steam), a two-component two-phase flow (e.g., air/water) or a multi-component two-phase flow (e.g., natural gas/hydrocarbon liquid/water).

It should be noted that the common terminology of the oil and gas industry has an ambiguity in this regard. It is fairly common for multiphase flow to be thought of as a natural gas/hydrocarbon liquid/water/other liquid flow where the liquid loading is high compared to a wet gas flow (e.g., a GVF < 80%). However, if a wet gas flow has multiliquid components the wet gas flow is then sometimes referred to as a multiphase flow.

Most wet gas meter testing done by the oil and gas industry has a Lockhart–Martinelli parameter of $X_{LM} \leq 0.3$. There are flows that some in industry label “wet gas flows” with a Lockhart–Martinelli parameter value greater than 0.3. However, some natural gas production companies cap wet gas meter use at a Lockhart–Martinelli parameter of ≤0.3, as beyond this increases the likelihood of severe slugging in the risers and pipelines and the risk of damage to the meters and associated equipment.

In none quantitative terms, wet gas flow is often simply defined by industry as gas, that contains a relatively small quantity of liquid. The definition of wet gas being any gas and liquid flow with $X_{LM} \leq 0.3$ means there can be a considerable difference in liquid to gas mass flow rates across the huge pressure (and therefore gas-to-liquid density ratio) range. This leads to different wet gas metering technologies being favored across different ranges within the $X_{LM} \leq 0.3$ definition. For example, the metering system and measurement techniques chosen to measure the flow departing a natural gas separator on an offshore platform where typically the gas will have a very small amount of liquid(s) may be different from a wet gas flowmetering system chosen for the wellhead of a gas condensate well to measure flow rates of gas with a relatively large quantity of liquid. Some in industry have therefore gone further than this Report in trying to create subsets of wet gas flow.

Nonmandatory Appendix C discusses the issues on why it is not scientifically possible to state equivalent wet gas flow definitions to the $X_{LM} \leq 0.3$ definition in other parameters such as GVF. Nonmandatory Appendix E discusses wet gas definitions of API and the Norwegian Society for Oil and Gas Measurement. Nonmandatory Appendix F shows worked examples of how to derive the wet gas nondimensional parameters from real industrial data as typically supplied by pipe line/systems engineers.

4 FLOW PATTERN

The flow pattern (or “flow regime”) describes the way the fluids are dispersed in a pipe flow. It is known that different meters can be affected in different ways (that are not always fully understood) by particular flow patterns so the flow pattern is important to wet gas flowmetering.

When gas and liquid flows simultaneously in a pipe, the phases can distribute themselves in a variety of patterns. The patterns differ from each other in the spatial distribution of the interfaces, resulting in different flow characteristics. This distribution is termed the flow pattern (or “flow regime”). The flow pattern in a given two-phase flow system depends on many variables, of which the following are considered to be of prime importance:

(a) gas and liquid flow rates
(b) pipe diameter and inclination angle
(c) the physical properties of the phases (i.e., gas and liquid densities, liquid viscosities, and the surface tension)

Many two-phase flow textbooks offer descriptions of flow patterns and names for particular common patterns. Note that there are no officially recognized definitions for common flow patterns, and, as a result, the same physical flow pattern can be called by different names by different industries, engineers and academics. However, there is now reasonable agreement on names for most types of flow pattern, and it is only in the areas of transition between these that ambiguities generally still exist.

Commonly accepted definitions for two-phase flow patterns are usually given in the literature for horizontal and vertical flows separately due to gravity having a significant effect. There is virtually no published information on inclined pipe two-phase flow patterns. Well-known sketches of flow patterns are shown in Fig. 4-1 for horizontal flows and in Fig. 4-2 for vertical up flows. Only typical flow patterns that can exist with a Lockhart–Martinelli parameter of less than 0.3 are shown in this document.

For horizontal flow: at low gas flow rates, the liquid flows at the bottom of the pipe with the higher velocity gas flowing above the liquid, driving the liquid by the shear force at the interface. This flow pattern is commonly called “stratified flow” or “separated flow.”

As the gas flow rate increases, instability at this interface increases and waves appear. This flow pattern is commonly called “stratified wavy flow” or simply “wavy flow.” In many cases, however, there is no distinction made between smooth stratified flow and wavy stratified flow, and the stratified wavy flow is just called “stratified” flow or “separated” flow.

If enough liquid is present, then the waves can get large compared to the pipe diameter and sporadically completely block the gas flow. This is called “semi-slug” flow.

At higher gas flow rates, the liquid travels in a nonsymmetrical ring (due to gravity) around the pipe periphery with the liquid droplet laden gas core traveling through the center of the pipe. This is often called “annular-mist flow” or “annular-dispersed flow.”

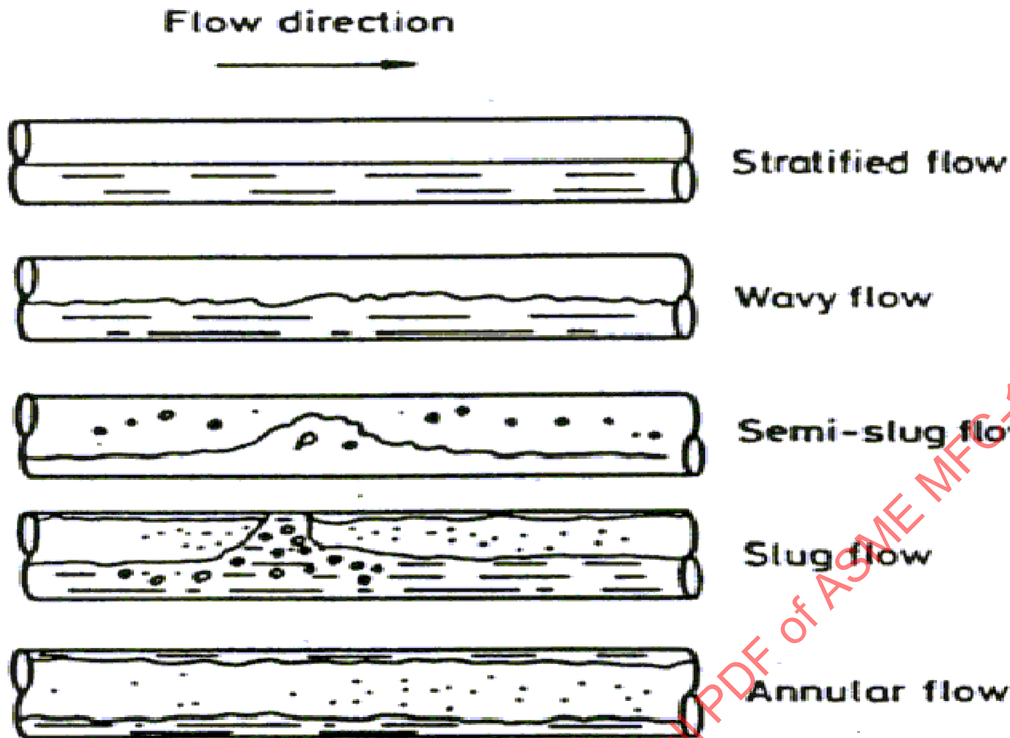
As the gas flow rate continues to increase for a set liquid-to-gas flow rate ratio, this liquid ring thins until there is nothing more than a very thin coating on the pipe wall and nearly all the liquid travels as droplets entrained in the gas. This is sometimes still called “annular-mist flow”/“annular-dispersed flow”/“dispersed flow”/“mist flow” (on account of there being no practical way to check if the ring of liquid is present or not). Typically, the higher the gas flow rate the smaller the average droplet size and the thinner any liquid film on the pipe wall. For horizontal flow there is always a gradient of droplet concentration where the higher concentration is at the base of the pipe. As pressure and gas flow rate (i.e., the liquid driving mechanism of the gas dynamic pressure) increases, the average droplet size reduces and the droplet dispersion throughout the pipe becomes more even.

In vertical upwards flow there are several general multiphase flow patterns but only three tend to exist in wet gas flows due to the relatively small volume of liquid flowing. These are “slug flow,” “churn flow,” and “annular-mist” flow.

In the event of a blockage in a pipe system caused by liquid gathering at a low point, gas pressure accumulates behind the liquid plug and eventually causes a “slug” of liquid (i.e., column of liquid filling the full cross-sectional area of the pipe) to be pushed up through the pipe work. This is “slug” flow.

For relatively high Lockhart–Martinelli parameters with low to moderate gas flow rates “churn” flow can occur. This flow pattern is unsteady in nature due to the constant gravitational force being countered by the continually varying gas dynamic forces (being applied to the liquid in a varying magnitude due to the continuing shifting spatial distribution of the liquid mass position and relative velocity to the gas stream). This vertical up flow pattern occurs due to the gas dynamic force being no more than the same order of magnitude to the liquid weight.

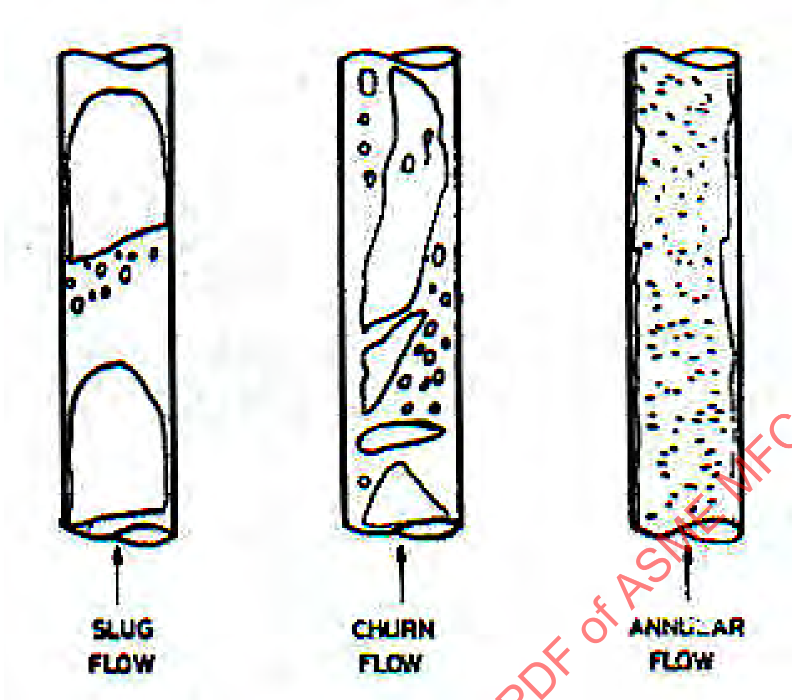
Fig. 4-1 Horizontal Wet Gas Flow Patterns



As the gas flow rate and/or the pressure increases for any given Lockhart–Martinelli parameter value, the gas dynamic forces increase to the point where the same flow pattern exists as in horizontal flow at high flow rates. That is, "annular-mist flow"/"annular-dispersed flow"/"dispersed flow"/"mist flow." Here the ring of liquid that exists with vertical flow is symmetrical due to the direction of the liquid's weight. As the gas velocity increases for a given liquid-to-gas ratio, the vertical up flow pattern behaves similarly to a horizontal flow pattern, and the liquid ring thins until there is only a very thin coating on the pipe wall and nearly all the liquid travels as droplets entrained in the gas. Again, typically the higher the gas flow rate, the smaller the average droplet size.

Vertical down flow pattern recognition is a poorly researched topic, and very little information exists in the literature on this subject. Most wet gas metering situations for vertical down flow are assumed to have "annular-mist flow" flow due to the gravitational and gas dynamic forces acting in the same direction. However, at low pressure and low gas velocity the liquid phase can be driven by its weight as much as, or even more by, the gas dynamic forces.

Fig. 4-2 Vertical Wet Gas Flow Patterns



Flow pattern knowledge of inclined flows is very rare. Little is published on this matter. However, it has long been recognized that with horizontal two-phase flows, slight positive inclination angles promote slug formation and slight negative inclinations promote stratification. A positive or negative inclination from the horizontal of as little as one degree can cause noticeable effects on the flow pattern.

It should be noted that in reality the boundaries between flow patterns are very difficult to judge. When viewing a two-phase flow it can be extremely difficult to decide which of the particular definitions fit best, as often the flow seems to be in continuous transition between the definitions. One of the basics of all multiphase and wet gas flows is that the flow patterns are generally unsteady flows, and are constantly changing, re-establishing themselves, and changing again. However, note that for many flows the overall flow rates remain approximately constant over time and so although the flow pattern is inherently unsteady at the metering point, by averaging the meter readings over a period of time it is often possible to obtain reasonably constant metering results. This fact is the reason that it is possible to develop wet gas flowmeters.

A further note on slug flow is required here. Note that in any gas flow system where liquid is present and there are low points in the pipe work, liquid will gather due to gravity. Even flows of gases with no more than trace liquids can have a significant liquid build up over time in low pipe work (or liquid trap) areas. When the liquid blockage increases, the resulting pressure rise behind the liquid can result in the liquid being suddenly forced down the pipe. Such liquid held up and then released with the gas is called a “slug.” Often the slug fills the pipe’s cross section completely. Thus in some wet gas metering situations, meters that appear to be operating in a steady wet gas flow with a steady flow pattern can suddenly be struck with very significant force by a slug of liquid. If flow exhibits periodic liquid slugs, it is said to be “slugging.” The condition of recurring and high-velocity slug creation by a flow is called “severe slugging.” System designers should be aware of the potential for slugs in flows with low liquid loading. In such cases single-phase meters are often used as the liquid-induced error on the gas flow rate prediction is low, but unlike actual single-phase flows, the pipeline components have to be able to withstand the force of periodic slug strikes. It should also be noted that in the oil and gas industry severe slugging is often the reality when a gas well has been “shut in” (and hence the pipe line flooded) and then reopened.

Unfortunately, even if slugging does not occur, the pipe work configuration in the vicinity of the meter could still have a direct effect on the flow pattern, so the choice of meter position (e.g., whether it is vertical or horizontal, at a high or low point in the pipe work or close to bends etc.) can affect the local flow pattern and therefore possibly the readings of some meter designs. This is analogous to the upstream pipe configuration for a single-phase meter affecting the flow profile and hence the meter's performance.

Predicting a wet gas flow pattern is not a precise science. Flow pattern prediction is difficult for meters installed in ideal locations (i.e., long straight upstream lengths) and is considered problematic in nonideal meter installations. Furthermore, in the nonideal locations the flow pattern can have been disturbed by an upstream pipe component and be in the process of returning to its natural undisturbed condition at the inlet to the meter. That is, it could be in transitioning while flowing through the meter. However, on a positive note, the little information that exists on this matter from the wet gas test laboratories suggests that a disturbance to a flow pattern by a pipeline component is damped out quickly (i.e., within a few pipe diameters).

The process of wet gas flow measurement in practical applications often requires that the measurement system be able to perform under a variety of flow patterns. In most industrial applications, however, there is no flow pattern determination (due to this determination being a complicated issue and a research topic in its own right). Furthermore, the flow regimes can change quickly with common operational interventions. For this reason, the development of wet gas and general two-phase (i.e., "multiphase") meters often includes checks that the system will operate in all flow patterns the system may encounter. As it is preferable to predict which flow patterns will exist over the expected range of flow conditions before choosing a method of metering, engineers can use the published flow pattern maps, although the predictions will be little more than a "best estimate."

5 FLOW PATTERN MAPS

A flow pattern map is a chart that attempts to predict the type of flow pattern expected under different flowing conditions. There is no set rule to what parameters the axis of such charts should use. This results in a myriad of different parameters used by different authors from the simple with clear physical meaning to the complex with obscure physical meaning. Many two-phase flow pattern maps have been created by different industries over the last few decades, but no single flow pattern map is regarded as the best for universal use. Flow pattern maps tend to be created from experimental observation and not from fluid mechanics theory. Typically they are created using data relevant to a particular industry, pipe orientation, set fluid types, specific pressure and temperature range, pipe size, phase average velocities etc., and hence their use can be limited. It is up to the individual engineer to make a judgment on which map is most suited to a particular wet gas metering application.

For all these maps the borders between flow patterns are not to be considered rigid. In reality the flow patterns do not change at any critical combination of parameters but rather tend to change gradually over a range of varying parameters. Therefore, if a flow condition in question is close to any border on a flow pattern map, then this indicates that the flow pattern at that point is likely to be transitional between the two flow patterns stated to exist either side of the boundary line. Figure 3 shows a sample of a popular flow pattern map used in the oil and gas industry. Note the axes are the gas and liquid densiometric Froude numbers [see eqs. (6) and (7)].

Figure 5-2 is a flow pattern map that shows another example of a horizontal pipe flow pattern map and is a reproduction of a map shown by API [2]. (Note that not all the patterns shown in this particular map, e.g., plug flow, are relevant to wet gas metering, as wet gas flow does not have the required liquid content with the gas to create such a liquid dispersion in the pipe.) Although in the majority of applications the meter operator does not have control over the changes in flow patterns, it is nevertheless important to recognize the impact of these flow patterns on the performance of a particular wet gas meter in question. The main use of flow pattern maps is to predict likely flow patterns that will be encountered.

Fig. 5-1 A Horizontal Flow Pattern Map (created by Shell Exploration and Production)

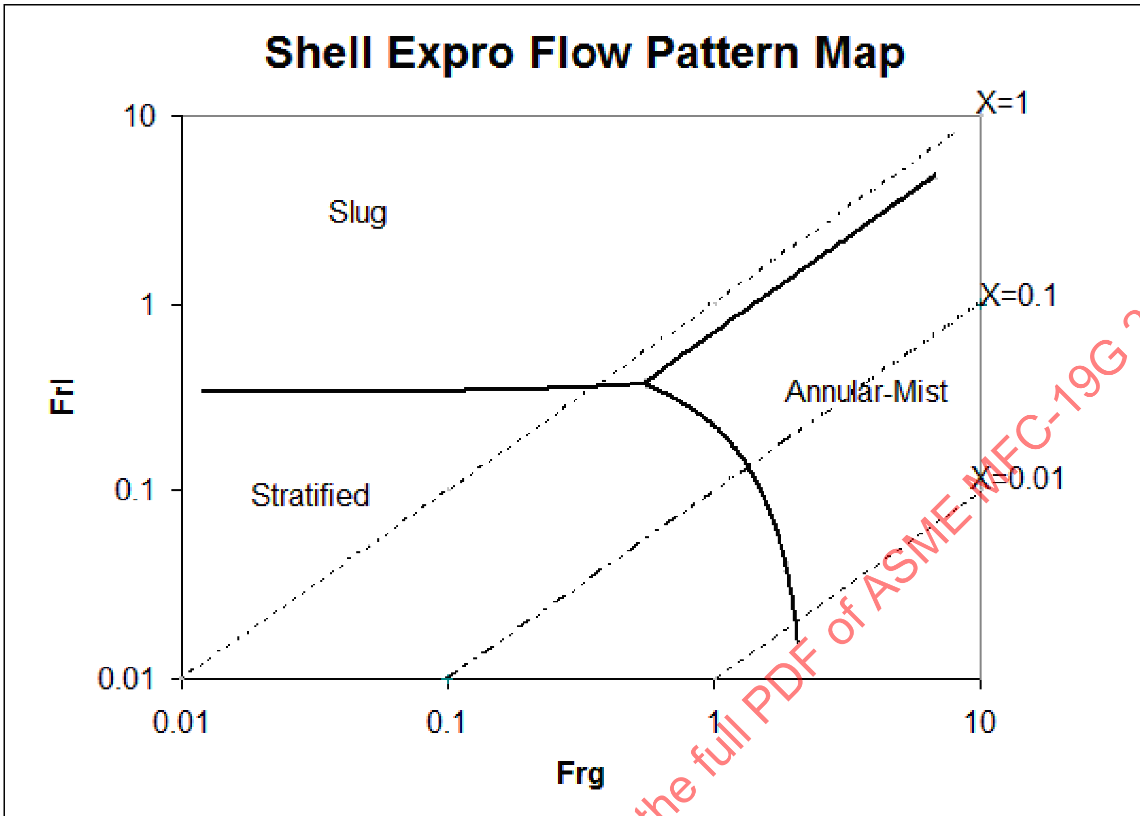
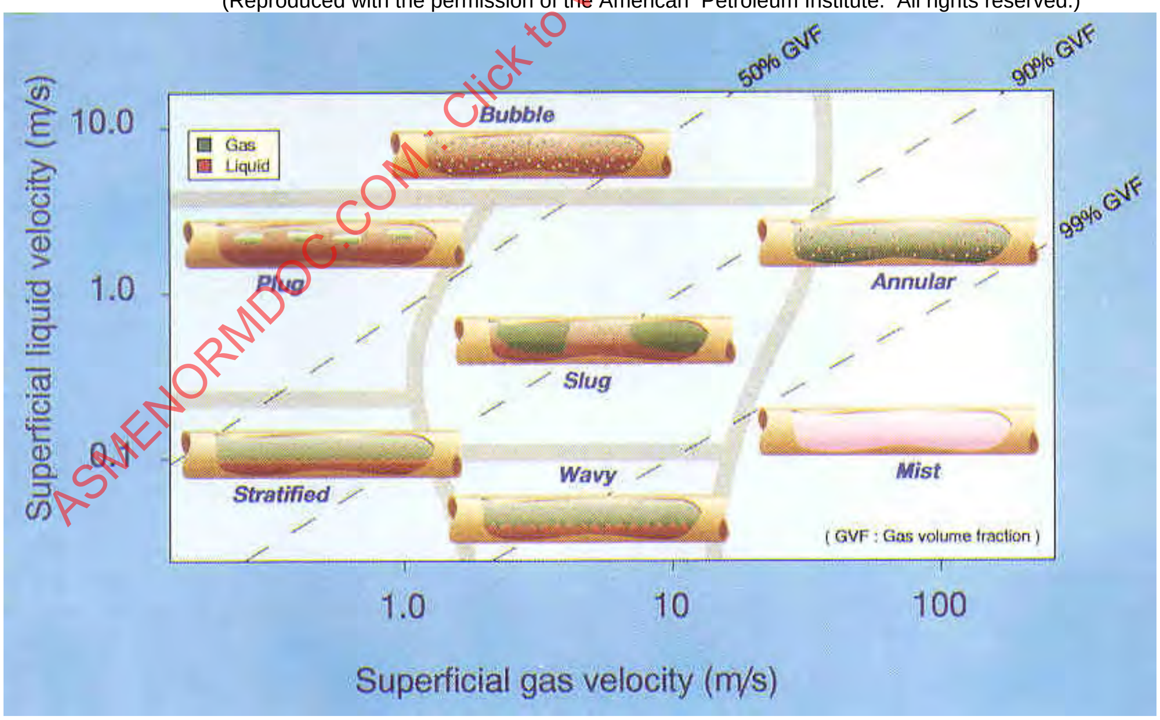


Fig. 5-2 A General Flow Pattern Map
(Reproduced with the permission of the American Petroleum Institute. All rights reserved.)



6 METERS USED WITH WET GAS FLOWS

Meters used with wet gas flows can be split into the following three general categories:

(a) *Category 1: Single-phase gas Flow Meters.* Single-phase gas meters are often applied to wet gas flows. There are generally two different applications for single-phase gas meter technologies when used with a wet gas flow. The first application is when the wet gas liquid loading is considered light by the meter user, and the liquid induced gas flow rate error is also considered acceptable by the user. Here the liquid flow rate is not metered or estimated. The meter's gas flow rate prediction is taken as the final gas reading, and the increase in uncertainty due to the liquid loading is accepted by the user. The second application is when the wet gas's liquid loading is heavier, and the liquid-induced gas flow rate error is considered to be significant by the meter user, and it is therefore required to be corrected for. When using a single-phase gas meter in these situations it is necessary to obtain the liquid flow rate or some form of liquid-to-gas flow rate ratio information from some source in order to correct the liquid-induced gas flow rate error using a wet gas flow correlation specified for that gas meter at those particular conditions.

(b) *Category 2: Wet Gas Flowmeters.* It is common practice in industry to define a "wet gas flowmeter" as a metering device that will predict both the gas and liquid phase simultaneously (i.e., in real time) with no liquid flow rate information being required to be supplied to the system by any source that is independent of the metering system, i.e., the primary, secondary, and tertiary devices. (The fluid properties are usually required to be supplied as with standard single-phase meters.)

Occasionally there is a necessity in industry to meter one phase flow rate of a wet gas flow while the other phase flow rate information is not so critical for the application. Currently, these applications tend to use wet gas meters as so far there are no meters known that predict the liquid flow rate alone in a wet gas flow. There are technologies appearing on the market at the time of writing that predict the gas flow rate alone in a wet gas flow but as yet they are new concepts and at the time of writing not widely used.

Wet gas meters tend not to be devices that can predict the water cut of the liquid phase. Typically, wet gas meters estimate the gas and liquid flow rates of a wet gas flow but give no information on the different components making up the liquid phase.

(c) *Category 3: Multiphase Meters.* Meters that estimate the gas and liquid phase flow rates and the fraction of different liquid components flowing simultaneously tend to be called multiphase meters. Multiphase meters are generally developed for the oil and gas production industry. Therefore, multiphase meters are metering devices that will predict the gas and liquid phase simultaneously and the water cut of the oil and water mix liquid component with no information being required to be supplied to the system by any source that is independent of the metering system (i.e., the primary, secondary, and tertiary devices) other than the fluid properties.

Traditionally the multiphase meters on the market have been aimed at general two-phase flows (i.e., $X_{LM} > 0.3$) encountered by the oil and gas production industry. The wet gas region (i.e., $X_{LM} \leq 0.3$) has traditionally been a difficult range for these multiphase meter technologies with uncertainties of results being significantly higher with wet gas than for general two-phase flow. However, in recent years multiphase meters have been developed to operate more successfully within the wet gas flow region. This development is continuing at the time of writing and the distinction between the wet gas and multiphase meter categories is becoming blurred. Wet gas meters that do not predict the water cut are not called multiphase meters. However, multiphase meters that can operate satisfactorily within the wet gas region (i.e., $X_{LM} \leq 0.3$) are sometimes also called wet gas meters.

With all three of these wet gas meter categories it should be noted that when discussing the alternative metering options there are metering technologies that are available to industry as commercial products and are now in actual service and metering technologies that are published conceptual designs that may or may not become commercial products.

Due to the commercial rivalry between manufacturers of different wet gas meter designs and the number of prototype meters in development, this Report establishes the following guideline:

(1) Meter types that have at least one working unit in service that has been bought by a user are considered to be commercial products available to industry.

(2) Meter designs that have some published research and offered to industry that have not yet been installed in a real application or meters that are installed in a real application but are on field trial for performance verification are considered to be "conceptual designs."

In this Report, only metering technologies that are available to industry as commercial products and are being used in service are discussed in the body of the Report. The Appendices discuss these technologies and also discuss metering technologies that are published conceptual designs that may or may not become commercial products in greater detail.

The following is an overview of the operating principle and performance of the dry gas meters used with wet gas correlations, wet gas meters, and multiphase meters in industry as claimed in information available to the public through journal papers, conference papers, and vendor press releases. All discussions are for incompressible steady state wet gas flow. ASME does not guarantee the veracity of the technical claims of any information quoted in this Report.

6.1 Single-Phase Gas Meters and Wet Gas Correlations

Single-phase meters can be split into those that use differential pressure (ΔP) techniques and those that use non-ΔP techniques. In general, ΔP devices are seen as the most robust and repeatable type of single-phase gas flowmeter for wet gas flow applications. However, other single-phase meters show promise of being successfully developed to operate with wet gas flow and on occasions any single-phase meter can be exposed to a relatively small quantity of liquid within a predominantly gas flow and a knowledge of the effect that this has upon the gas flow measurement can be important. Liquid flowing with a gas will cause a single-phase gas meter to have a liquid-induced error. To correct for this, the response of the meter to different liquid loadings must be found by experiment to be repeatable and predictable, and this response must be expressed by a wet gas correction factor (or "correlation" or "algorithm"). The amount of liquid present in the gas stream must be measured by some means. The following discussion attempts to cover most of the common types of single-phase gas meters that have been applied to date in wet gas flow measurement.

6.1.1 Single-Phase Gas Differential Pressure (ΔP) Meters. The differential pressure (DP) flowmeter is a device that utilizes the physical laws of the conservation of mass and energy. A reduction of the cross-sectional area of a conduit will ensure by these physical laws that the flow will increase velocity and the pressure will reduce. The difference in pressure before and after the change of area is therefore related to the velocity of the flow and therefore the flow rate. The physical geometry that causes the change of area is called a "primary element." The different standard[1] DP meters on the market all operate with the same generic flow equation, and the different parameters in this generic flow equation are all dependent on the primary element used. Common primary elements are orifice plates, nozzles, cones, etc. The wet gas flow response of a DP meter is dependent on the primary element used.

Differential pressure (DP) meters are historically the most commonly used meters in wet gas flows as a direct consequence of the orifice plate meter being one of the most common and widely used industrial flowmeter. The sustained drive to develop wet gas flowmetering technology is a relatively recent occurrence. For many years, due to lack of alternatives or due to general ignorance of the potential problems, orifice plate meters (with or without drain holes) were used for wet gas metering. Since the late 1950s a number of DP meter wet gas (and general two-phase) flow research papers have been published. Most of these wet gas flow papers were for orifice plate meters but there are several papers discussing the Venturi, nozzle, and cone-type DP meters. There are no known papers discussing the characteristics of eccentric orifice plate and wedge meters when used to meter wet gas flows, although these may well prove to be adequate wet gas meters. Much of the orifice plate meter data concerns wet steam with a selection of other fluid types while most Venturi and cone-type DP meters have wet gas data from experiments aimed at the natural gas production industry. These DP meter

[1] Laminar flow elements and pitot static based technologies do not come under this general description and are out of the scope of this Report.

papers indicate that the majority of the research is in agreement and that DP meters have a similar response to wet gas flow.

It is generally accepted that liquid in a gas flow causes the differential pressure from the DP meter to be higher than that, which would be indicated if the gas phase of the wet gas flow flowed alone. There is therefore a positive error (often called the "overreading") associated with DP meters when they are used with wet gas flows.[2] The parameters influencing the overreading have been major research areas for DP meters. The published literature has indicated that the overreading for a given wet gas flow condition is not the same for the different primary elements. Individual DP meter types and geometries have unique wet gas overreading properties.

It was first suggested by Schuster [4] that an orifice plate meter read high when the gas was wet. Murdock [5] then effectively showed that the orifice plate meter has an overreading dependent on the Lockhart–Martinelli parameter (although Murdock did not call the parameter by this name). Chisholm and Leishman [6] showed that a nozzle meter also has an overreading, dependent on the Lockhart–Martinelli parameter (in the only known nozzle meter wet gas research paper). Chisholm also advanced orifice plate meter knowledge [7,8] by adding that the overreading was also dependent on the pressure of the flow (or the gas-to-liquid density ratio). Lin [9] confirmed the Chisholm findings. De Leeuw [10] found these orifice plate meter findings were valid for Venturi meters also and added further knowledge by showing that the gas densiometric Froude number was also influencing the overreading of the Venturi meter when the flow pattern was annular-mist flow. Steven [11] confirmed de Leeuw's findings for Venturi meters and that the same parameters affected the overreading of cone-type meters [12–15]. As of 2006 there is no known research publication that discusses the gas densiometric Froude number (Fr_g) effect on the orifice plate meter.

Figure 6.11 shows a reproduction of Murdock's original plot (with the x-axis converted to the Lockhart–Martinelli parameter). The y-axis shows the ratio of the meter reading to the reference meter reading. This type of single-phase gas meter wet gas flow data plot is often referred to as a "Murdock Plot."

Figure 6.1.1-2 shows typical Murdock plot for a Venturi meter [11]. The plot shows NEL wet gas data for a 6 in. 0.55 beta ratio Venturi meter in terms of positive percentage error (or "overreading") to the Lockhart–Martinelli parameter.

Figure 6.1.1-3 shows the same data highlighting the three different pressures tested. The legend in the graph gives the gas-to-liquid density ratio (DR) adjacent to each test pressure. Clearly there is a *DR* effect with higher pressure (or higher *DR*) having a lower overreading for a set Lockhart–Martinelli parameter (*XLM*).

There is some scatter evident in the constant pressure results in Fig. 6.1.1-2. It can be seen from Fig. 6.1.1-3 with the 40 bar data that the scatter is due to a Fr_g effect. The graph shows that higher values of Fr_g give a higher overreading.

All published Venturi meter and cone-type DP meter data shows these same trends with DR and Fr_g numbers. As of 2006 no other DP meter types are known to ASME to have been tested with wet gas flows to this level and the results published. It is currently assumed that all DP meters follow these general trends.

Experimental data for a particular DP meter can indicate the approximate percentage error. If the liquid loading is sufficiently small to give an acceptable gas flow rate error for a particular application, then in practice an increased gas flow rate uncertainty without applying any wet gas flow correction may be acceptable.

Where the liquid flow rate or some form of liquid-to-gas flow rate ratio is known, or can be estimated, an equation of the form shown as eq. (29) can be applied to the single-phase equation.

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[2] Some orifice plate meter papers (e.g., [3]), for very low liquid loading, contradict the generally accepted relationship that liquid causes a positive error in the gas flow rate estimation. One proposed reason for this is considered to be a lubrication effect of the trace liquid on the meter tube and orifice plate in which the effective surface roughness has been marginally reduced.

$$\dot{m}_g = \frac{EA_t Y C_d \sqrt{2\rho_g \Delta P_{tp}}}{f\left(X_{LM}, \frac{\rho_g}{\rho_l}, Fr_g\right)} \tag{29}$$

The numerator in eq. (29) is the generic single-phase equation for any DP meter in question with the appropriate constants for that particular geometry. The denominator represents the particular wet gas correcting function fitted to that particular DP meter for those particular conditions. The precise nature of the function " f " is found by experimental data obtained for particular fluids and meter geometries where *XLM*, *DR*, and Fr_g are known. Historically, not all correction factors include all three of these parameters.

It should be understood that all forms of these single-phase gas DP meter correction factors require the liquid flow rate or liquid-to-gas flow rate ratio to be specified or derived. That is, note

Fig. 6.1.1-1 Reproduction of Murdock's Two-Phase Flow Orifice Plate Meter Plot

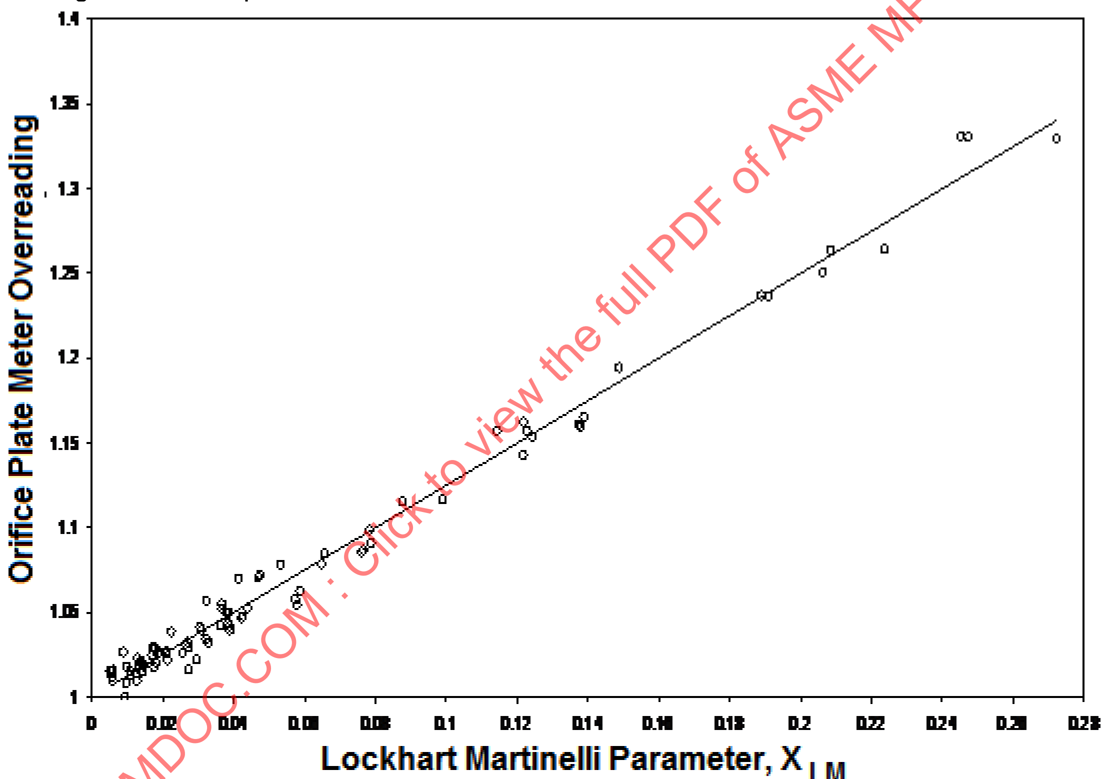

Fig. 6.1.1-2 Wet Gas Flow Venturi Meter Data

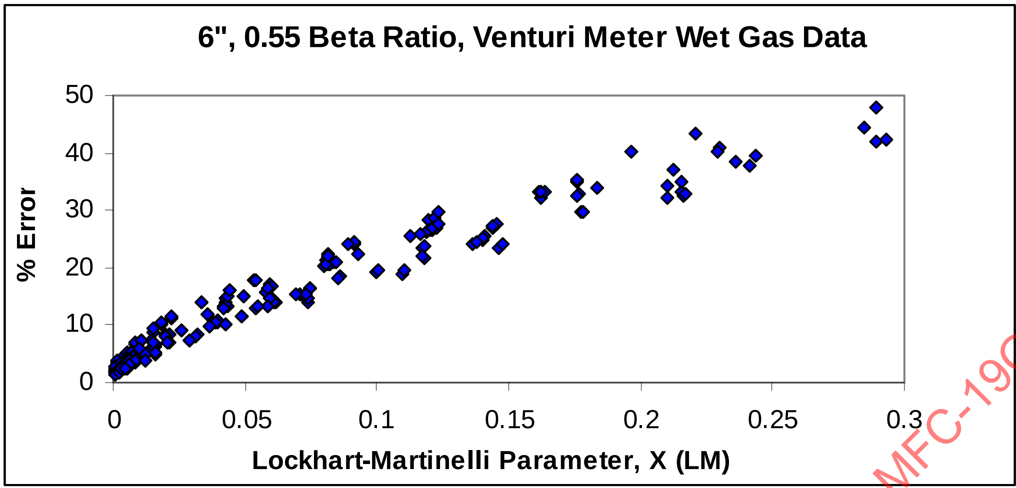

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Fig. 6.1.1-3 Wet Gas Flow Venturi Meter Data With Separated Pressure

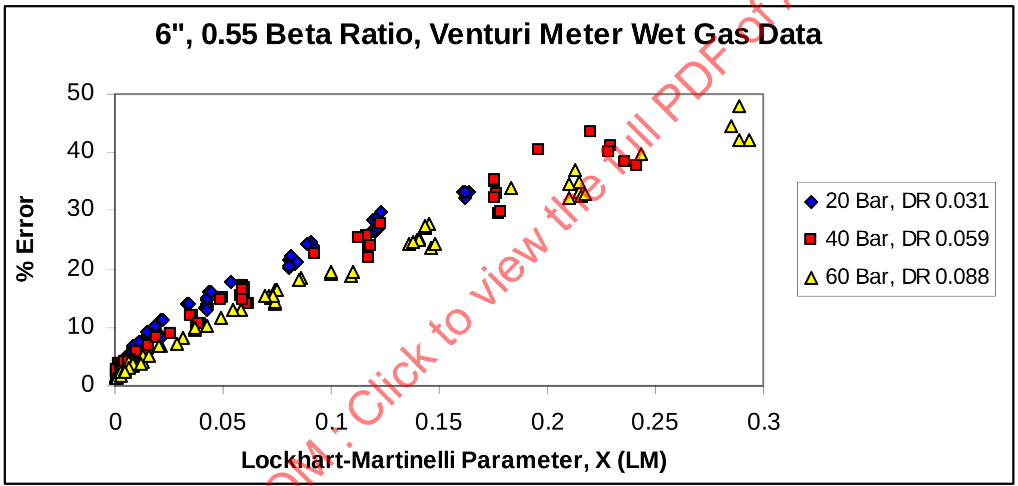
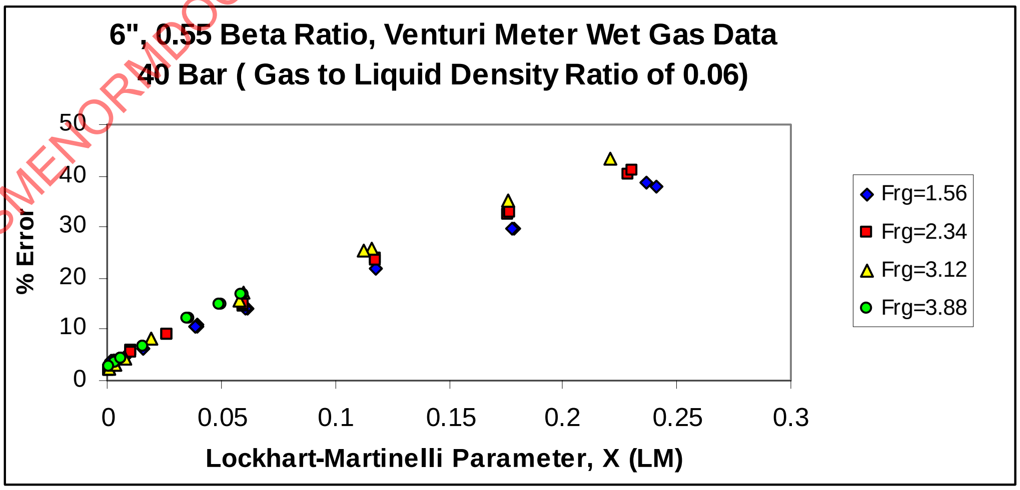


Fig. 6.1.1-4 Gas Flow Venturi Meter Data With Separated Fr_g



the denominator in eq. (29) includes the Lockhart–Martinelli parameter [which is expressed in eq. (30)].

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} \quad (30)$$

Hence, for known gas and liquid densities, solving eq. (29) by iteration of the gas flow rate [noting Fr_g is a function of the gas flow rate — see eqs. (3) and (6)] requires direct knowledge of X_{LM} [which is the same as knowing liquid-to-gas flow rate ratios — see eq. (30)], or, the liquid flow rate information needs to be known directly.

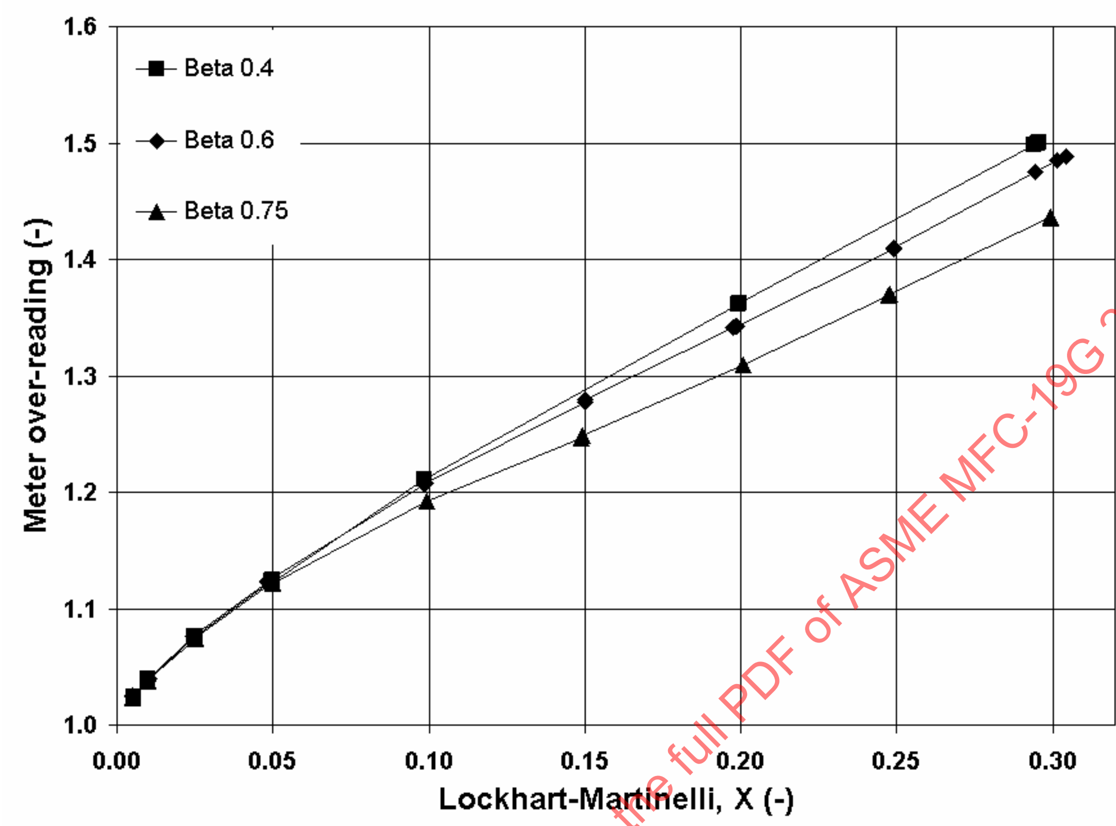
Typically an uncertainty in the gas mass flow rate prediction of 2% is quoted for a known X_{LM} /liquid flow rate. The gas flow rate result uncertainty quoted in most of the literature is generally for knowing the liquid flow rate to a low uncertainty (i.e., the liquid reference meters of the wet gas flow test center). Increased uncertainties in this liquid flow rate prediction add to the gas flow rate uncertainty quoted in these cases.

It should be noted that along with the Lockhart–Martinelli parameter, gas-to-liquid density ratio and the gas densiometric Froude number, the beta ratio (β) of DP meter-type meters needs to be considered, as this has been reported to also have an effect on the overreading by Stewart et al. [12, 16]. That is, data gathered at NEL for a series of wet gas cone type meters and Venturi meters (β = 0.4 to 0.75), show that there is a further dependence of the overreading on the β-value itself. Figure 6.1.1-5, for example, shows the difference in a 4 in. Venturi meter overreading measured at three values of β, for a fixed pressure (i.e., gas-to-liquid density ratio) and gas velocity (gas densiometric Froude number). The overreading differs by as much as 7% (in absolute terms) at the highest Lockhart–Martinelli parameters tested (which was the wet gas limit of X_{LM} = 0.3). The larger the beta ratio (the smaller the primary element relative to the pipe size) the smaller the overreading for a set wet gas flow condition.

Influences of significant meter diameter and fluid property variations on wet gas flow performance are the latest topics of research at the time of writing. Relatively little data has been released on these topics.

Reader-Harris [17, 18] showed that the gas properties in a wet gas flow do not affect the DP meters wet gas overreading (other by dictating the gas density at a given pressure). Liquid property effects have been addressed by Reader-Harris [17, 18] and Steven [15, 19, 20]. These initial studies suggest that for horizontal wet gas flows liquid properties affect flow patterns and therefore wet gas overreading of DP meters. At relatively low pressures and gas flow rates (i.e., stratified flow) there is no appreciable difference between a DP meter's reaction to wet gas with hydrocarbon liquid and a wet gas with water. Both types of liquid will cause the meter to overread the gas flow in a similar way. At relatively high pressures and gas flow rates (i.e., transition between stratified and annular mist flow) there is an appreciable difference between a DP meter's reaction to wet gas with hydrocarbon liquid and a wet gas with water. The water tends to remain stratified until higher gas dynamic forces are reached and the overreading for a water based wet gas flow then lags a light hydrocarbon liquid based wet gas flow for the otherwise identical flow conditions. That is the gas/water wet gas flow has a lower overreading versus Lockhart–Martinelli parameter gradient than the gas/light hydrocarbon liquid. NEL [17, 18] has also shown as pressures and gas flow rates increase further this difference in gradient diminishes. Figures 6.1.1-6 through 6.1.1-8 reproduce sample NEL graphs.

Fig. 6.1.1-5 NEL Wet Gas 4 in. Venturi Data for 31 bar(a), Fr_g=1.5



Steven [19, 21] has discussed very preliminary findings with regard to the effect meter diameter has on the DP meter wet gas overreading. Initial theoretical and data set analysis suggests that a diameter effect may exist. The smaller the DP meter for set geometries (including beta ratio), Lockhart Martinelli parameter, gas-to-liquid density ratio, gas densiometric Froude number, and similar liquid properties the smaller the overreading. Figure 6.1.1-9 shows the graph presented by Steven [21]. It must be remembered that one data set is all that has been analyzed. Therefore, Steven states that this is as yet only a postulation and not a proven DP meter wet gas flow phenomenon.

It should be noted that the experimental tests used to develop the correction factors do not usually document the flow patterns. However, provided the correction factors are applied to similar meter geometry and fluids, it can be assumed that flow pattern influences in the DP meter performances are included. Most DP meter wet gas correlations are from horizontal meter runs; consequentially, some vertical installations may be forced to use the horizontal correlations. It is unknown what amount of error this will cause.

If a method of estimating either the liquid flow rate or liquid-to-gas flow rate ratio to an acceptable uncertainty is available then the use of a single-phase gas DP meter and a wet gas correlation can be an economical way of metering the gas flow in a wet gas flow. However, care must be taken in choosing the wet gas correction factor to be applied as all the available correlations are all for specific geometries

Fig. 6.1.1-6 NEL 4-in., Schedule 80, 0.75 Beta Ratio Venturi Meter, Gas-to-Liquid Density Ratio of 0.046, Gas Densiometric Froude Number of 1.5

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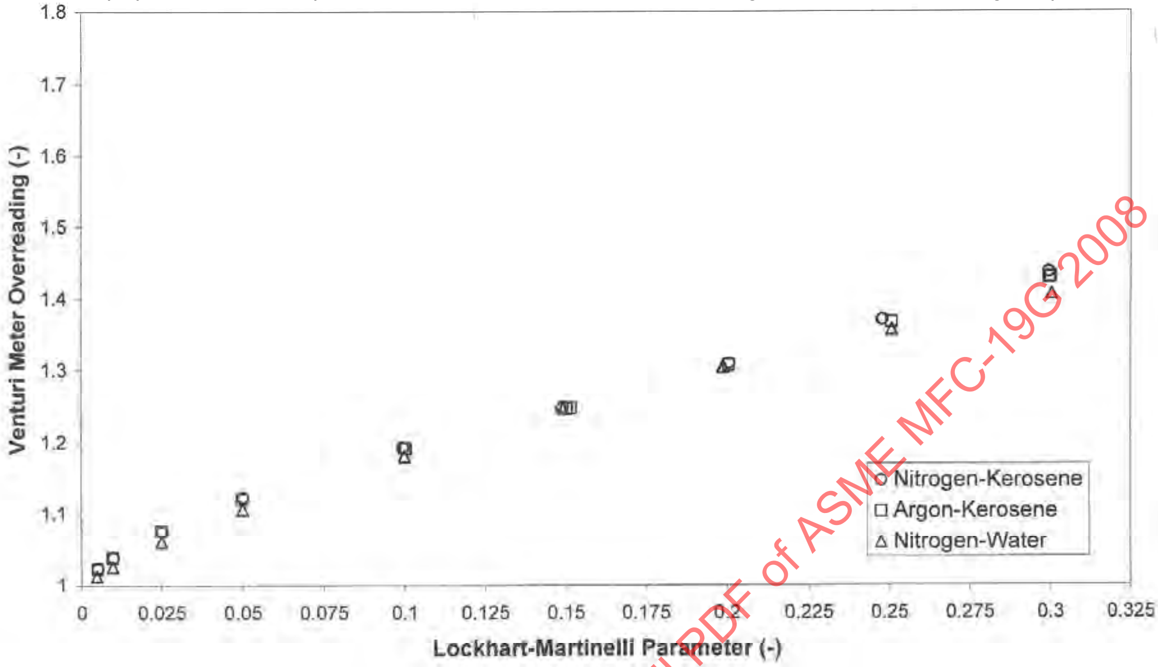

Fig. 6.1.1-7 NEL 4-in., Schedule 80, 0.75 Beta Ratio Venturi Meter, Gas-to-Liquid Density Ratio of 0.046, Gas Densiometric Froude Number of 2.5

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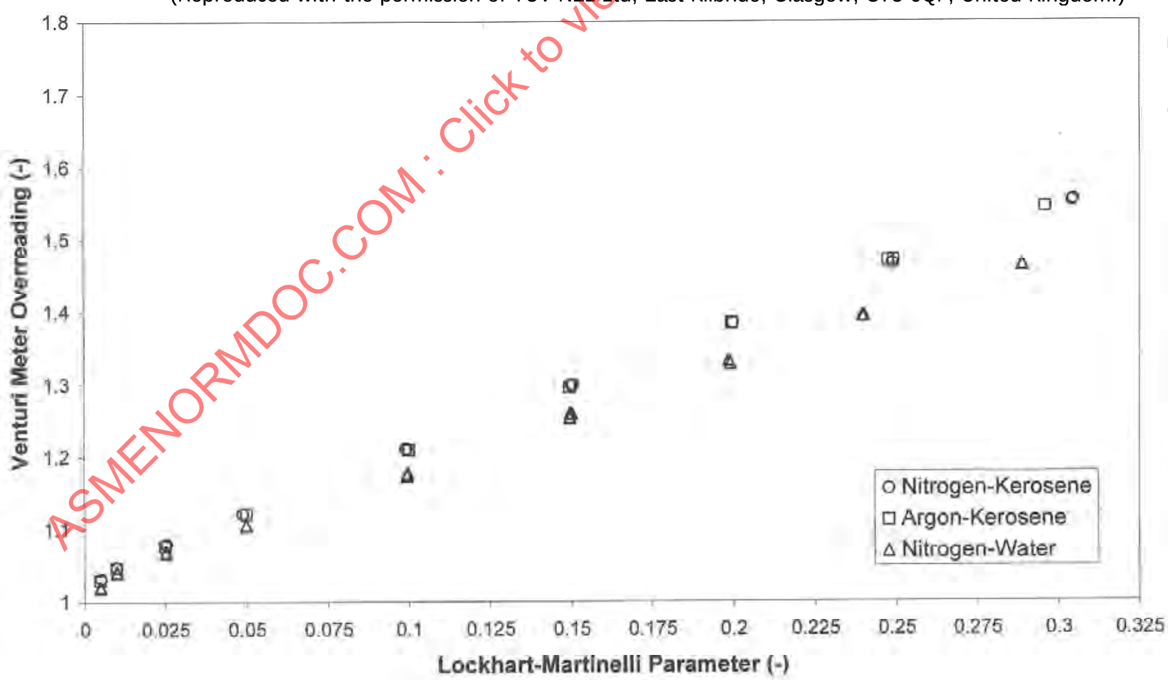

Fig. 6.1.1-8 NEL 4-in., Schedule 80, 0.75 Beta Ratio Venturi Meter, Gas-to-Liquid Density Ratio of 0.046, Gas Densiometric Froude Number of 4.5

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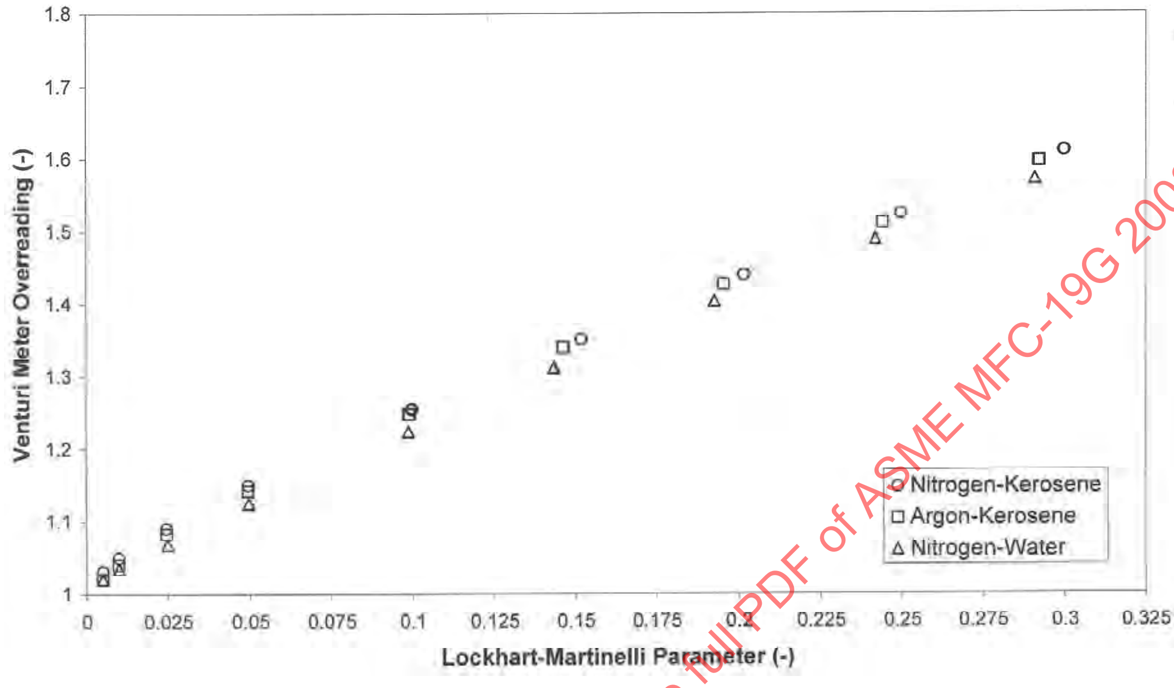
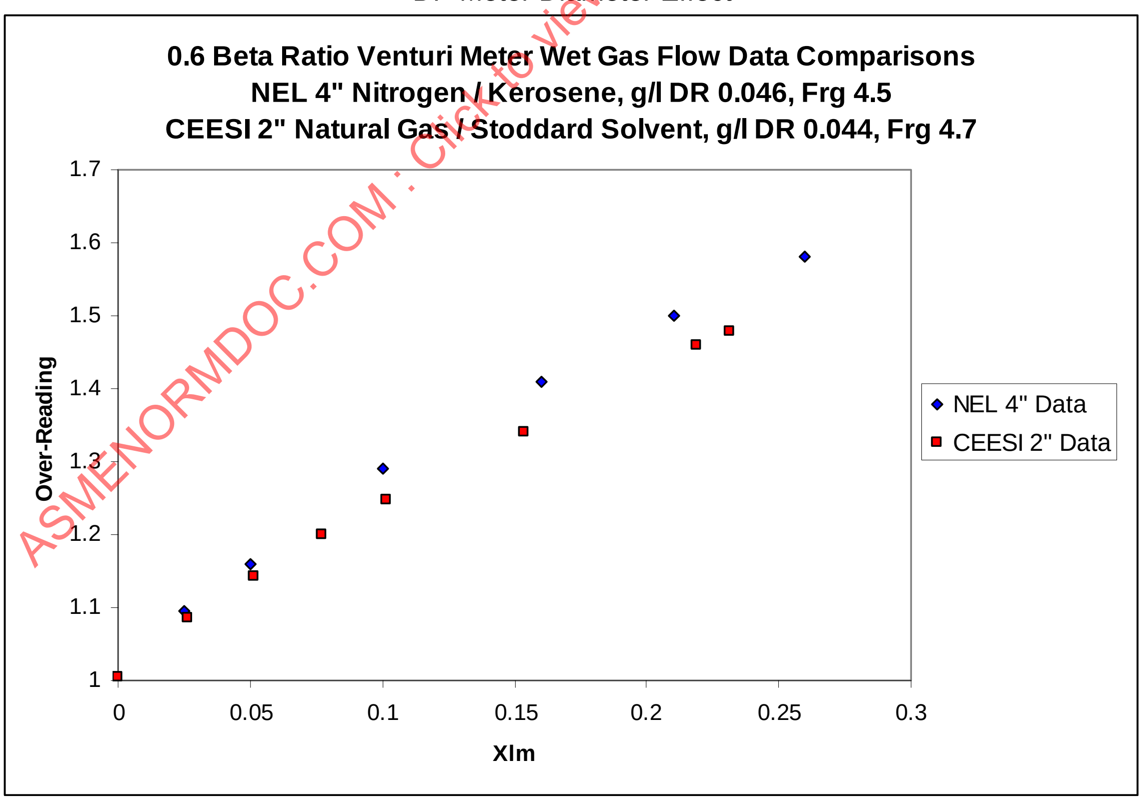


Fig. 6.1.1-9 4-in. and 2-in. Venturi Meters With Similar Wet Gas Flows Showing a DP Meter Diameter Effect



of DP meters and for a specific range of flow conditions. Use of any particular wet gas correlation on a different geometry DP meter (e.g., different type/meter size/beta ratio) and/or at a flow condition outside of the data set parameter range used to create the correlation (e.g., different liquid type/gas-to-liquid/density ratio/gas densiometric Froude number/Lockhart–Martinelli parameter) would lead to additional uncertainties in the gas flow rate prediction. It should further be understood that most wet gas correlations have been developed with test meters in long, straight runs of pipe. Installation effects for meters with wet gas are largely unknown and therefore disturbances close to the inlet may increase the uncertainty of the resulting gas flow rate prediction.

Details of many of the published correlations are given in Nonmandatory Appendix F.

6.1.2 Non-Δ*P* Single-phase Gas Meters. Other dry gas meters, which do not rely on differential pressure measurements, exhibit different responses to the presence of liquid in the flow. The most common types are turbine, vortex shedding, Coriolis, and ultrasonic meters, which are discussed below.

6.1.2.1 Turbine Meters. It is generally accepted that turbine meters should not be a meter of choice for wet gas flows due to the adverse effects on the meter caused by the liquid phase. However, as with most single-phase gas meters there are instances in industry where turbine meters have been subjected to low liquid loading wet gas flows; it is therefore of relevance to discuss turbine meter performance under such conditions.

Gas flow turbine meters measure the volume flow rate of a gas flow by passing the flow over a centrally mounted turbine rotor to drive its rotation. Sensors measure the angular velocity of the rotor in the form of counts as it rotates. As the gas flow rate increases the rotational velocity increases. The sensor count appears as a frequency that is directly related to the gas flow rate via the meter factor (usually called the K-factor as it is algebraically denoted by the letter “K”). A gas turbine meter’s “K-factor” is found by calibration at a gas flow test facility. Due to manufacturing tolerances each gas turbine meter has to be individually calibrated.

Only two papers by Ting [22] and Stewart [23] and one technical report by NEL [24] (from which Stewart’s paper is a summary of the turbine meter section of that report) have been published that discuss research on the performance of a turbine meter when subjected to a light liquid load wet gas flow in horizontal flow. (There is no research known regarding turbine meter wet gas performance in vertical wet gas flow.) The two existing papers do not fully agree.

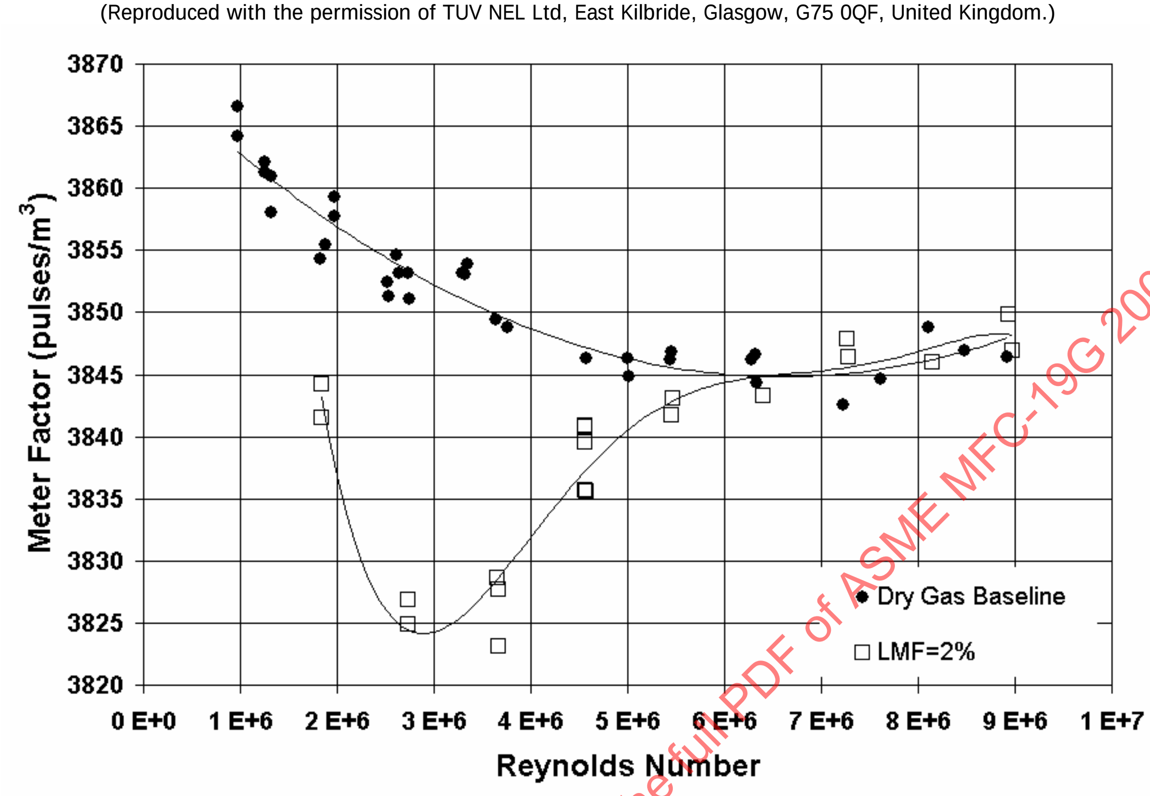
Stewart reports a Lockhart–Martinelli parameter range of $0.0006 \le X_{LM} \le 0.0070$ (for nitrogen and kerosene at 61 bara up to a Reynolds number to nine million at ambient temperature).

Ting reports a Lockhart–Martinelli parameter range of 0.00012 (0.07 bbl/MMSCFD) to 0.000255 (0.14 bbl/MMSCFD) for air and water at 51.6 bara up to a Reynolds number to ten million at ambient temperature.

Crucially both papers report that the liquid-induced error is less than 1% and hence there is agreement that a turbine meter will continue to work when it is temporarily exposed to trace liquids. However, there are considerable differences in the details of these gas turbine meter’s responses to the liquid’s presence.

Stewart/NEL [23, 24] shows that under wet gas conditions the 6 in. gas turbine meter tested by NEL in nitrogen/kerosene has a tendency (at least for the lower values of Reynolds number) to “underread” the gas flow rate. (“Underreading” is a term used to describe a negative error induced by the liquid on the meter’s gas flow rate prediction. It is analogous to “overreading,” which describes a positive error induced by the liquid on the meter’s gas flow rate prediction.)

Fig. 6.1.2.1-1 NEL/Stewart's Turbine Meter Wet Gas Response for Liquid Mass Fraction of 2%
(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom.)



The error increases with liquid mass fraction. The response of this turbine meter to the presence of liquid in the flow stream is shown in Fig. 6.1.2.1-1.

In Figure 6.1.2.1-1 Stewart/NEL shows a plot of pulses per volume of gas (i.e., the K-Factor) versus Reynolds number for the liquid loading value of 2% liquid mass fraction (LMF). Clearly there is a liquid-induced error at Reynolds numbers below six million. At Reynolds numbers greater than this the liquid-induced error has disappeared (which is not the same finding reported by Ting).

The behavior of the turbine is not fully understood. Stewart recorded the flow patterns seen by an installed camera and notes that the liquid-induced error magnitude coincided with the type of flow pattern.

At *Re* $< 1.8 \times 10^6$ the flow was reported to be completely stratified. The liquid running along the meter tube base would oppose the motion of the blades and hence reduce the rotor's angular velocity. Such a small amount of liquid (<2% LMF) would not cause a significant blockage to the gas and hence not cause a significant increase in gas velocity. Therefore in this stratified flow pattern the meter underread the gas flow. Note, however, that the maximum measured underreading (at 2% LMF) for the stratified flow pattern was reported to be less than 0.5%.

For $2.7 \times 10^6 \leq Re \leq 5.4 \times 10$ the flow was reported to be in transition between stratified and annular-mist flow. The fluctuating nature of the flow pattern introduced a greater level of scatter into the data. The underreading increased significantly with liquid content reaching a maximum value (for LMF = 2%) of 0.75%. This indicates the rotor was slowed by liquid resistance to a greater extent than it could accelerated by the increased gas dynamic force due to the liquid blockage.

At *Re* $> 5.4 \times 10^6$ the flow approached annular-mist flow. In this region, as the gas velocity increases, slip reduces as the size of the liquid droplets reduce and they become more evenly distributed throughout the flow. The result of this was for *LMF* ≤ 2% the turbine meter behaved as a single-phase meter.

Ting et al. [22] showed a polynomial curve fit on the dry gas data for a 6 in. gas turbine in Fig. 6.1.2.1-2 (i.e., the K-factor to the flows Reynolds number curve usually known as the "turbine curve") and the calculated K-factors for the individual wet gas flows conditions tested at CEESI. Note that the units that were chosen here to describe the liquid loading are barrels (i.e., volume) of liquid per million standard cubic feet of gas. Up to a Reynolds number of approximately five million there is no significant liquid effect, and beyond this, a small K-factor shift is evident. The plot of the K-factor deviation found between the wet and dry gas flows is reproduced in Fig. 6.1.2.1-3.

The liquid-induced error reported by Ting is an underreading (as it was with the NEL tests reported by NEL/Stewart). However, significant differences in the reports are described below:

For a maximum liquid loading of X = 0.000255 Ting found an underreading of 0.43% compared to NEL/Stewart finding at X = 0.007 an underreading of 0.75%. (It is noted in this Report that at these extremely small liquid loadings the repeatability of each turbine meter in dry gas becomes an issue when analyzing and comparing data.)

As the Reynolds number increased for a set liquid loading the underreading increased and then leveled off in Ting's tests while it increased and then decreased with Stewart's tests.

No wet gas turbine meter correlation is known to have been published.

Clearly more work is required in order to understand the precise nature of a gas turbine meter's relationship with light liquid loadings as the two available reports have contradictions. However, crucially both agree that for very light liquid loadings ($X \leq 0.007$) a liquid-induced error of less 1% will occur. Ting concludes that "… a Turbine meter could be used in the short term as a master meter for in-situ wet gas proving but that turbine meters should not be used in continuous unprocessed (i.e., wet) gas operations."

A long-term disadvantage of gas turbine meters being used with wet gas flow is their susceptibility to mechanical damage. Blades can be damaged by impact with the liquid phase, wet gas flows are seldom clean, bearings can be contaminated, and the blades of a gas turbine are relatively fragile and therefore susceptible to damage by slugs. For these reasons the current designs of turbine meters are an unlikely choice to deliberately place in a wet gas environment.

Fig. 6.1.2.1-2 Ting's Turbine Meter Wet and Dry Gas Flow Rate Results at CEESI

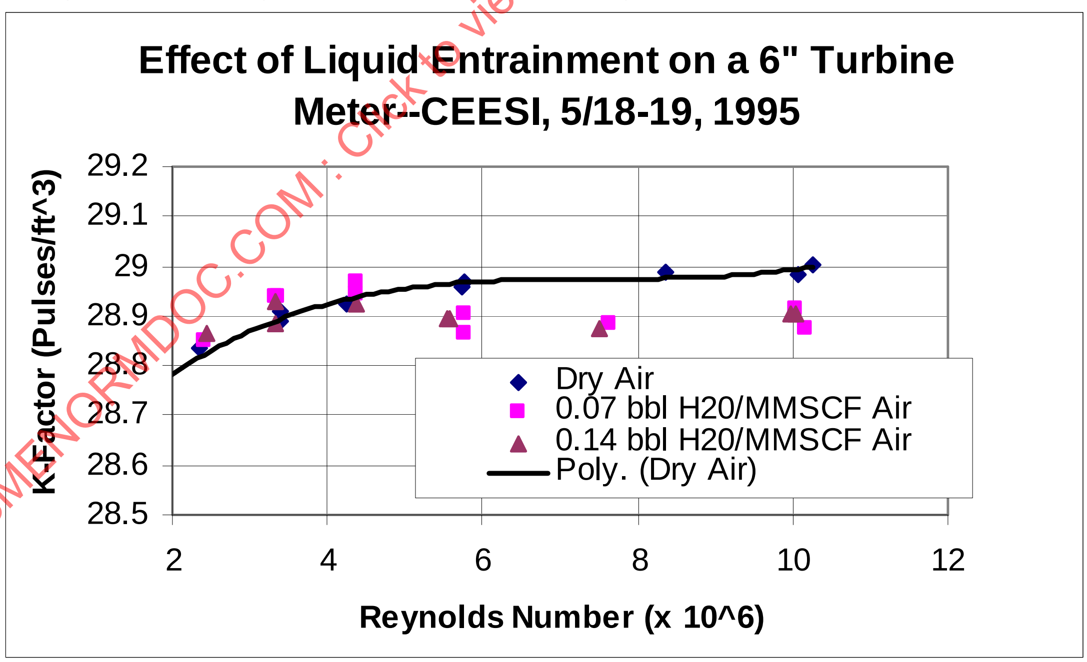
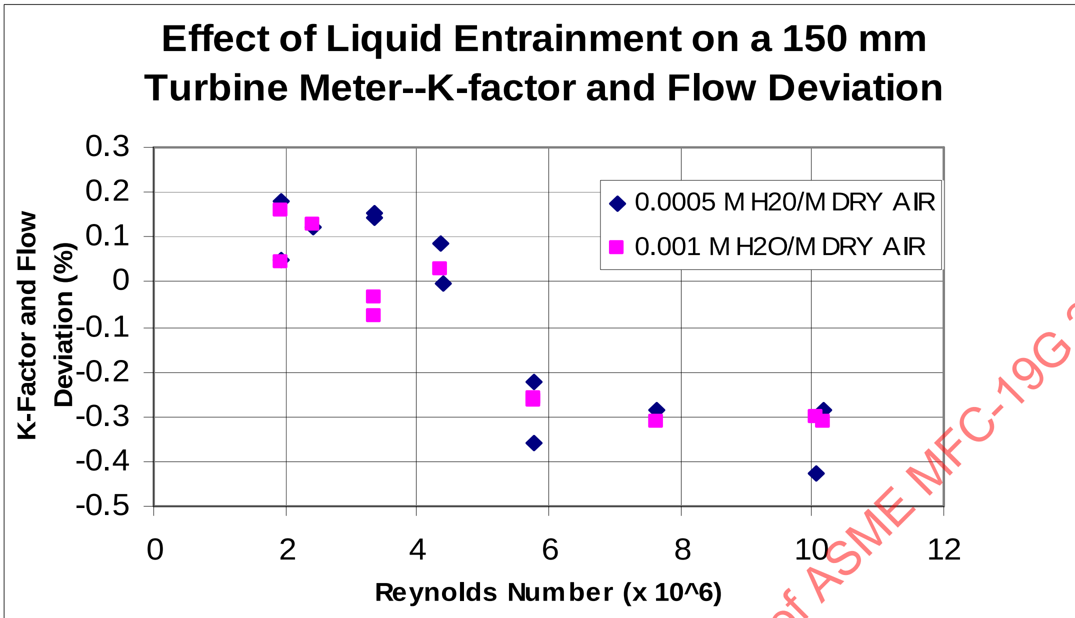


Fig. 6.1.2.1-3 Turbine Meter Wet Gas K-Factor Deviation Results



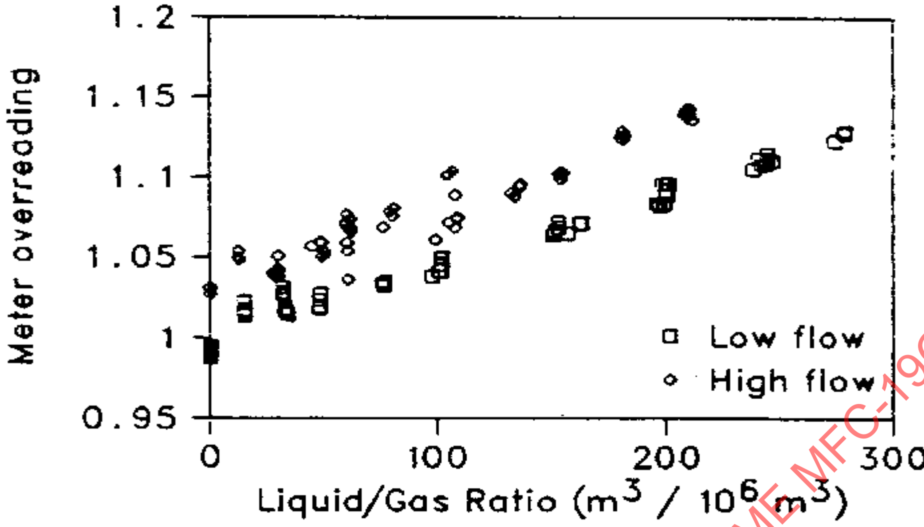
6.1.2.2 Vortex Shedding Meters. The gas flow vortex meter operates on the principle of a gas flow passing a bluff body (i.e., a square or triangular bluff body) producing a series of alternate vortices shed from either side of the body in turn. Behind the bluff body a "von Karman vortex street" is created. Sensors measure the frequency of this vortex shedding, and this frequency is directly related to the gas velocity via a meter factor (usually denoted by the letter "K") and hence the volume flow rate.

A limited amount of wet gas flow data has been published for vortex shedding meters. These devices are known to have been tested periodically with wet gas flow by Joint Industry Projects (JIPs) and individual companies but little of this work exists in the public domain. However, four published papers [25, 26, 24, 23] are cited here.

Washington et al. discusses in two related papers [25, 26] results of analysis on vortex shedding meters in the field and in a laboratory in wet gas flow. (The meter's bluff body design was not disclosed.) Washington reported that horizontally mounted 75 mm (3 in.) vortex shedding gas meters overread with wet gas (i.e., the liquid induces a positive error on the gas flow rate) and that the overreading is related to the liquid loading and the gas flow rate. Figure 6.1.2.2-1 shows Washington's presentation of the results. Low and high gas flow rates indicate 85,000 and 135,000 Nm^3/D, respectively. Note the liquid loading is represented as actual cubic meters of liquid per normal million cubic meters of gas. (The maximum Lockhart–Martinelli parameter here is approximately 0.1.)

Washington states the results were not repeatable at lower liquid loadings. It is stated that between laboratory and field tests the results were different and this was reported to be due to the vortex shedding meter overreading being dependent on the flow pattern. Washington concludes, "...liquid increases the reading of a velocity-type flowmeter in gas service. The actual increase depends on the slip between the liquid and gas phase and the resulting liquid hold up in the line. This in turn depends on the installation geometry, and the gas and liquid flow rates and properties."

Fig. 6.1.2.2-1 Washington [25, 26] Field Data for Wet Natural Gas Flow



Washington [25, 26] also states that vertical down wet gas flow tests showed that a vertical installation "…is not a practical way of reducing the effect of entrained liquid on a gas vortex meter."

No wet gas flow Vortex Shedding meter correlation was offered by Washington et al.

NEL and Stewart [23, 24] released the only other published wet gas flow vortex shedding meter research information. (The meter's bluff body design was not disclosed, but it is known the vertically installed meter had the bluff body parallel to the ground.) It is stated that this work "…tended to agree qualitatively with previously published data (Washington et al.); however the absolute overreading values are much lower for a given LGR [i.e., liquid-to-gas ratio]." Stewart discussed test data and analysis for three pressures and many gas and liquid flow rates. Stewart introduces a level of doubt on the gas velocity (or gas densiometric Froude number) effect previously reported by Washington as no predictable gas velocity effect was found with the new laboratory data. Figure 6.1.2.2-2 shows a sample graph of the 30 bar data.

Figure 6.1.2.2-3 shows the three pressure data sets (15, 30, and 60 bar) separated. The individual pressure data sets maximum Lockhart–Martinelli parameters plotted are the limits for which the meter appears stable (i.e., beyond these values the meter overreading was seen to become extremely erratic as is seen when considering the 30 bar data set presented in Fig. 6.1.2.2-2.) There is clearly a pressure effect. The higher the pressure, the lower the overreading. The large spread in results for each set pressure and Lockhart–Martinelli parameter combination is due to the reported erratic gas flow rate effect. In Fig. 6.1.2.2-3 linear fits to the capped Lockhart–Martinelli parameter data are shown. The "cap" is at the maximum Lockhart–Martinelli parameter before the meter response is judged to go unstable.

Figure 6.1.2.2-4 shows the reported results of using these linear fits for each pressure across the stable Lockhart–Martinelli parameter region to correct the gas flow rate for a known liquid flow rate. (The gradients of the liner line equations were not offered.) The correlations when applied to the individual pressure sets that were used to create them show a gas flow rate uncertainty of "…mostly within 2% and all within 5%."

The only Vortex meter wet steam paper known is the Hussein and Owen paper [27]. This paper describes tests of a 2 in. Vortex meter with wet steam at two static pressures (4 and 6 bar) and (quality $x \geq 0.84$, Lockhart–Martinelli parameter $X_{LM} \leq 0.012$). Hussein and Owen plotted the data as quality to overreading. Again it is reported that as liquid loading increases the overreading increases, although the magnitude of the error is large compared to the other reported research. The correction offered is the multiplication of the uncorrected Vortex meter wet gas flow rate result by the factor $1/\sqrt{x}$, where x is the flow quality [see eq. (16)]. This gives an estimated total flow rate. There is no literature regarding independent checks of this correlation and the correlation uncertainty is unknown.

For wet gas flowmetering applications where the liquid flow rate or some form of liquid-to-gas flow rate ratio is known to be (or can be estimated to be) relatively small experimental data has shown it is likely the error will be proportionately small. For higher liquid loadings with larger liquid-induced errors it may be possible to produce a useable wet gas correlation for any particular vortex meter up to a stated maximum Lockhart–Martinelli parameter. This maximum Lockhart–Martinelli parameter may be dictated by instability in the reading and not the maximum desired limit for an application. As with DP meters the choice of any

Fig. 6.1.2.2-2 NEL Nitrogen/Kerosene 30 bar Vortex Shedding Meter Data
(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom.)

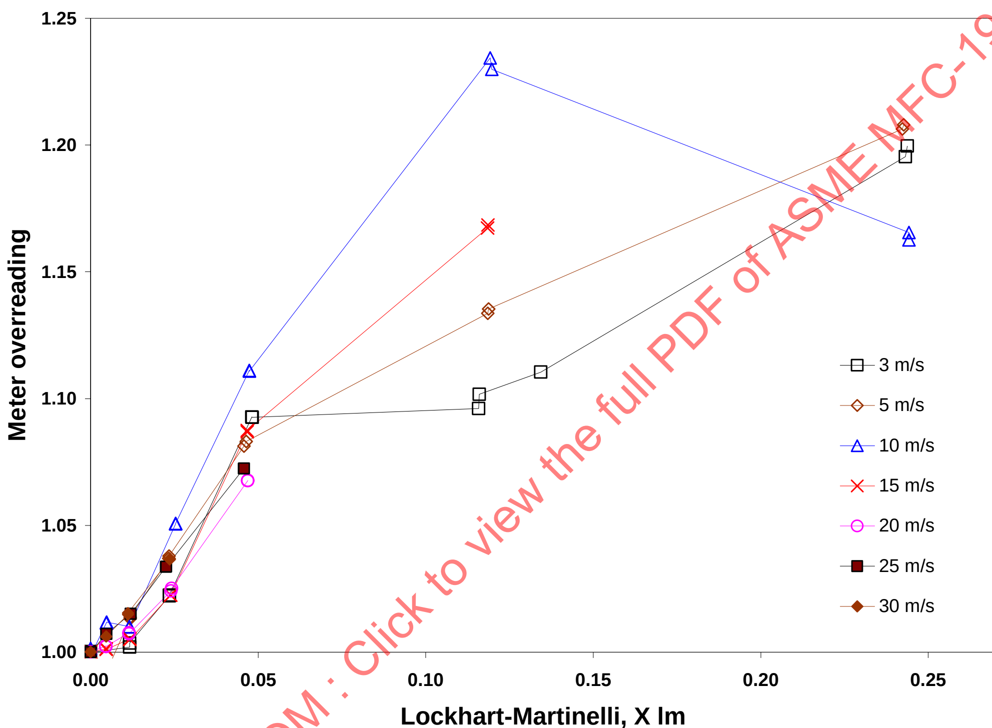


Fig. 6.1.2.2-3 NEL Nitrogen/Kerosene Vortex Shedding Meter Data Capped at Maximum Lockhart–Martinelli Parameters Before Data Becomes Erratic

(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom

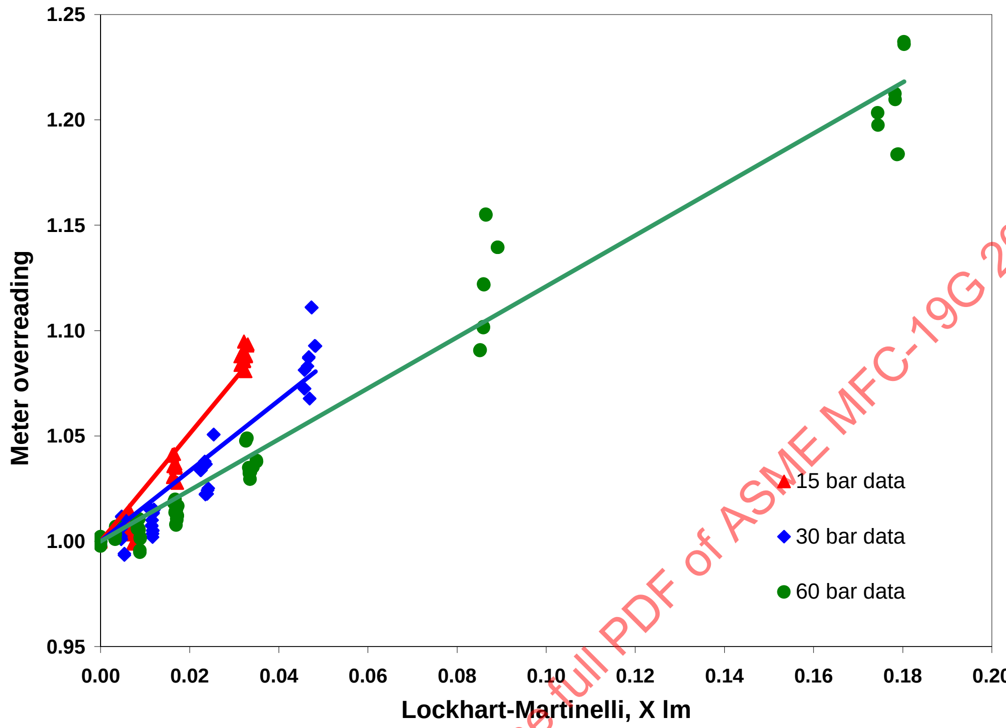


Fig. 6.1.2.2-4 Results of the Linear Fit Wet Gas Correlations Presented in Fig. 6.1.2.1-2 for Known Liquid Flow Rates

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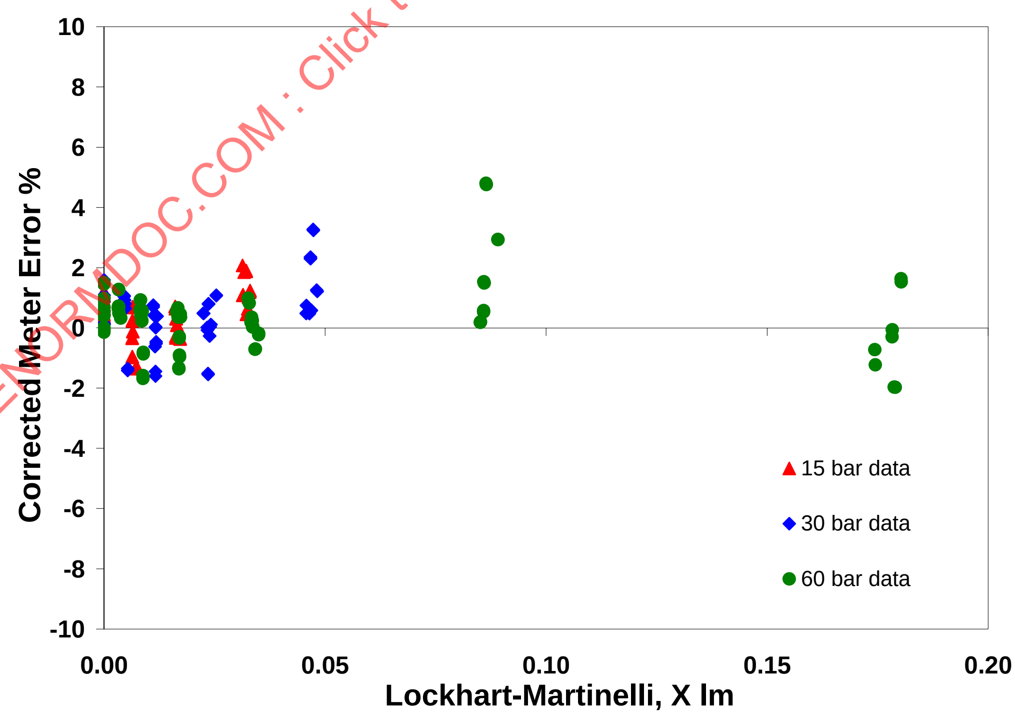


correlation would need to be considered with care. In reality with such limited data available individual meter geometries would need to be tested with wet gas flow and a wet gas correlations specially created from that data. The same caution on the extrapolation of such a correlation beyond the data set limits that created it is given here as was given concerning DP meters.

6.1.2.3 Coriolis Meters. The Coriolis flowmeter is a mass flowmeter that uses the effect of the Coriolis force, due to fluid mass flow on a forced conduit vibrating tangential to the flow direction. The produced Coriolis force is directly proportional to the fluid mass flow, and this produces a strain on the pipe (i.e., the meter spool). Measurement of the strain (i.e., meter spool distortion) therefore indicates the measurement of mass flow rate. The various different designs of Coriolis mass flowmeters all work according to this same principle. Coriolis mass flowmeters are good meters in single-phase gas or liquid flowmetering applications.

There is limited information in the public domain regarding the performance of Coriolis meters when they are used with wet gas flows. There are a few technical papers with some information on Coriolis meters with two-phase flows (i.e., $X_{LM} > 0.3$) but none of these cover the range defined by this Technical Report as wet gas flow. There are only two wet gas flowmeter research reports [24, 110] and an associated technical paper [23] known to the authors. NEL and Stewart [23, 24] report the independent wet gas tests on a 4 in. Endress + Hauser Promass 63F Coriolis meter. Wet gas flow tests are reported up to 5% LVF at 31 bar (a maximum Lockhart–Martinelli parameter of 0.24) and at 61 bar (a maximum Lockhart–Martinelli parameter of 0.18). The 31 bar results are reproduced in Fig. 6.1.2.3-1.

Clearly, as with most meters, liquid presence in the gas flow significantly affects the flowmeter's ability to predict the gas mass flow rate. The magnitude of the gas flow rate error was found by NEL to be related to the pressure (or gas-to-liquid density ratio), the liquid loading (or the Lockhart–Martinelli parameter), and the gas flow rate (or the gas densiometric Froude number). However, Stewart does not report any correlation being created that could predict the error for a known liquid flow rate or liquid-to-gas flow rate ratio.

As the Coriolis meter is a device designed to measure the mass flow and density of a single-phase flow, NEL plotted the meter's total mass flow error versus the actual total mass flow. Figure 6.1.2.3-2 presents the 30 bar results. The error increases with the Lockhart–Martinelli parameter. Whereas the gas flow rate error could be positive or negative (see Fig. 6.1.2.3-1) NEL reports the total mass flow rate error as always negative. It is shown in Fig. 6.1.2.3-2 that, in all but the highest liquid loading, increasing the gas flow rate for a set pressure and Lockhart–Martinelli parameter reduces the total mass flow rate error. It was noted that the error was marginally less at higher pressures. NEL stated that these findings are evidence of the flow pattern moving towards annular mist (i.e., the two phases becoming a single homogenized flow where the meter should operate successfully as a single-phase meter). However, the errors at the highest gas flow rates tested are still substantial, suggesting full mist flow with atomized droplets (i.e., a homogeneous mix) is not achieved at these flow conditions.

From NEL's tests across the full wet gas range, it could be concluded that the exact behavior of Coriolis meter designs with wet gas is not yet fully understood. Stewart concludes: "…the meter exhibited significant errors in wet gas flow compared with both the dry gas reference mass flow and the total reference mass flow."

However, Stewart also stated in conclusion: "…many modern meters have the built in capacity to sense the presence of liquid from the increased power drawn by the drive coil due to liquid damping and can stop measuring temporarily while liquid passes."

Britton et al. [110] showed wet gas flow test results from CEESI on two 2 in. Coriolis mass flowmeter designs (i.e., an Endress + Hauser Promass 83F and a Micro Motion CMF design). After excellent single-phase test runs the meters were tested in wet gas flow. The test conditions were as follows:

(a) nominal static pressures: 175 psig and 500 psig
(b) nominal gas velocities: 80 ft/sec to 160 ft/sec
(c) nominal liquid loads: concentrating on ≤25% by mass (with a few higher values)

Fig. 6.1.2.3-1 NEL 4-in. Coriolis Meter 30 bar Wet Gas Data
(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom

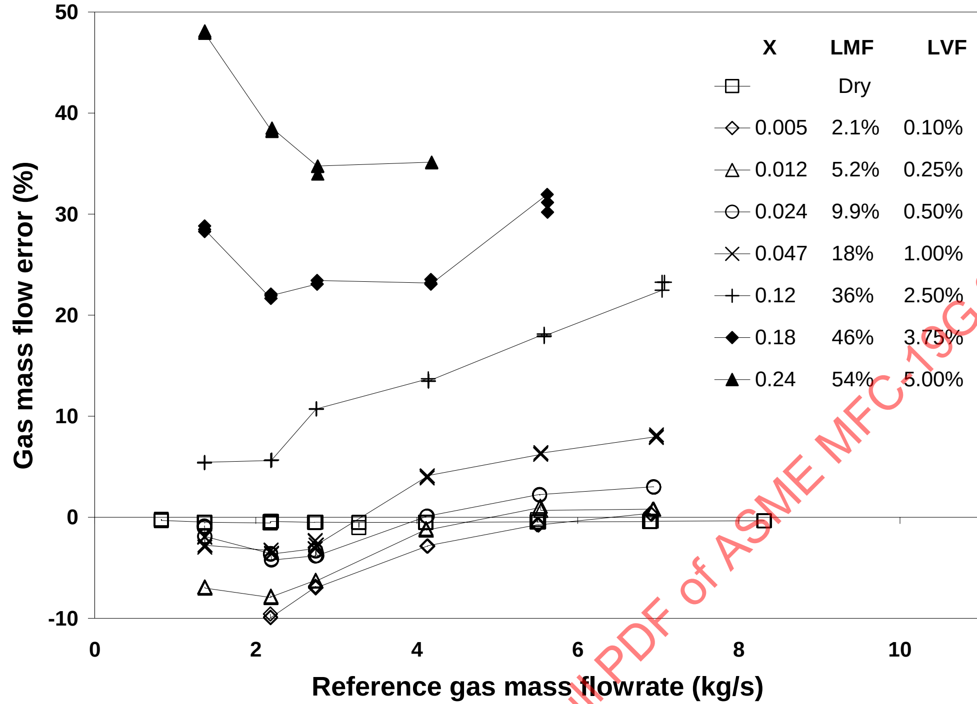
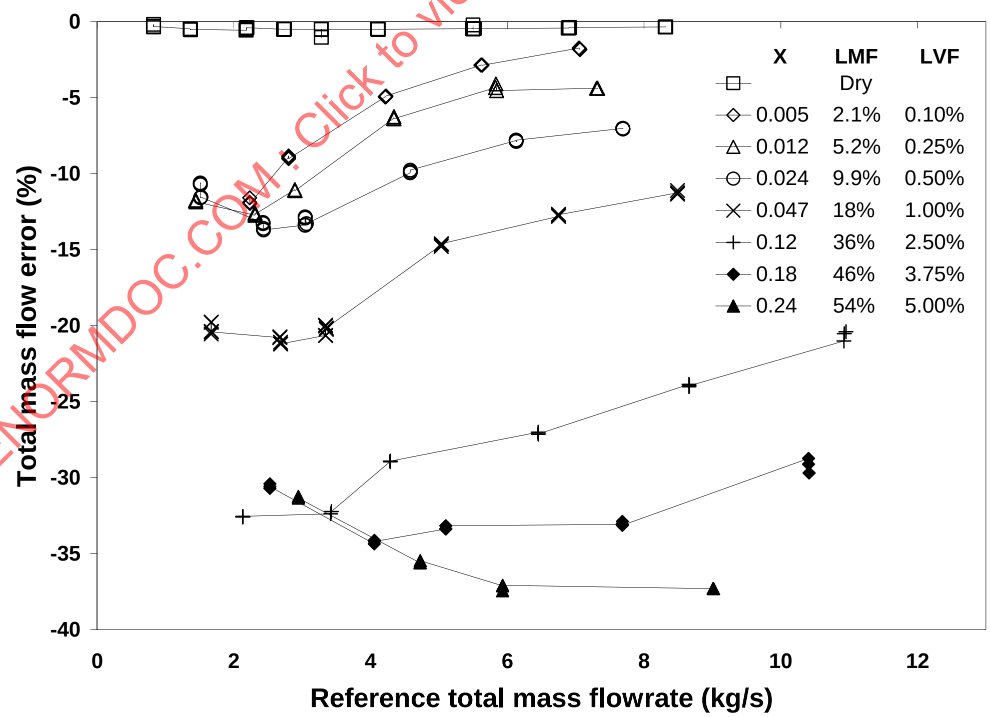


Fig. 6.1.2.3-2 NEL 4-in. Coriolis Meter 30 bar Total Mass Flow Rate Wet Gas Data
(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom



The equivalent Lockhart–Martinelli parameter range of these tests was concentrated on $X_{LM} \leq 0.05$ (although a few higher values were recorded as seen in Fig. 6.1.2.3-3.) The gas velocities are very high compared to the tests reported by NEL. Britton [110] suggests the test data is likely to be for a mist/homogenized flow pattern in these cases.

Both 2-in. meters (Micro Motion and Endress + Hauser) are reported to have behaved in a similar manner. Figure 6.1.2.3-3 shows the Micro Motion meter data plotted by Britton [110] as Lockhart–Martinelli parameter versus "Percent Deviation" (i.e., the overreading). The meter has a stable and predictable wet gas error for $X_{LM} \leq 0.035$ (which corresponds to a liquid-to-gas mass flow rate ratio of approximately 20% at the higher pressure). In this range, the wet gas overreading is approximately linear to the Lockhart–Martinelli parameter as with DP meters, although the scale of this overreading is higher.

Figures 6.1.2.3-4 and 6.1.2.3-5 show in more detail the low liquid loading data from both types of Coriolis meter. The comparison shows that both designs have a very similar reaction to the liquid's presence with increasing Lockhart–Martinelli parameter up to 0.035 meaning an approximately linearly increase in overreading. For these low liquid loading data sets both meters show a pressure (or gas-to-liquid density) effect. The lower the pressure the higher the overreading. A wet gas correlation that could predict the gas flow rate for a known liquid flow rate appears possible for a low liquid loading but as yet none has been published.

Britton et al. [110] concluded: "A Coriolis meter will begin to produce a reasonably consistent over-registration result for Lockhart–Martinelli numbers less than 0.035 (or liquid loads less than 20%). The magnitude of over-registration is dependent upon meter type and model number.

The accuracy of a Coriolis meter will remain within 1% for wet gas flowing conditions as long as the liquid load is less than 1% or the Lockhart–Martinelli number is less than 0.0012."

Fig. 6.1.2.3-3 2-in. Micro Motion Coriolis Flowmeter Wet Gas Test Data

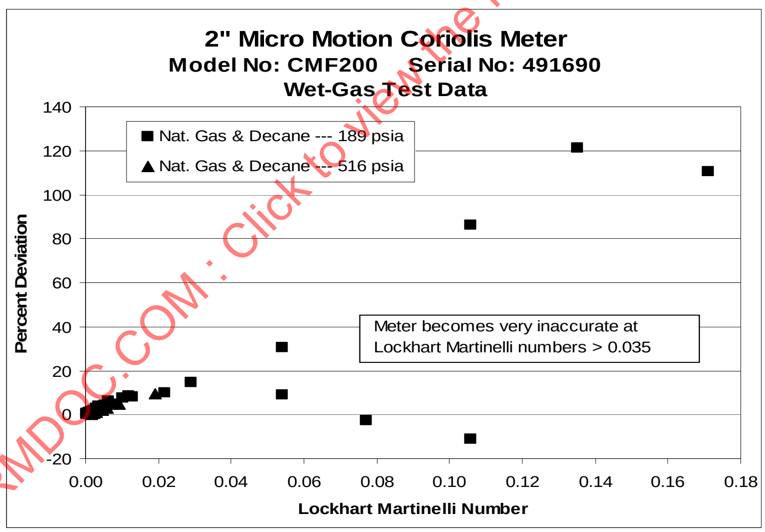


Fig. 6.1.2.3-4 Endress + Hauser Coriolis Meter, $X_{LM} < 0.035$

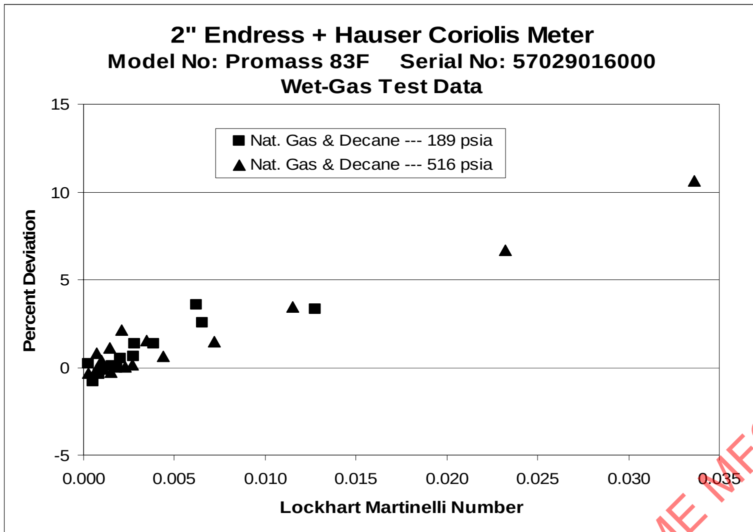
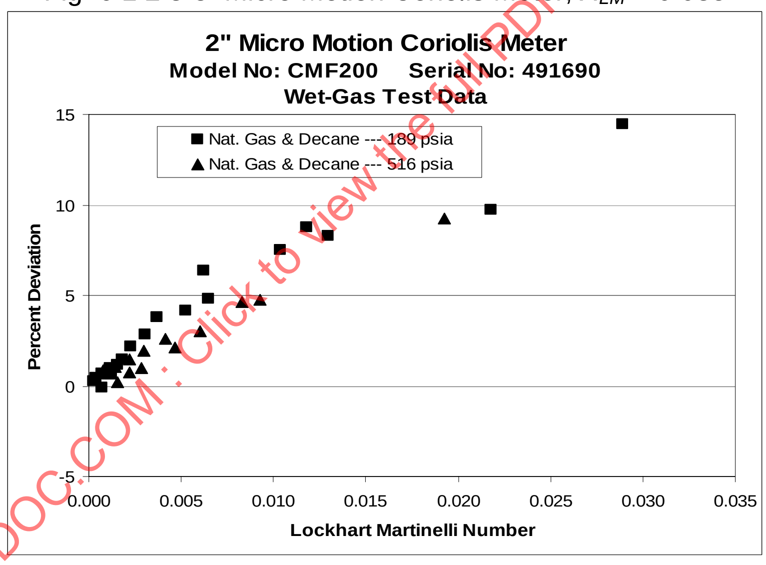


Fig. 6.1.2.3-5 Micro Motion Coriolis Meter, $X_{LM} < 0.035$



"The wet-gas data presented in this Report was obtained at nominal gas velocities of 80 ft/sec and 160 ft/sec. At these gas velocities, the liquid dispersal in the gas stream is most likely in "mist" form. At lower gas velocities the liquid dispersal will become what is known as "stratified" or "annular." Under these conditions, proportionally more liquid is traveling along the wall for the same liquid loadings reported herein. An accumulation of liquid on the vibrating tubes of the Coriolis meters will most likely affect the meter's performance. It is suggested that a test program be initiated to determine the lower gas velocity limits applicable to the various designs of Coriolis meters….," and "…As yet no such research is known to have occurred."

6.1.2.4 Ultrasonic Meters (USM's). The transient time ultrasonic gas meter transmits and receives acoustic waves sent diagonally across the fluid flow in both directions, with and against the flow, and measures the respective transit time of flight of waves in each direction. The difference in the transit times is proportional to the gas velocity along that path. Multipath meters can have different numbers, locations, and orientations of the paths and they can be direct path (transducer to transducer) or reflected (transducer to wall to transducer). The path velocities are then integrated, by various proprietary mathematical methods, to find the average flow velocity and hence the actual volume flow rate from the cross sectional area of the meter.

The ultrasonic gas meter can produce a large amount of diagnostic data in addition to the volume flow rate. This diagnostic data has traditionally (since the development of the concept into a commercial gas flowmeter in the early 1980s) been used in dry gas applications to assure the user of the correct operation of the USM. As of the 1990s there was considerable discussion within the oil and gas industries with regard to whether these USM diagnostic capabilities could be utilized to meter wet gas flows. Beecroft, Zanker, and Stobie presented technical papers on the issue [28, 29, 30].

A problem with discussing wet gas flow effects on ultrasonic meters is that all ultrasonic gas meters are patented designs and therefore there is no generic meter design that can be discussed here. Each manufacturer has a unique design. However, some of the designs are similar and therefore it is possible to discuss their performance in general terms.

(*a*) *Transient Time Four Path Horizontal Ultrasonic Meters*. The four-path horizontal ultrasonic meter has the most wet gas flow research published. This design was tested extensively by the Ultraflow (1 and 2) Joint Industry Projects (JIPs) starting in the early 1990s. Much of this research has been published by Wilson [31]. Further research has been published by Zanker et al. [32, 33].

An initial industry concern was the survivability of the ultrasonic meter and, in particular, the survivability of the transducers in wet natural gas production applications. Early problems appear to have been overcome and Wilson, Zanker, and Stobie have shown that the ultrasonic meter can withstand the rigors of wet gas flows.

The performance of the four path ultrasonic meter was reported to be dictated by the wet gas flow pattern [31, 32]. Tests at two horizontal wet gas flow facilities demonstrated that the meter results are similar when the flow pattern is similar. Figure 6.1.2.4-1 shows JIP results [31]. Of the two test centers it is reported that one produced stratified flow only and the other produced both stratified and annular mist flow. Both sets of stratified flow data produced the similar results, i.e., a gas flow rate overreading with liquid content, and a relationship that can be fitted to a gradient of 5 line. The overreading is stated to be due to the liquid blockage reducing the area, causing an increased gas velocity, and the gas meter still programmed to use the full pipe area, therefore, overestimates the gas flow rate. Mist flow also produced an overreading but much lower than the stratified flow. The mist flow linear line has a gradient of unity so each additional 1% increase in the LVF means an additional 1% error on the gas flow rate prediction. The JIP data reports a maximum Lockhart–Martinelli parameter value of approximately 0.2 for LVF of 4%. Ultrasonic meter vendors therefore state that standard multipath ultrasonic flowmeters should be operated at as close to mist flow conditions as possible without affecting measurement integrity. (Mist flow conditions for set liquid-to-gas flow rate ratios are promoted by high pressure and high gas velocities.) If mist flow occurs, experimental data indicates that the total gas measurement percentage error will tend towards the homogeneous flow model. In general for multipath meters an overreading is observed that for all other parameters held constant increases with the LVF (i.e., the Lockhart–Martinelli parameter), decreases with increasing pressure (i.e., gas-to-liquid density ratio), and decreases with the superficial gas velocity (i.e., the gas Reynolds number and turbulence levels for set pressures).

Figure 6.1.2.4-2 shows the meter error versus superficial gas velocity for various LVF values from a horizontally installed 6 in. four-path ultrasonic meter at 50 bar. All liquid loadings show the error increasing and then decreasing as the superficial gas velocity increases. At the maximum gas flow rate when the flow pattern is expected to be mist flow the gas error for any given LVF is at its lowest. The original test data indicates that as the pressure increases and the flow pattern tends more to mist flow the gas flow rate error reduces.

Fig. 6.1.2.4-1 JIP Four-Path Ultrasonic Meter Wet Gas Results

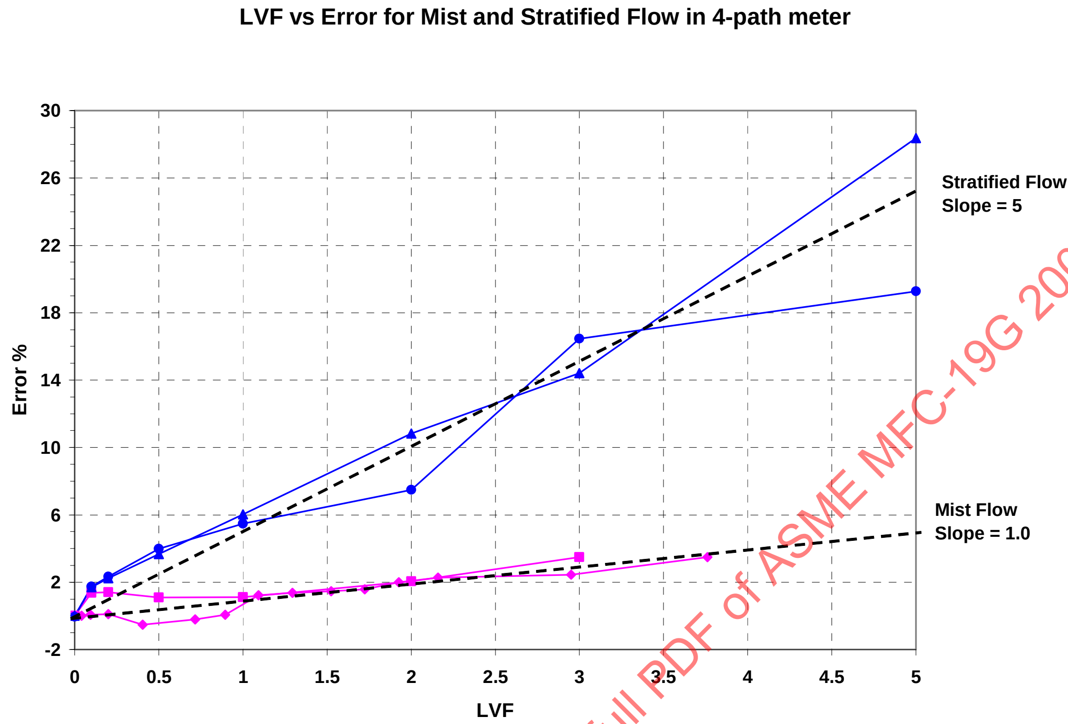


Fig. 6.1.2.4-2 Gas Flow Error of a 6 in., Four-Path Ultrasonic Meter With Wet Gas Flow at 50 bar (Superficial Velocity in m/s)

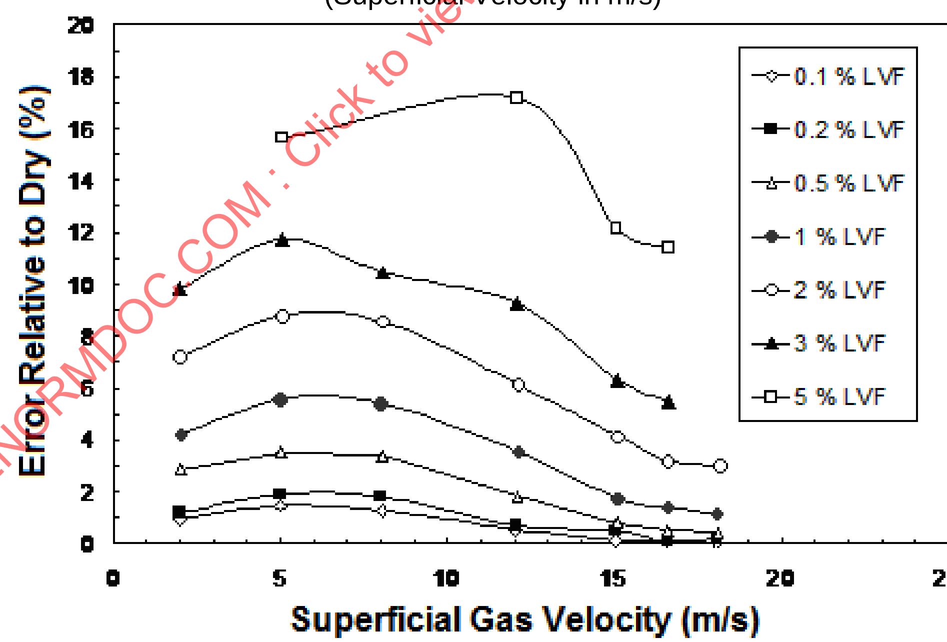


Figure 6.1.2.4-1 indicates that the flow pattern affects the ultrasonic meter. The details of what happens in the transition zones between stratified and mist flow is not well documented. If mist flow patterns were easily attainable, the suitability of the ultrasonic meter as a wet gas meter would improve dramatically. However, predicting flow patterns in wet gas is not easy and other flow patterns, which

result in much greater errors, are commonly encountered. No simple wet gas correlations are known to be published for the nonmist flow applications.

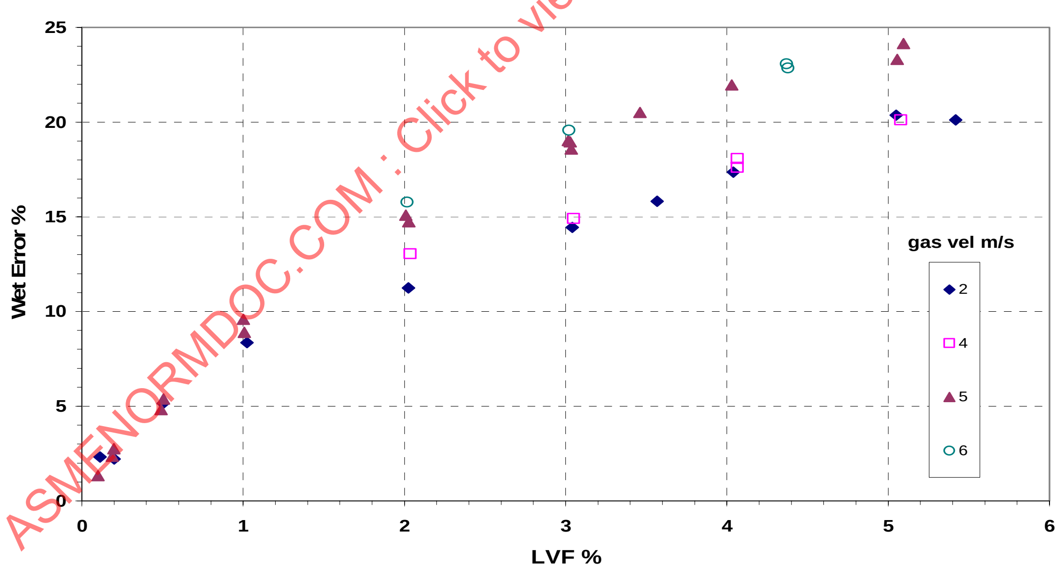
For wet gas flowmetering applications where the liquid flow rate or some form of liquid-to-gas flow rate ratio is known to be (or can be estimated to be) relatively small, the experimental data for the four path ultrasonic meter may indicate to a user what the approximate percentage error is. If the liquid loading is sufficiently small to give an acceptable gas flow rate error for a particular application then in practice users could accept an increased gas flow rate uncertainty without applying a wet gas flow correction.

(*b*) *Two-Path Horizontal Ultrasonic Meters*. The two-path ultrasonic meter wet gas flow data was released as part of research papers on prototype designs of horizontal ultrasonic wet gas meters for stratified flow patterns. (Nonmandatory Appendix L discusses these design concepts.) However, it is useful to know the performance of a two-path ultrasonic gas meter when exposed to wet gas flow without any corrections being applied. The following discussion relates to the ultrasonic design shown in Fig. L-7. Note that the wet gas flow research now discussed relates to a two-path USM where the installation was such that the chords were vertical and horizontal for a wet gas flow stratified flow pattern. The uncorrected gas meter therefore estimates the gas flow from the single horizontal path.

Figure 6.1.2.4-3 shows results presented by Zanker [33] for a 6-in. two-path ultrasonic gas meter with stratified wet gas flow. The overreading, called in Fig. 6.1.2.4-3 the "wet error," is shown in for 25 bar. The wet error increases smoothly with increasing LVF, reaching values up to 25% at the highest achievable LVF of 5%. It is therefore a very similar result to those reported for four path ultrasonic meters, which is to be expected. Overreading is tending to increase with the gas flow rate. This is the same result for gas velocities less than 6 m/s seen for the four-path meter in Fig. 6.1.2.4-1.

The limited data suggests that four and two-path meters tend to behave in a similar fashion when exposed to wet gas flows.

Fig. 6.1.2.4-3 6-in. Two-Path Ultrasonic Flowmeter Wet Gas Overreading Vs. LVF%



No technical information is known to have been released about ultrasonic meters in vertical installations metering wet gas flows.

(c) Doppler Ultrasonic Meters. No technical information is known to have been released on Doppler-type ultrasonic meters.

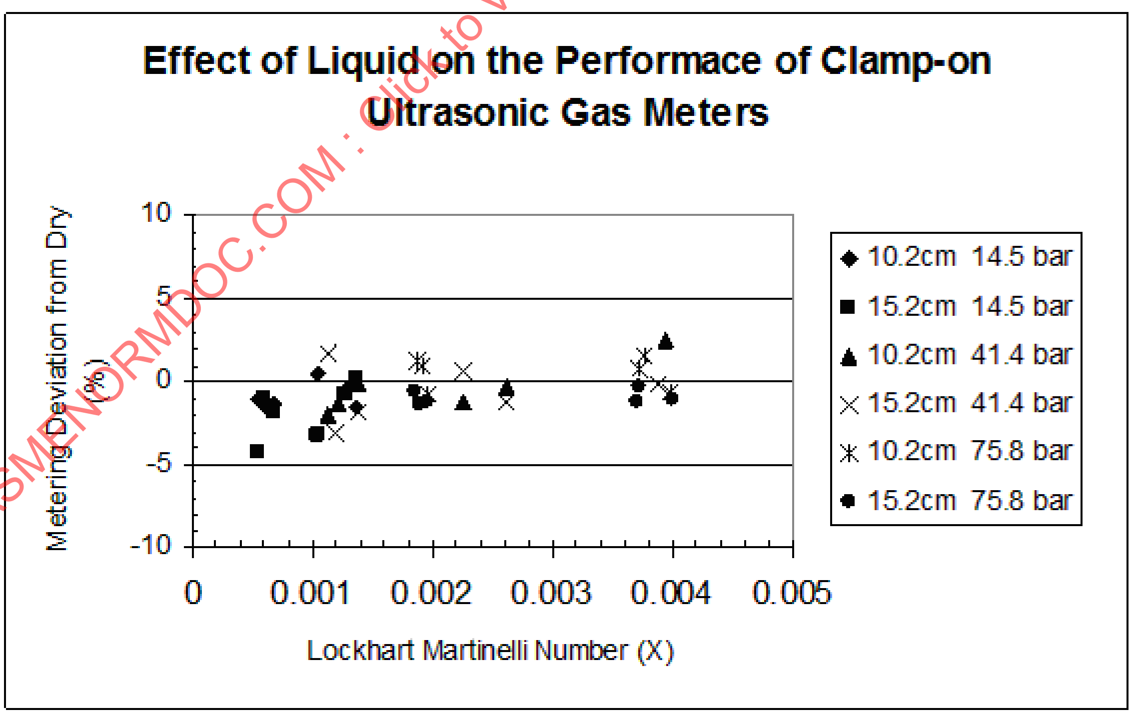
(d) Clamp-On Ultrasonic Meters. A limited amount of wet gas data has been published with clamp-on ultrasonic meters. Vedapuri et al. [34] has reported some development work on clamp-on ultrasonic equipment installed with wet gas flows, and Ting [35] has reported a test of a clamp-on ultrasonic device with wet gas flow.

Vedapuri states that clamp-on ultrasonic devices with wet gas flows have to be able to cope with the same main problem facing single gas phase flow clamp-on ultrasonic devices. That is the low impedance of the gas phase (compared to the high impedance of liquids) means the signal-to-noise ratio can be high. These meter's performances are therefore said to work better with gas and wet gas flows as the density (and therefore pressure) increases. The system was also reported to work better as pipe wall thickness reduced. Therefore, whether such a metering system could be used would depend on the application's gas density and the schedule of the pipe. Vedapuri discusses wet gas tests where the minimum GVF is 90% but the papers do not give any comparisons between reference gas meter readings and the clamp-on ultrasonic device gas flow rate prediction so the liquid induced percentage error (i.e., overreading or underreading) is not publicly known.

The only known information regarding the performance of a clamp-on ultrasonic device when used with wet gas flows comes from Ting et al. [35]. Ting reports that at CEESI 4-in. and 6-in. pipe in series had clamp-on ultrasonic equipment installed and wet gas tests where conducted simultaneously for both meters at very small liquid loadings (i.e., $X_{LM} \leq 0.004$) for three pressures (14.5 bar, 41.4 bar, and 75.8 bar). The fluids were natural gas and decane. The 4-in. meter was found to have an uncertainty of 2% with a dry gas flow. At the very low liquid loading tested it is reported that "…the effect of liquid entrainment is still within the deviation of the dry gas meter performance variation of ±2% for all tests."

The 6-in. meter was found to have a dry gas uncertainty of 2% at 75.8 bar but at lower pressures the dry gas uncertainty rose to 4%. At the low liquid loading tested it is reported of the 6-in. meter that "the effect of liquid on the performance of gas flow measurement is also within the same deviation range as dry gas flow except for a few lower pressure data points at 14.5 bar and 41.4 bar." Indications from the accompanying plot [35] shows up to 4% uncertainty in these cases (see Fig. 6.1.2.4-4).

Fig. 6.1.2.4-4 6-in. Clamp-On Ultrasonic Gas Meter Wet Gas Flow Performance



6.1.3 Summary of Single-phase Gas Meter Performance With Wet Gas Flows. At the time of writing, differential pressure (DP) meters are widely regarded as being sturdy, and for wet gas flows are the most repeatable and predictable single-phase gas meters available. There is, relative to the other gas meter technologies, a significant amount of wet gas flow research in the public domain. The response of several DP-type meters to a wet gas flow is well documented and there are correlations available for different primary elements that will correct the liquid-induced error for when liquid flow information is available. Multiple experiments show that for low values of Lockhart–Martinelli parameter the gas flow rate prediction error is relatively small and may be acceptable in some industrial applications. Greater values of Lockhart–Martinelli parameter produce larger gas flow rate prediction errors, which are often correctable when liquid flow information is available. However, there is little independent research into the validity, repeatability, and extrapolation qualities of most of the individual DP meter wet gas correlations. Furthermore, the little information that does exist shows some discrepancies for some DP meter wet gas flow test results. It is also a fact that the published data does not cover the scope of what industry needs and often therefore wet gas correlations for DP meters need to be created from a specially run wet gas test, or existing correlations need to be extrapolated with unknown consequences on the uncertainty of the gas flow rate result.

Turbine meters are relatively fragile because the turbine blades are thin and the system has moving parts. Research shows for very small liquid loadings a turbine meter can continue to operate with a small increase in gas flow rate prediction uncertainty, but even very small liquid loading flows can experience periodic slugging. This is likely to damage the turbine meter. Furthermore, long-term operation with wet gas flows will degrade the rotor parts faster than when the meter is operated in dry gas.

Vortex meters have a limited amount of wet gas flow research data associated with them. However, the little that exists indicates that for lower liquid loading (and the maximum liquid loading limit a vortex meter can operate at is dependent on pressure) the vortex meter gives a reasonably repeatable overreading that is related to the liquid loading (i.e., the Lockhart–Martinelli parameter) and pressure (i.e., gas-to-liquid density ratio). There is some suggestion the liquid induced gas flow rate prediction error could be related to the gas flow rate (i.e., the gas densiometric Froude number). However, for low LVFs (typically less than 1%) at otherwise set flow conditions vortex meters exhibit a linear error with increasing liquid fraction that can be approximated for practical industrial use to be independent of pressure and the gas flow rate. Furthermore, it is seen from the test data that like DP and ultrasonic meters the vortex meter has small overreadings with low liquid loadings and therefore they can be used to predict the gas flow rate of a wet gas flow if the increased gas flow rate uncertainty is acceptable to the user. However, at higher liquid loadings in the wet gas flow range the liquid-induced error can be considerable and the vortex meter's response to significant liquid loading is difficult to predict.

The publicly available knowledge of Coriolis meter response to wet gas flows is limited. The data available indicates that Coriolis meters are very sensitive to the presence of liquids in the flow stream, and the meter types response to wet gas is not yet fully predictable.

Ultrasonic meters are reported to give repeatable and predictable responses to wet gas as long as the flow pattern is known to be either stratified or mist flow. Under these conditions ultrasonic meters can be used to meter wet gas flows when liquid flow rate information is available for use in the liquid induced gas prediction error correction process. Furthermore, it is seen from the test data that like DP and vortex meters the ultrasonic meters have small overreadings with low liquid loadings and therefore they can be used to predict the gas flow rate of a wet gas flow if the increased gas flow rate uncertainty is acceptable to the user.

For all gas meters used in wet gas flows where a correction is required for the liquid-induced error the obvious major drawback to this method of gas metering is it is usually not a simple matter to predict the required liquid information. A generally unspoken truth about gas meter wet gas correlations is that the gas flow rate prediction uncertainty stated for the correlation in question is based on knowing the liquid flow rate information to the test equipment's liquid flow rate reference meter uncertainty. In real applications this is rarely, if ever, the case. Increased uncertainty in the liquid flow rate input to the correlation has a knock on effect on the gas flow rate prediction uncertainty. (This was explained for DP meter technology by Steven [13].) Naturally, it is beneficial to predict the liquid flow rate information with as low an uncertainty as is possible. The following section discusses methods used to obtain the liquid flow rate information so as single-phase gas meters can be used to meter wet gas flows by means of wet gas flow correlations.

6.1.4 Supplemental Systems Required for Single-Phase Meters Used With Wet Gas Flows

6.1.4.1 Separator Approach to Wet Natural Gas Production Flowmetering. In the oil and gas industry producing wells are grouped together and the flows are separated in group or bulk separators. At the outlets of these separators are meters to measure the quantity of fluids flowing. In these cases the meters are single-phase metering devices. However, separators are known to be less than 100% efficient and hence liquid carry over into the gas flow and gas carry under into the liquid flows is a source of error in this system. In the case of liquid carry over the performance of a gas meter with trace liquids can become important.

Without operator control interference many dry natural gas production flows from wells produce gas at around the desired gas flow rate for those valve settings over many weeks and months. It is also often assumed (albeit with less certainty) that wet natural gas production flows produce gas (and whatever associated liquid flow rates flow with the gas) at around the desired gas flow rate for those valve settings over many weeks and months. The changes in production flow rates are typically small when considered in terms of days and weeks. This situation often begins to change only when the well is in the later stage of production and heavier hydrocarbon gases are "dropping out" (i.e., changing to a liquid phase), water begins to be produced, gas lift is needed to maintain production quota, etc.

This approximately constant production flow rate leads to the practice of test separators being used to find the gas and liquid flow rates of a production pipeline over a period of time. Test separators are employed to test the flows from individual wells. Test separators are smaller than the bulk separators, and are used to periodically flow test the wells' production. The compromise is that operators have to assume the last spot check flow rate measurements are still valid. The value of liquid flow rate may be used as an input to a single-phase gas meter wet gas correction algorithm. The uncertainty of this liquid flow rate value is an additional uncertainty above that quoted for any wet gas meter correlation used. Separators are not always 100% efficient and liquid exiting a separator can contain entrained gas just as gas exiting a separator may contain some entrained liquid. However, the baseline for well testing has until now has been the test separator and its associated measurement systems. Separators rely on gravity in order for the gas, oil and water to naturally separate, and in some cases may require heat and chemicals (defoamers and demulsifiers).

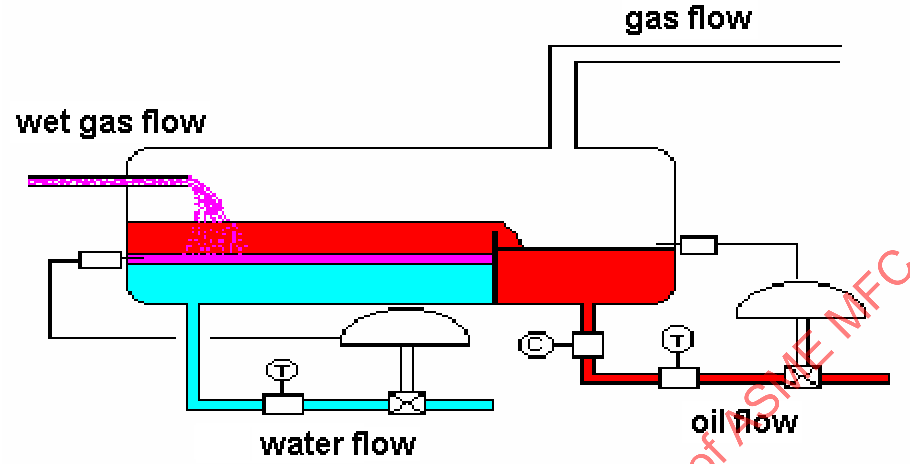
Water as the densest fluid tends to sink to the bottom of the separator; hydrocarbon liquids, which are generally less dense than water, float on the water; and the relatively light gas occupies the top of the separator vessel. In a simple separation scenario the only requirement is time for the separation to take place and the amount of time required is dependent on the separator vessel volume and the fluid flow rates. However, in reality it is more complicated than this idealized model. The hydrocarbon liquids and gas can combine in a "foam" at the hydrocarbon liquid–gas interface and the foam can be carried over in the gas flows. Hydrocarbon liquids and water can combine in an "emulsion" at the hydrocarbon liquid–water interface. In order to reduce or eliminate the foam and emulsions, chemicals (defoamer and deemulsifier) are required, which can affect the performance of the process systems further downstream. In addition to the chemicals mentioned, heat is also often required.

Test separation often requires a relatively large separator with separators in service having weights ranging from 5 tons to hundreds of tons. The large weight is due to the volume required to provide the time for gravity, chemicals, and heat to produce complete separation. However, even then separators can be less than 100% efficient.

NOTE: It is the combined associated problems of possible poor efficiency and the cost and technical problems of producing, transporting, and supporting on offshore production platforms such vessels that has driven the oil and gas industry to attempt to develop the more advanced wet gas and multiphase meters that may have benefits to all industry.

Typically, the gas, oil, and water separator is designed in one of two ways. One design is to separate the gas from the liquid and measure the gas by a conventional single-phase gas meter and the liquid volume by a conventional single-phase liquid with a water cut meter to give the ratio of oil to water. The alternative design separates the gas, oil, and water, and each are individually metered by an appropriate single-phase metering device. (It is inherent assumption here that no phase change will occur after the phases have been separated even if the thermodynamic conditions vary downstream of the separator.) A sketch of such a separator vessel of this is shown in Fig. 6.1.4.1:

Fig. 6.1.4.1 Separator Vessel That Separates Gas, Oil, and Water



Bulk separators are operated for groups of wet gas or two-phase flows from multiple wells, but test separators are used to check the performance of an individual wet gas or two-phase flow from a particular well. The test separator approach is generally perceived to be the best wet gas metering method available for wet gas and two-phase flow producing wells. However, for it to be so there are several requirements needed for the design and maintenance of the separator system. In section 11 the practical requirements of running a separator to get the best performance from it are discussed.

A major disadvantage to using test separators is that they provide only a periodic measurement of the flow rates. If the flow rates are not known to be constant over long periods, the test separator measurement may not represent the actual average flow rates over that long period. However, test separators have other important uses that can make them indispensable. Test separators allow accurate samples of the individual fluids to be obtained, which is an extremely difficult task when the wet gas or two-phase flow is not separated (as discussed in section 12).

When a separator and its associated metering systems are correctly sized, correctly used, and correctly maintained, users claim at best each of the gas, oil, and water measurements have an uncertainty of 2%. If the separator and its associated metering systems are not correctly sized, correctly used, and correctly maintained this uncertainty in measurement can rise significantly.

It should be noted there are other lesser known designs of separator. There is a patented design for saturated steam on the market called the "splitigator." This is essentially a large inclined pipe section. There are also prototype designs on rotary separators that use centrifugal forces to separate the gas and liquid phases [36].

6.1.4.2 The Throttling Calorimeter. For steam flow the application of enlarged pipe sections at an inclined angle as wet steam separators is a direct consequence of the difficulty in estimating the quality [see eq. (16)] of a single component wet gas flow. Traditional oil and gas separators are of limited value as the separated phases of a one component saturated vapor can easily change phase with small changes in the conditions. With industry desiring the measurement of wet steam flows in many applications (such as district heating, geothermal well flows, steam injection facilities at heavy crude oil wells, power stations, etc.) no universally accepted method exists for finding a wet steam flow quality. There is an ongoing debate in industry on how to achieve this.

One method is to throttle a wet steam flow (which is an isenthalpic process). The system designed to do this is called a throttling calorimeter. Figure 6.1.4.2-1 shows a schematic diagram of such a system.

The pressure and temperature of the flow are recorded thereby allowing the vapor (steam) and liquid (water) enthalpy values (h_v and h_l, respectively) to be found by use of steam tables (or steam programs). A sampling tube allows a steady sample flow of wet steam to be "throttled" by passing it through a small orifice into a lower pressure chamber.

Fig. 6.1.4.2-1 Schematic Diagram of a Throttling Calorimeter

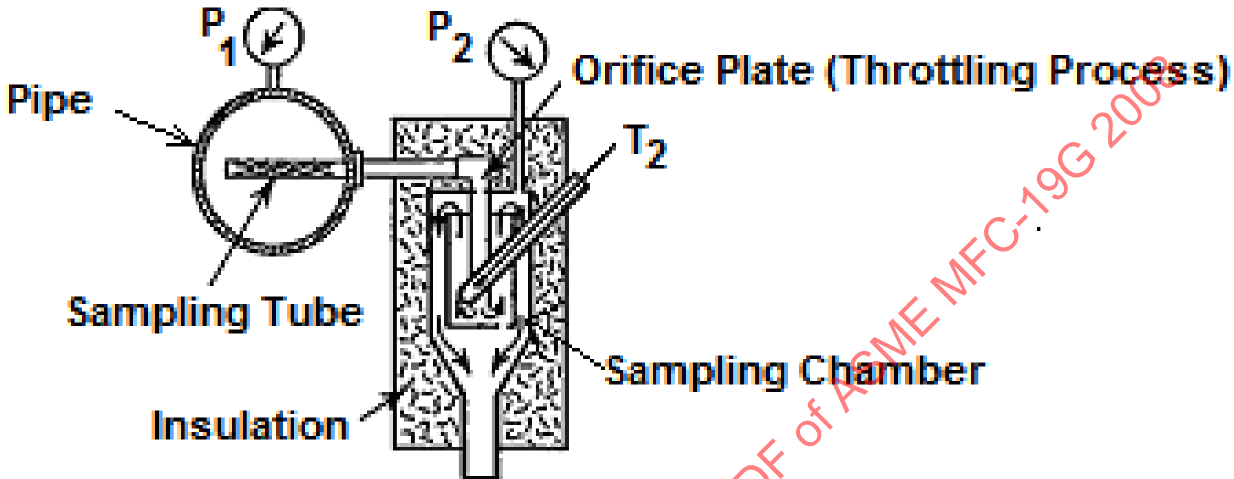
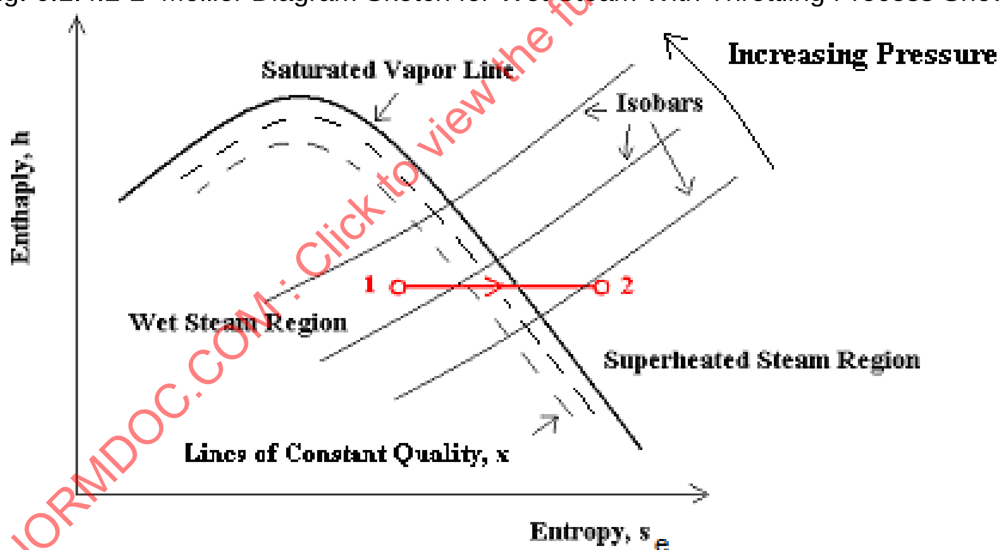


Fig. 6.1.4.2-2 Mollier Diagram Sketch for Wet Steam With Throttling Process Shown



The crucial requirements are that this throttling process is isenthalpic and that the conditions in the chamber allow the flow to be superheated. Figure 6.1.4.2-2 shows the process line sketched on a Mollier [enthalpy(h)–entropy(s_e)] diagram. With the chamber pressure and temperature measured the superheated steam sample can have the chamber enthalpy found by use of steam tables. As the expansion is isenthalpic the enthalpy (h) of the main wet steam flow has also been found as it is equal to the enthalpy of the steam in the sample chamber. As the vapor (h_v) and liquid (h_l) enthalpy values are known the quality (x) can be found from eq. (31).

$$x = \frac{h - h_l}{h_v - h_l} \tag{31}$$

As the throttling method of predicting the steam quality is wholly dependent on the flow conditions allowing an expansion that would turn wet steam into superheated steam it is unfortunate that there is a very significant limitation to this method due to a large isenthalpic pressure drop being required to turn even high quality wet steam flows to superheated steam. The higher the line pressure the lower the quality measurement limit. However, even for high pressures, the minimum quality measurable is typically greater than 0.9. Within the operating range typically an uncertainty of ±1% of read steam quality (*x*) is reported. Usually the sample is exhausted to the atmosphere. A worked example is shown in Appendix I along with a table showing typical minimum qualities for given saturated steam pressures for which the method will successfully throttle the saturated steam to superheated steam (thereby allowing the method to work). Some throttling calorimeter manufacturers produce a separating throttling calorimeter that separates out some of the liquid before the throttle, and this is reported by the vendors to reduce the minimum quality readable. Potential users need to obtain information on particular throttling calorimeter designs from suppliers.

This quality measurement can be used to calculate the liquid and vapor flow rates either directly for a closed cycle steam flow (i.e., where the total mass flow in the pipe is known) or via a wet gas metering correlation for an open steam flow (where the total mass flow in the pipe is not known). The uncertainty in the steam quality prediction is an additional uncertainty above that quoted for whatever wet gas correlation used.

6.1.4.3 Tracer Injection Methods. One method of determining the free liquid flow rate in a wet gas flow is to use a tracer dilution technique. The technique consists of injecting a carefully chosen tracer liquid at a precisely known flow rate into the wet gas flow. The tracer liquid is usually either a (low intensity) radioactive source of a known strength (such as a sodium isotope tracer) for the steam industry or an inert dye with a known fluorescent intensity for the oil and gas industry. The oil and gas industry have developed florescent dyes that will only be absorbed by one liquid component and no other component in the wet gas flow. Therefore, separate estimations of the water and oil flow rates can be made.

After tracer injection a mixing distance is required to allow full mixing of the tracer and the produced liquids. A rule of thumb is to allow 150 pipe diameters to allow for complete mixing, but it should be noted that this originates from liquid flow tracer test standards (since withdrawn) for single-phase multi-component where the liquid flow velocities are often less than in a wet gas flow. Full mixing in a wet gas flow may occur considerably before this and it is standard practice to aid the mixing process by choosing the injection and sampling points to be on opposite sides of pipe line components such as intrusive flowmeters, bends, valves, etc. In practice, it is up to the user's engineering judgment where to carry out injection and sampling.

Once full mixing is assured samples of each liquid component needs to be analysed. Note that for multiple liquid component flows the liquid component samples do not have to be in any way representative of the actual ratio of the liquid component quantities. All that is required is that each component sample is large enough to be analyzed. These technologies are now advanced enough that the sample analysis can read very small concentrations and therefore the tracer injection flow rate is very small and does not significantly affect the fluid properties of a wet gas flow. The sample analysis can be sophisticated and often includes flash and liquid property calculations that account for the sample analysis being conducted not at line conditions but at ambient conditions. The analysis is usually done at site with portable equipment. The flow rate equation for a liquid component is given by eq. (32).

$$\dot{Q} = \frac{C_o}{C_s} . \dot{q} \qquad (32)$$

where

C_o = concentration of the tracer solution injected into the stream

C_s = concentration of the sample taken downstream of injection point

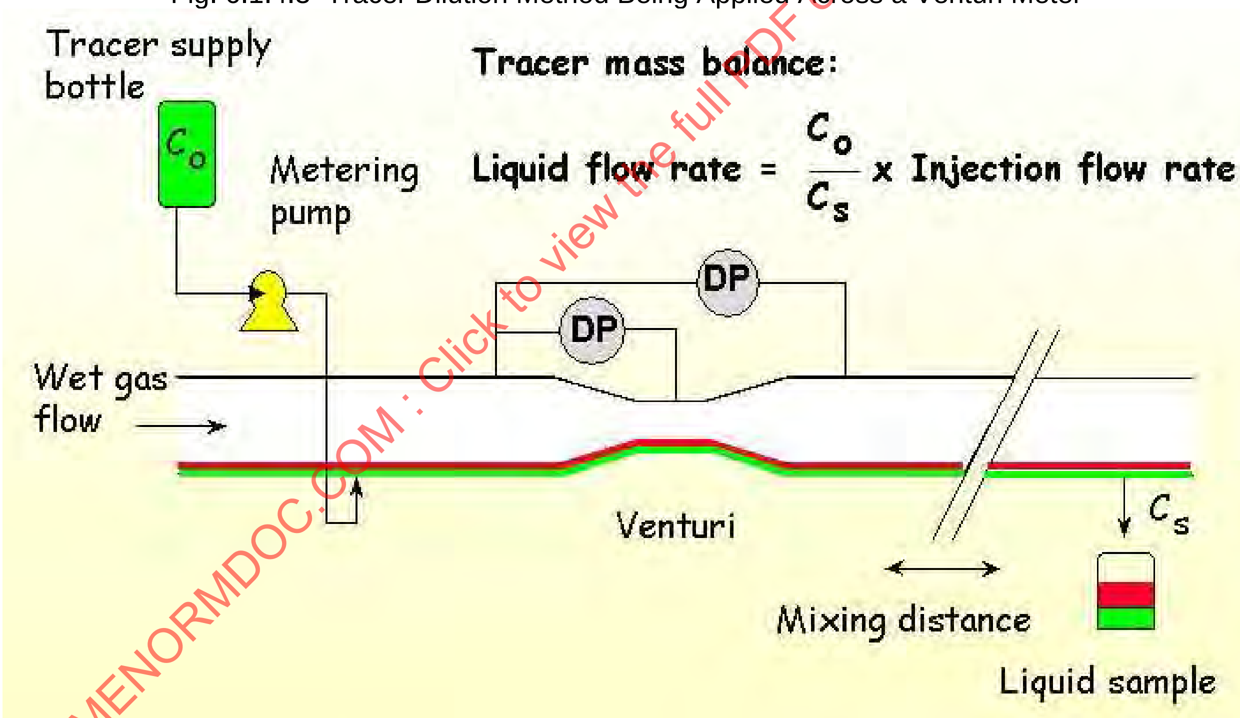
$\dot{Q}$ = liquid flow rate of that compound

$\dot{q}$ = quantity of tracer per unit time

Figure 6.1.4.3 shows a sketch of a typical application. This figure suggests stratified flow only but the technology works regardless of flow pattern, i.e., stratified or mist flow. (The mixing distance in the sketch has been compressed here to help fit the sketch into the page. Typically the sampling point would be further down stream.)

The main limitation of tracer dilution techniques is, as with test separators, the users have to assume the last spot check liquid flow rate measurement is still valid. An uncertainty in the order of 10% on each of the liquid phases is quoted by the service companies that offer these services. This 10% liquid uncertainty has a knock on effect on the wet gas meter correlation gas prediction uncertainties. This is discussed in some detail by Steven [13]. Nilsson [37] and Van Maanen [38] give a detailed overview of this technology in terms of practical application.

Fig. 6.1.4.3 Tracer Dilution Method Being Applied Across a Venturi Meter



There is no theoretical reason why gas tracers could not be developed to estimate gas flow rates. However, although this has been periodically discussed by industry, to date we know of no service offered to industry.

6.1.4.4 Capacitance Meters. Capacitance meters can offer value to a wet gas metering system. Currently they can be used as liquid indicators. That is, they will indicate to a systems operator when a gas flow has become a wet gas flow. A further use of the capacitance meter is to give a water cut estimation for a wet gas flow with an oil/water liquid component when the total liquid flow rate is estimated by a wet gas metering system (see para. 6.2). However, this procedure is greatly complicated by the capacitance meter's output being directly affected by the flow pattern. Currently we know of no stand-alone capacitance meter device that predicts the liquid flow rate.

6.1.4.5 A Concluding Statement on Single-Phase Gas Meters and Wet Gas Correlations. There has been a considerable amount of effort put into understanding how single-phase gas meters operate when the flow is a wet gas. As a result there is a general understanding of which single-phase gas meter designs are capable of being utilized with wet gas flows (i.e., which are sturdy and give repeatable and predictable errors for a known liquid quantity) and those that are not (i.e., those that are not sturdy enough and/or do not give repeatable and predictable errors for a known liquid flow rate). These single-phase gas meters that show promise with wet gas flow applications continue to be included in wet gas flowmetering research projects for two reasons.

First, there are wet gas metering applications where the liquid flow rate is known or can be estimated. In such cases knowing the wet gas response of a single-phase gas meter allows the liquid-induced error to be predicted and therefore corrected for. At the present time industry has far from complete knowledge of single-phase gas meter responses to wet gas flows. The known wet gas responses of meter designs are typically for set geometries (e.g., pipe diameter, beta ratio, meter orientation) and set flow condition ranges (fluid types, pressures, phase flow rates, etc.). Furthermore, the data that does exist is often not checked with further testing for repeatability (by either independent or the original researchers). The validity of extrapolating these published wet gas flowmeter responses is an open question. Researchers are therefore continually updating industry's knowledge of single-phase gas meters wet gas responses as industry often requires the use of these meters with wet gas flows. Such systems are typically simpler and considerably more cost effective than using a wet gas flowmeter (i.e., a metering system that will meter both the gas and liquid flow rates continuously in real time).

The second reason single-phase gas meters that show promise with wet gas flow applications continue to be included in wet gas flowmetering research projects is that the majority of the wet gas meters on the market are based around these single-phase gas meter technologies. Nonmandatory Appendix J describes the generic technologies that are used to create wet gas and "multiphase" meters, and in most cases, at least one gas meter device is present as part of a larger system. The manufacturer quoted performances of wet gas meters are, like they are for single-phase gas meters with wet gas, generally based on relatively limited test data sets (compared to the possible range of field conditions) as they have been tested on the same test facilities. Repeatability and extrapolation of performance characteristics are therefore important to wet gas meters. As single-phase flowmeters are incorporated into most wet gas meter designs tests on their wet gas flow performance are of fundamental importance to most wet gas meter designs.

The main component technologies that make up the various designs of wet gas meters are now discussed.

6.2 Wet Gas Meter Component Technologies

Although research into single-phase gas meter's response to wet gas flow is of great interest to industry, the reality is the majority of real industrial flows do not have liquid flow rate information available for use with a single-phase gas meter wet gas correction factor. This situation means that often engineers are forced to estimate the liquid flow information or periodically check it with spot checks by an appropriate method. Poor liquid flow rate estimations or shifts in the liquid flow rate between spot checks inevitably lead to increases in gas flow rate prediction uncertainty. For these reasons research into metering systems that measure both the gas and liquid flow rates in real time has been growing for the last few years.

There are wet gas meters on the market that are designed to meter the gas and the liquid flow rates in real time (but not designed to predict water cut). For these designs the liquid is generally considered to be one component, and if the liquid is multi-component, then homogeneous mixing of liquid components is generally assumed. (Nonmandatory Appendix F, example 1 shows an example of how to calculate the homogeneous liquid mix of two liquids flowing in the same pipe at different flow rates.)

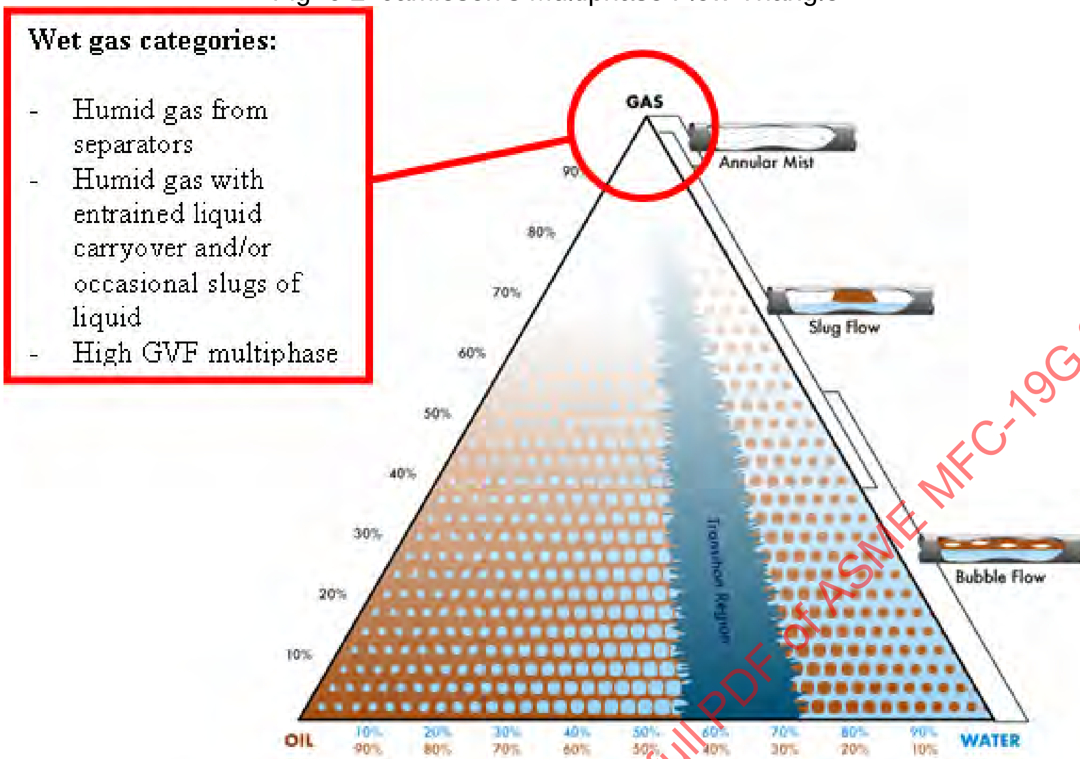
In the oil and gas industry there has been a significant push to remove the requirement for separators for two-phase flows (including general two-phase multiple liquid component flows) and therefore there has been a significant push towards the research and design of such meters. As described in the definition for “multiphase flow” (para. 2.3.1) the oil and gas production industry tends to call two-phase flows with more than one liquid component “multiphase” flows. Therefore, multiphase meters are defined as meters that for all mixes of gas and liquids measure the gas flow rate and the individual liquid component flow rates (or the gas and liquid flow rates and the water cut as it is sometimes described for hydrocarbon liquid/water multiphase flows). In reality as these meters are designed for the oil and gas industry, that means metering gas, hydrocarbon liquid, and salinated/freshwater flow rates.

Several multiphase meter designs exist but several of the earlier designs at least had reported trouble with flows at high GVF. At this point in the technological development history there was a distinction in where a wet gas meter could be applied and where a multiphase meter could be applied. Typically it was said that a wet gas meter would give poor performance at GVF’s below 90% or 95%, and a multiphase meter would give poor performance at GVF’s above 80% or 85%. This left the obvious problem of determining the metering technology to apply when a production flow was within the range 80% < GVF < 95%.

In 2001 Jamieson [39] gave a review of the multiphase metering situation and produced a figure (reproduced as Fig. 6.2 here) that is now well known in the oil and gas production industry. In Fig. 6.2 note that the text box describes the wet gas as including humid gas—i.e., gas with a relative humidity less than 100%, which is a different definition to this Report and the Norwegian Society of Oil and Gas Measurement [1]. Note that the diagram is for high-pressure high flow rates, which indicates that the wet gas flow pattern is annular mist flow. Jamieson [39] summed up the situation in 2001 with the following comments: “Most multiphase meters measure primarily in terms of gas and liquid volume flow rates, and then split the liquid fraction into oil and water. Thus this representation is a valid indication of how multiphase meters work. On this plot [Jamieson’s triangle — Figure 6.2], “wet gas” occupies the tip of the multiphase triangle, but there is no simple way of defining where the boundary with other multiphase applications lies. Some companies set it at about 5% by volume of liquids. Others extend it out to about 10% by volume of liquids. This simply reflects the history of the development of “multiphase” and “wet gas” meters. Several of the significant multiphase meter developments began as liquid/gas meters targeted at about 60% to 70% gas volume fraction, thus about two-thirds of the way along the oil/gas side of the triangle. With further development they could handle water and higher gas volume fractions, and so the performance envelope moved into the body of the triangle and up and down the oil/gas side of the triangle. It was difficult to extend the performance to GVF greater than about 85%. Wet gas metering development started at the apex of the triangle and development was targeted at handling increasing amounts of liquid. It was difficult to extend the performance to GVF less than about 95%. In some applications, there were several gas/condensate fields at 90% GVF. “Thus there was a “no man’s land” that could not be tackled by ’multiphase‘ or ’wet gas’ metering techniques.”

Therefore, at the turn of the 21st century the reality of the situation was that no technology was available for within the 80% < GVF < 95% range that the users found satisfactory. This situation, in part, led the API to define “type 3” wet gas flow. With the available multiphase meter technologies not operating well above 80% due to the “high” GVF, some in the multiphase meter circles tended to informally call GVF greater than 80% wet gas flow. API allowed for this understanding by including type 3 wet gas flow (see Nonmandatory Appendix E). However, within the last few years several multiphase meter designs have now been developed to operate into the two-phase flow region defined in this Report as wet gas flow (i.e., $X_{LM} \le 0.3$). With this development the distinction between traditional multiphase meters and wet gas meters has become blurred. Furthermore, some multiphase meter technologies use similar generic methods to predict the gas and liquid flow rates as wet gas meter technologies and then add additional technologies to predict the liquid’s water cut.

Fig. 6.2 Jamieson's Multiphase Flow Triangle



GENER AL NOTE: See reference [39].

It is immediately noticeable to any literature reviewer on wet gas and multiphase meter technologies that there are recurrent themes common to many of the available technologies. That is, several core technologies and ideas are common to different meter designs. In fact, although all marketed wet gas and multiphase meters have their own particular combination of applied technologies, and the test data sets and mathematical analysis of the raw instrument readings are usually unique for each meter type, in general, the available meters are all based on selected combinations from a list of tried and tested core technologies. These core technologies are now discussed.

The core technologies that most wet gas meters and multiphase meters with high GVF capabilities use are single-phase meters that have different wet gas flow responses in series, recovery pressure readings for DP meters, high frequency read instabilities in two-phase flows of instrument readings, extending the throat of DP meters to achieve flow pattern equilibrium and cross correlation, microwave, gamma ray, capacitance/conductance, and partial separation technologies.

The individual wet gas meters and multiphase meters with high GVF capabilities available to the market are referenced by their published literature in Nonmandatory Appendix K.

6.2.1 Multiple Single-Phase Meters in Series. One method of metering both the liquid and gas phases simultaneously is to use two or more single-phase meters in series that react differently to any given wet gas flow condition. It is then possible to solve the two unknowns (i.e., gas and liquid flow rates) by the simultaneous equations (or other mathematical techniques) offered by individual meters' wet gas correlations.

An attempt to meter liquid and gas flows in real time by using two DP meters in series was first reported by Sekoguchi. The original paper was published within Japan and not available at the time of this publication. However, Lin [40] gave a good review of Sekoguchi's work. Experiments with air/water flows through different combinations of segmental and eccentric orifice plate meters were discussed. However, the analysis of the data did not utilize the now relatively well-known DP meter wet gas parameters (see section 2) and the data was analyzed in an obscure way. The reported uncertainty of both phase predictions was 30%. However, in recent years researchers have returned to this concept backed with the increased knowledge of single meter performance with wet gas flows and considerably better performance is now achieved. The following is a description of the principles of using two single-phase gas meters is series to predict the gas and liquid flow rates in real time.

For any single-phase gas meter the presence of liquid in the gas flow will cause the meter to give an incorrect gas flow rate prediction. The scale of the error is typically related to the amount of liquid flowing with the gas (see para. 6.1). Each meter type has its own response to wet gas flow. Therefore, each meter can have a unique correlation where that meter's incorrect gas reading (for example, "$\dot{m}_{g,Apparent}$") is related to the liquid flow rate and possibly other parameters (such as the gas flow rate itself and the gas-to-liquid density ratio, etc.) by some function. That is:

$$\dot{m}_{g,Apparent} = f\left(\dot{m}_g, \dot{m}_l, \rho_g/\rho_l, ...etc\right) \quad (33)$$

The meters in series wet gas meter theory is based on dissimilar meters with different functions (i.e., correlations) being in series. For a pair of dissimilar meters in series:

Meter 1: $$\dot{m}_{g,Apparent1} = f_1\left(\dot{m}_g, \dot{m}_l, \rho_g/\rho_l, ...etc\right) \quad (34)$$

Meter 2: $$\dot{m}_{g,Apparent2} = f_2\left(\dot{m}_g, \dot{m}_l, \rho_g/\rho_l, ...etc\right) \quad (35)$$

where all parameters except the gas and liquid mass flow rates are known. Therefore, by algebraic manipulation these equations can be rearranged to give:

$$\dot{m}_l = f_1^*\left(\dot{m}_g, \dot{m}_{g,Apparent1}, P, ...etc\right) = f_2^*\left(\dot{m}_g, \dot{m}_{g,Apparent2}, P, ...etc\right) \quad (36)$$

where the superscript "*" indicates the appropriate rearranged functions.

From this expression the gas mass flow rate can be found either by separating it out of the above equations or if this is not algebraically possible (wholly dependent on the form of the original correlations) then by iteration (with one of the uncorrected meter readings being used to initiate the iteration to ensure a quick convergence). With an estimate of the gas mass flow rate obtained it can be used as an input into functions f_1^* or f_2^* to derive the liquid mass flow rate. A simple example of this is shown for two different DP meters with Murdock type equations in Nonmandatory Appendix J.

The meters in series method theoretically works for any meter combination as long as the meters used have significantly different responses to wet gas flows. In practice this can be difficult to achieve. Research into single-phase meter responses to wet gas flow have shown that different standard DP meter designs can have liquid-induced errors with too similar wet gas responses for this measurement by difference method to work well. This can result in relatively large uncertainties in the estimated flow rates (especially for the liquid flow rate). In order to get an appropriate difference in wet gas performance between DP meters, unorthodox DP meter designs have been researched and systems are now on the market. A disadvantage is the relatively high head loss multiple meters can create (and the possibility of phase change that high head loss can produce).

Daniel et al. [41] and Downing et al. [42, 43] discuss a two DP meter wet gas system. Nonmandatory Appendix K references two independent papers [44, 45] that describe investigations into the performance of this design.

Another combination of single-phase gas meters is the combination of the positive displacement meter and one or more other single-phase gas meter (usually DP meters but a vortex meter has been known to have been added as well). Nonmandatory Appendix J discusses the work of Medvejev [46] in 1972 and Chen [47] in 1982 (also summarized by Lin [40]). These works describe two-phase flowmeter system tests where the system was a differential pressure (DP) meter and a positive displacement (PD) meter in series. (This type of meter combination to create a wet gas meter has been developed commercially by the Agar Corp. Agar has also commercially developed a vortex/DP meter wet gas flowmeter).

6.2.2 Differential Pressure Meter Classical DP/Total Head Loss Wet Gas Meters. A recorded method of attempting to meter the liquid and gas flow rates in real time is to use a classical differential pressure reading and the total head loss reading from a differential pressure (DP) meter. The suggestion that the total head loss across a DP meter with wet gas flow may contain liquid flow rate information was first reported by de Leeuw [10, 48]. It was noted that for tests on a 4 in., 0.4 beta ratio Venturi meter the wet gas flow recovery was less than when that quantity of gas flowed alone. That is, the liquid presence affected not only the classic differential pressure reading but also the total head loss. De Leeuw defined the "Pressure Loss Ratio" as the ratio of the overall pressure drop (i.e., total head loss) to the traditional DP read between the upstream and throat pressure tappings. Figure 6.2.2 shows a sample of de Leeuw's published results.

The pressure loss ratio appears to be more sensitive to changes in the Lockhart–Martinelli parameters at low Lockhart–Martinelli parameters and there is a superficial gas velocity (or gas Densiometric Froude number) effect. This has been independently verified by NEL during work for the U.K. government's Flow Programme (see Nonmandatory Appendix J).

A practical application of this knowledge was immediately realized by industry. A Venturi meter with wet natural gas flow having the downstream pressure recorded could allow system operators to monitor for possible changes in the wet gas flows liquid flow rate. This then could be used an indicator for when another liquid flow spot check is required (by test separator or tracer injection method etc.) when using a single-phase meter with a wet gas correlation (see para. 6.1.4).

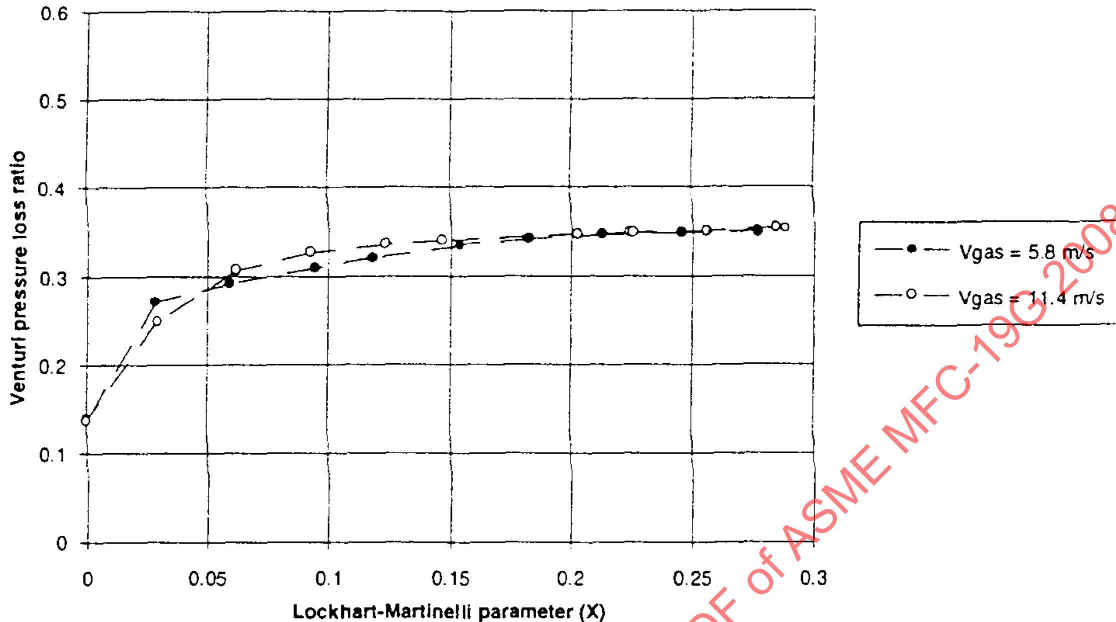
A commercial company has developed a wet gas metering concept from this research. The general concept is to fit to experimental data a function that expresses the pressure loss ratio to the parameters that are found to influence the pressure loss ratio. These have been stated to include the Lockhart–Martinelli parameter, the gas-to-liquid density ratio and the gas densiometric Froude number. Let ΔP_{hl} denote the total head loss across the DP meter and ΔP_t denote the traditional upstream to throat differential pressure created by the DP meter. Therefore the pressure loss ratio has a function (f) fitted:

$$\frac{\Delta P_{hl}}{\Delta P_t} = f\left(X_{LM}, \frac{\rho_g}{\rho_l}, Fr_g \right) \qquad (37)$$

A DP meter can have a wet gas flow overreading correlation fitted as follows:

$$\dot{m}_g = \frac{\dot{m}_{g,Apparent}}{g\left(X_{LM}, \frac{\rho_g}{\rho_l}, Fr_g \right)} \qquad (38)$$

Fig. 6.2.2 4 in., 0.4 Beta Ratio Venturi Meter Pressure Loss Ratio Vs. Lockhart–Martinelli Parameters at 45 bar



Therefore, there are two equations for two unknowns (i.e., the gas and liquid flow rate, as the Lockhart–Martinelli parameter and the gas densiometric Froude number can be reduced to these unknowns for a set meter geometry and gas and liquid densities), which can therefore be found. More details on this wet gas metering method are described in Nonmandatory Appendix J.

6.2.3 Fast Response Sensor System. A fast response sensor system is a multiphase metering technology that can be used for wet gas metering. The system uses the high-frequency readings of the natural fluctuations of pressure, differential pressure, and temperatures across an obstruction (typically but not exclusively a DP meter) to predict the flow pattern and phase flow rates of general multiphase flows including wet gas flows. Such devices are recent developments, which are still therefore seen largely as experimental devices. However, some systems are now used in practice, and these meters are continuously being developed.

All obstructions in pipe work have naturally occurring pressure, temperature, and differential pressure fluctuations even in single-phase flows that are considered on the most part to be steady. These fluctuations are products of the random turbulence that exists in all real fluid pipe flows and the random turbulence signals are considerably enhanced by wet gas flows. The frequency and magnitude of the fluctuations of pressure, temperature, and differential pressure can be related to the liquid content in a gas flow. That is, the pressure, temperature, and differential pressure (P, T, DP) and the magnitude of the pressure, temperature, and differential pressure fluctuations (for example, δ_P, δ_t, and δ_{DP}, respectively), and the frequency of these parameters' fluctuations (for example, ω_P, ω_t, and ω_{DP}, respectively) can be related through a neural network to give the flow pattern and the gas and liquid flow rates. The neural network becomes more accurate with the addition of new data sets. These systems are said to "learn by experience" and be the application of artificial intelligence to wet gas metering. The computer "learns" from the data sets the flow patterns and flow rates that gives rise to fluctuations and fit appropriate functions for the flow rates. That is:

$$\dot{m}_g = f\left(P, T, DP, \delta_p, \delta_T, \delta_{DP}, \omega_P, \omega_T, \omega_{DP}\right) \quad (39)$$

and

$$\dot{m}_l = f_1\left(P, T, DP, \delta_p, \delta_T, \delta_{DP}, \omega_P, \omega_T, \omega_{DP}\right) \quad (40)$$

where "f" and "f_1" denote functions that are continually being updated by the software as more data is added.

The drawback of such a system is the reliance on multiple data sets and reports that individual wet gas test loops can influence the parameter fluctuation characteristics (or gives a "fingerprint" that is difficult to replicate at another test facility). Also, it is not yet theoretically possible to accurately predict the effect parameter changes such as pipe diameter, fluid type, pressure, etc. will have on these fluctuation magnitude and frequencies. As the fluctuation magnitude and frequencies do not necessarily behave linearly with changing parameters (and no data on this issue is know by us to be published) it is not clear what effect extrapolation of the data sets will have on a system's gas and liquid flow rate uncertainty.

A small number of such systems are in use with the oil and gas industry for specific wet gas flows, but these have largely been developed with dedicated separators offering single-phase mass flow references. It is up to the potential user to judge if this technology is suitable for any given wet gas metering application. Such a decision would be based on whether the vendor has existing data sets that are similar to the proposed application or whether the wet gas flow will initially have a reliable phase flow rate reference method (such as a test separator in wet natural gas production facilities) long enough to allow the neural network to "learn" (i.e., gather the required amount of data across the full flow range expected in use). Toral et al. [49 – 52] discusses the development of such a system for general multiphase flow (including wet gas flow) and claims the wet gas flow uncertainty on each phase is 15%. Note that this system used an orifice plate meter and in some installations a capacitance meter. No independent research checking these uncertainties is known to us.

6.2.4 Prototype Wet Gas Meters. There are wet gas meter concepts that have been discussed in the literature that are not currently commercial products. Whereas this topic is not within the scope of this Report, Nonmandatory Appendix L discusses some of these more commonly known concepts.

6.3 High GVF Multiphase Meter Design System Components

Many of the high GVF capable multiphase meters available to the oil and gas industry use re-current themes common to many of the available technologies. That is, several core technologies and ideas are common to different meter designs. Most use a differential pressure meter as the core of the system and then phase fraction devices are incorporated into the meter body. The combined information from the DP meter and the phase fraction device/s are inputs to the calculation procedure that is usually held in confidence by the meter manufacturer. This calculation procedure is often the result of data fits from data attained from multiple tests at different test facilities and the manufacturer holds this as confidential due to the considerable financial outlay involved. It is understood by ASME to be the case that whereas general industry sees these systems as black box technologies, potential or repeat customers of the manufacturers are sometimes privy to the data analysis and calculation methods. However, little data or independent analysis of these systems is in the public domain.

The phase fraction devices used as part of multiphase meter systems are typically available as independent systems. It is of interest to engineers involved with the metering of wet gas and multiphase flows to understand the physical principles of these phase fraction devices. A simple discussion is given in Nonmandatory Appendix M, but the details of this subject are currently not within the scope of this Report.

6.4 Nodal Analysis, Integrated Modeling, and Virtual Meters

The oil and gas industry sometimes chooses to use a different methodology than flowmeter devices to predict the flow of gas, wet gas, and multiphase flows in production pipelines. This methodology is collectively termed "nodal analysis, integrated modeling, and virtual metering." These "metering methods," if indeed that is the correct description of these processes, predicts the flow rate from individual wells where it is not physically possible or economically viable to install an appropriate metering system. (Typically this means subsea or even downhole multiphase/wet gas flowmetering requirements.) These methods are more usually applied to general multiphase flows but they can and are on occasion applied to wet gas flows.

Nodal analysis, integrated modeling, and virtual metering take known (or estimated) values of variables or other information such as the pressure and temperature at individual points in the flow line, choke Cv curves, choke setting, differential across chokes, information from downstream separators, any flowmeter outputs, pipe friction coefficients, thermal loss coefficients, well trajectories, pipe internal diameter, PVT information, chemical injection points, geothermal gradients, etc. and apply these inputs or "boundary conditions" (i.e., a single variable is a "node") to mathematical multiphase flow models for the system. That is, the measured or estimated conditions at different points in the flow line are use to predict unknown parameters at the nodes and conditions at other nodes in the flow line where no information was previously known.

Nodal analysis is generally considered a manual operation and it can be time and resource consuming and more susceptible to operator error compared to virtual metering, which is generally considered to be an automated process (that can use additional information than that of nodal analysis) and operates continuously online. The virtual meter is considered to work better in network situations (i.e., with several linked pipelines) compared to looking at each pipeline in turn. By setting up the mathematical models in the network situation, common nodes are used for several pipe flows — increasing the number of effective nodal inputs into the model and generating redundancy in the process. In contrast, the manual nodal analysis method can become extremely complex and cumbersome when applied to pipe work networks.

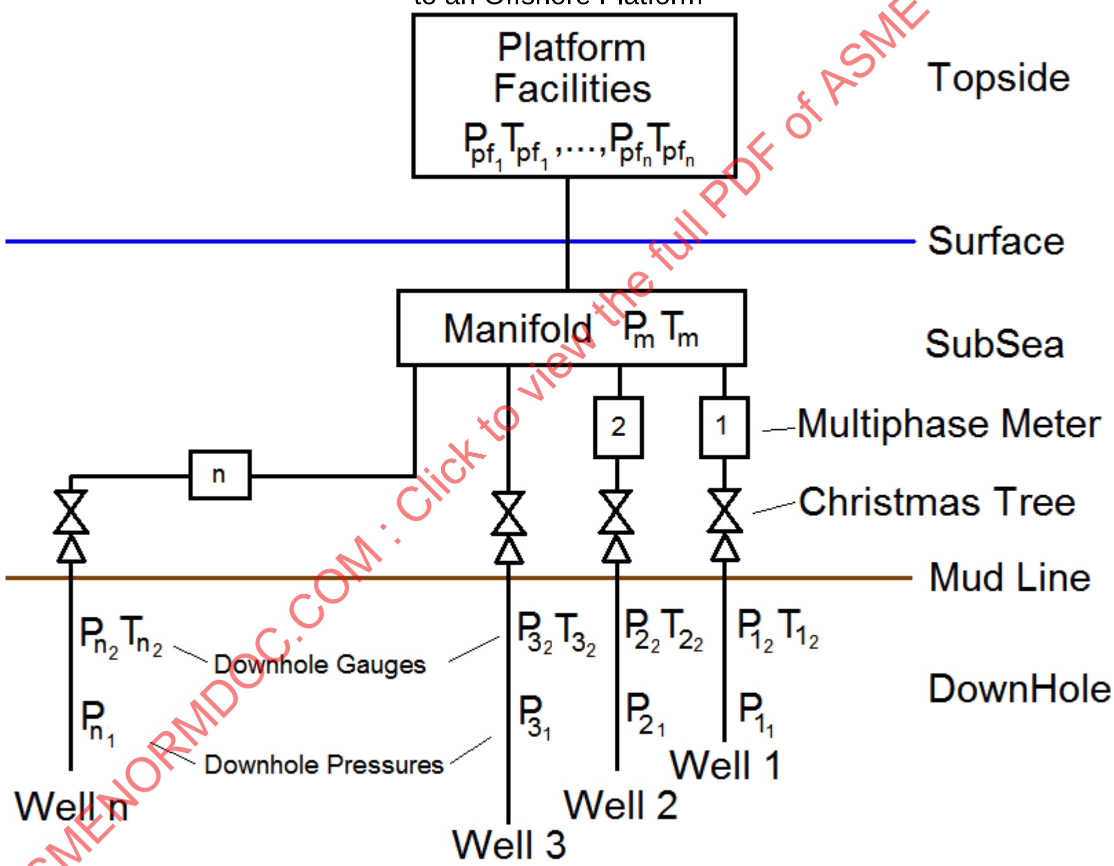
It is generally believed that the closer a node is to the well, the greater the magnitude between consecutive node parameter values, the more the technique is checked against references, the better the design and maintenance of the multiphase/wet gas meters (if any) being used, the more a multiphase/wet gas meter output is verified after significant shifts in flow conditions, the more nodes included in the input to the model then the more precise the flow condition predictions should be. PVT calculations are heavily relied upon and hence their accuracy affects the uncertainty of the flow prediction methods. The uncertainty of the analysis's output is directly related to the number of available nodes to the number of unknowns. The idea is to attempt to get instantaneous flow information from each well where the method is applied. It is generally understood that the technique should be applied to stable flows as transient flow conditions can cause significant increases in flow rate prediction uncertainty.

With multiphase flow information on each individual flow line being very useful to the oil companies and the extreme expense of fitting meters subsea and downhole, the use of this virtual type of flowmetering is very attractive to oil companies. The use of such methods is increasing especially in the most difficult deep sea applications for metering systems. The increasing power of the available computers and the improving modeling techniques incorporated into the software used, coupled with the problems and costs of installing real meters in some deepwater applications, is making these multiphase flow prediction methods more common in well surveillance applications.

A disadvantage of virtual meters is that they do have to be recalibrated or tuned if there are major changes in the pipe flows (due to changes in the reservoir behavior). The virtual meter will tell you that there is a water breakthrough occurring, but once you have had this change, it should be re-tuned to better measure for this condition. If the compositions change (PVT), they will need to receive the new PVT file (but note so would the physical wet gas and multiphase meter designs). One vendor stated to ASME that its virtual metering systems are generally retuned once every 6 months, with some installations operating for a couple of years between tunings, and others for about a month at times of large changes.

The following is a generic example of the principle. Figure 6.4 shows a simple schematic of an offshore production facility where there are "*n*" number of wells all being tied back to a manifold for one offshore platform. The pressure in each well is measured (i.e., the first set of nodes, P_{11} to P_{n1}). The pressures and temperatures at a location in each of the downhole pipes are measured (with pressure and temperature being individual nodes, i.e., P_{12} T_{12} to P_{n2} T_{n2}, etc.). Possibly the pressures and temperatures at the Christmas Tree (i.e., sea floor valve/choke arrangement at the exit from downhole) would be measured and used as nodes (although this is not shown in Fig. 6.4). Each well, or perhaps more interestingly for this methods use, only some of the wells, may have dedicated multiphase meters. Note in this example the pipeline from well three has no multiphase meter. The outputs of the available meters are nodes. The pressure and temperature in the manifold are nodes (i.e., P_m, T_m). Finally, the instrumentation (including meters) top side (i.e., on the platform) can be used as nodes (i.e., P_{pf1} T_{pf1},…, P_{pfn} T_{pfn}). The input nodes to the mathematical model allow the software to analyze and predict all the missing information that is not available as an input. In this case the system output would include a prediction of the flow rate from well 3. The system is wholly dependent on the applicability of the software/mathematical model that predicts the flow.

Fig. 6.4 Schematic of a Generic Multi-Wet Gas/Multiphase Flow Satellite Well Tie-Back to an Offshore Platform



Due to current technical and/or financial metering limitations downhole and subsea, for certain systems, there has been limited acceptance of these techniques in a case-by-case basis by some regulatory authorities (e.g., the U.S. Minerals Management Service). However, such systems have been available for the last 15 years; currently the number in use in offshore hydrocarbon production is comparable to the number of multiphase meters in use.

7 WET GAS SAMPLING

The fact that the oil and gas industry need to obtain sample(s) of the wet natural gas flow being metered stems from the requirements imposed by existing wet gas meter technologies, whether single-phase meters with wet gas correlations, wet gas meters or multiphase meter designs as described in Section 6. These requirements are typically gas–liquid ratio and/or gas and liquid phase densities.

The conditions required for the data above are

(a) in situ pressure and temperature at the meter location

(b) gas–liquid ratio at flowing condition, i.e., representative of void fractions

While the data requirement and the conditions necessary to obtain them for the proper use of the correction correlations are clear, obtaining them in practice is not a simple process.

The issues that often arise, and have become the sources of endless debate and confusion in wet gas sampling, can be traced to two topics:

(1) fidelity of the sample composition

(2) representative of flowing conditions

Sample fidelity raises the question of whether a representative sample was taken during the sampling process. The other important question, but one that is often rarely asked is, "What is the sample representative of?" Of the in situ reservoir fluid, the wellbore or near wellbore fluid, or the fluid that arrived at the meter?

Flowing condition requirements come from the phenomenon of slip (see para. 2.3.2). The presence of slip between the liquid and gas phases precludes the direct mapping of phase fractions obtained under static equilibrium conditions.

While it is worthwhile to be aware and knowledgeable of these issues, it is perhaps equally important to be aware that the questions posed cannot currently be answered practically. The other important factor, covered in the next section, is the effect of pressure and temperature on phase behavior. To illustrate the point, suppose a representative fluid sample was indeed obtained at the meter location, and the sample was then taken to a facility where the pressure and temperature changes but where the measurements for the relative gas–liquid amounts were taken and reported to be that at in situ conditions. The changes in pressure and temperature cause the relative volumes to change due to compressibility, thermal expansion, but most importantly, due to mass transfer between phases. Thus, the data obtained at the facility may be correct at the facility conditions but it is incorrect to apply this information as if it is the correct information at the required meter conditions.

Finally, a number of techniques and equipment have been developed and commercially available for wet gas sampling. This section covers some of these techniques with the intent of providing the reader with the available options. It should be noted that each methodology carries a level of uncertainty, operational ease/difficulty as well as cost. It is not, however, the intent here to provide a comprehensive comparative analysis. Good practice methods coupled with an understanding of the limitations of the technique being employed is often capable of providing quality data for the purposes of wet gas measurement.

7.1 Sampling Techniques

As discussed earlier, a key difficulty of the sampling approach is how to ensure a representative distribution of the liquid phase within the gas stream (sample fidelity). In mist flow this is not too problematic, but in stratified or annular flow, some form of mixing device is generally required to disperse the liquid throughout the gas phase. Sample probes must also be carefully designed (and positioned) to average out any remaining distribution variances within the flow. Furthermore, the sampling should ideally be isokinetic, that is to say, the linear velocity of the fluid entering the sample probe should be equal to that of the undisturbed flow stream at the sample point. If the velocity in the sample tube is higher (or lower) than the velocity in the main pipe, then the pressure in the tube will be lower (or higher) than main pipe. Lighter elements of the flow stream (i.e., gas rather than liquid droplets) may then be preferentially expelled from (or "sucked into") the sample tube, distorting the overall sample composition. To avoid phase changes, the samples themselves may have to be maintained at the same pressure and temperature as the wet gas stream during offtake, accumulation, and analysis.

Despite these challenges, new developments are gradually reaching the marketplace that claims to produce accurate samples of a wet gas stream, even though few independent evaluations of their performance currently exist.

8 PRESSURE, VOLUME, AND TEMPERATURE (PVT) PHASE PROPERTY CALCULATIONS

Most wet gas meter technologies require the phase properties as an input to the phase flow rate calculation. In the oil and gas production industry samples of reservoir fluids are ideally collected at an early stage in the reservoir's producing life and dispatched to the laboratory for a complete PVT analysis. This applies to the full range of hydrocarbon mixtures (i.e., from dry gas to black oil). In the case of a gas (particularly wet gas, condensate, and retrograde) producing formation, it is best to obtain samples of the undersaturated fluid where the average reservoir pressure is above the dew point pressure of the gas.

There are two ways to collect fluid samples for PVT analysis. Samples can be taken either by direct subsurface sampling or by surface recombination of the gas and condensate phases. Whatever technique is used, the same basic problem exists, and that is, to ensure that the proportion of the gas to the condensate in the composite sample is representative of the in situ fluid in the reservoir.

8.1 Subsurface Sampling

Subsurface sampling is a direct method of obtaining reservoir fluid sample. A special sampling container is run in the hole, on wireline, to the reservoir depth and the sample collected from the subsurface well stream at the prevailing bottom hole pressure. The problems associated with sampling an initially saturated or undersaturated reservoir, in which the bottom hole flowing pressure has fallen below the saturation pressure, can largely be overcome by proper well conditioning prior to sampling.

If the well has already been flowing, it should be produced at a low stabilized rate for several hours. This reduces the drawdown and increases the bottom hole flowing pressure, thereby allowing some, if not all of the condensate to re-vaporize in the vicinity of the well. Following this, the well is shut in for a reasonable period of time during which the fluids flowing into the wellbore under an increasing average pressure will result in the best possible homogeneous sample that is representative of the reservoir fluid.

If the reservoir is initially at or suspected to be close to the saturation pressure, the subsurface sample should be collected with the well shut in. If it is known that the reservoir is initially undersaturated, the sample can be collected with the well flowing at a very low stabilized rate so that the bottom hole flowing pressure remains above the saturation pressure during sampling. With proper well conditioning a representative subsurface sample can usually be obtained.

One of the drawbacks of this technique is the small sample volume obtained per trip. Typically several samples are taken and the saturation pressure at ambient temperature for each sample is measured and compared at the well site. Large deviations of the ambient saturation pressure indicate samples with varying composition, and thus, are not representative of the reservoir fluid.

8.2 Surface Recombination Sampling

In collecting fluid samples at the surface, separate volumes of condensate and gas are taken at separator conditions and recombined to yield a composite mixture. The well is produced at a steady rate for several hours and the separator gas-oil ratio (GOR_{SEP}) is monitored during this period. If the separator GOR is steady during the period of measurement, then one can conclude that recombining the gas and condensate in the same ratio will yield a representative composite sample of the reservoir fluid.

Note that the separator GOR is reported in volume of separator gas and condensate at separator pressure and temperature. If the GOR measurement is reported in standard volume of separator gas per stock tank volume of oil, then an adjustment must be made to determine the actual ratio in which the samples will be recombined. This is because the condensate sample is collected at separator pressure and temperature, whereas the later GOR is measured relative to stock tank conditions. The required recombination ratio, R_{sep} is given by

$$R_{sep}\left[\frac{scf}{sep\ bbl}\right] = R\left[\frac{scf}{stb}\right]\times S_h\left[\frac{stb}{sep\ bbl}\right] \quad (41)$$

where S_h is the condensate shrinkage factor from separator to stock tank conditions.

Note that the process of well conditioning as mentioned in the previous section on subsurface sampling applies to surface sampling as well.

9 WET GAS FLOWMETERING PRACTICAL PROBLEMS AND RECOMMENDED PRACTICES

With industry in the last decade increasingly choosing wet gas metering options there are various reports of wet gas metering experiences (both successes and failures) and these experiences have led to organizations having recommended practices. This information can be split into general wet gas flow information and meter specific information. These reported experiences are discussed in this section. There are also DP meter saturated steam flow specific issues to wet gas metering discussed in Nonmandatory Appendix O (which can be used as a general guide to all single component fluid wet gas flowmetering).

9.1 General Wet Gas Flowmetering Practical Issues

There are a number of general wet gas flow related industrial problems discussed in the literature. These problems are the type that would affect all meter designs as they are pipe-flow problems rather than meter-specific problems. These are now discussed.

Is the Gas Wet?

The question of whether a gas flow is wet and, if so, how wet should be asked at a systems conceptual design stage. Often problems with single-phase flowmetering occur as the system engineers have failed to allow for appropriate metering installations at the design stage. This situation is also true for wet gas flowmetering applications. If it is an industrial requirement to meter a flow that is a wet gas it is important the technical challenges associated with this are recognized at the earliest stage in the overall system design.

One problem is that it is not always easy for predictions to be made on the wetness of a gas flow prior to the system coming into service. Although it is technically difficult to predict wet gas flow conditions, it is necessary for the engineers or scientist to attempt this using the best practices available to them, as their prediction will affect the system designers' choice of metering strategy. If the wet gas has a low liquid loading, the economic, practical solution could be to install a standard dry gas meter. If it is a wet gas flow ($X_{LM} \leq 0.3$) but not general two-phase flow ($X_{LM} > 0.3$), a wet gas meter design is perhaps the prudent choice. A report of general two-phase flow (especially with a mix of liquid components as is common in the oil and gas industry) could result in a multiphase meter being chosen. The financial costs of getting this prediction wrong can be considerable. A multiphase or wet gas meter applied to what turns out to be a dry gas flow is a pointless capital expenditure. Worse perhaps, could be a dry gas meter installed when the flow turns out to be two-phase flow and no metering is capable with the system in service. Cooley et al. [53] discuss a real situation where wet gas meters were installed sub-sea for natural gas production and the flow from the well was incorrectly predicted at the conceptual stage. Higher Lockhart–Martinelli parameters were predicted than was found in practice. The actual liquid loading was found to be that of trace liquids. The wet gas meter technology was therefore found to not be required after the capital expenditure was complete. Wood et al. [54] discuss a real situation where wet gas meters were applied top-side (i.e., on a natural gas production platform) and the flow turned out to have a Lockhart–Martinelli parameter greater than 0.3. That is, it was not a wet gas but a general two-phase flow. The solution was to build a replica meter and incur significant extra expenditure in testing the replica meter under similar flow conditions. (It should be noted that another issue in this example was that the meter in question had a diameter of 14 in. and the correlations originally applied were for much smaller meters. This is a common dilemma for engineers, as little data is available to show what affect, if any, diameter has on a meter design's wet gas performance.) A further difficulty for engineers dealing with wet natural gas flows is that the conditions can change considerably over the years of production meaning that over the life time of the well no one metering technology will be suitable for the range of flows expected. Example 6 in Nonmandatory Appendix F discusses this situation in some detail.

In the steam industry it is common for any two-phase saturated steam flow to be called a wet steam flow regardless of the dryness fraction/quality (i.e., regardless of the Lockhart–Martinelli parameter). It is important for engineers working with steam to recognize the mismatch in terminology here, because if the steam is of low quality, many stated wet gas flowmetering techniques will not be appropriate. However, most practical working wet steam flows have high qualities, a fact that reduces this problem somewhat.

Stobie in his discussion of wet gas flowmetering industrial trials at a wet gas seminar in Paris in 2001 stated that "…it is generally true to say that all the rules that apply to dry gas with respect to developed flow profiles and swirl are equally applicable to wet gas." There is nothing in the literature stating that the single-phase meter standards are not relevant for wet gas flow, but there are many incidences of extra requirements when wet gas flowmetering is to be carried out. These are discussed below.

(a) Hydrates. Many physiochemical phenomena influence the performance of wet gas and multiphase meters in the oil and gas industry. Certain chemical components in the flow stream can react or encounter phase transitions within the operational envelope of the flow line and measurement system. These components include water, hydrocarbons, diluents, and salts, which lead to the formation of hydrates, waxes, and scale. Their presence at particular operating conditions alters the chemistry, fluid properties, flow geometries, and fluid dynamics of the system.

Hydrates are solid, ice-like inclusion compounds. However, hydrates differ from ice in their crystal structures, phase boundaries, and fluid properties. When hydrates form in a flow line they can create a number of problems, the foremost being flow line blockages that reduce production and create safety issues. In addition to blockages, hydrates cause many measurement errors. The causes of the errors include changes to the flow geometry, fluid properties, fluid amounts, sensor performance, and flow conditioner among others.

Fig. 9.1-1 Hydrate Blockage in a Section of Pipe

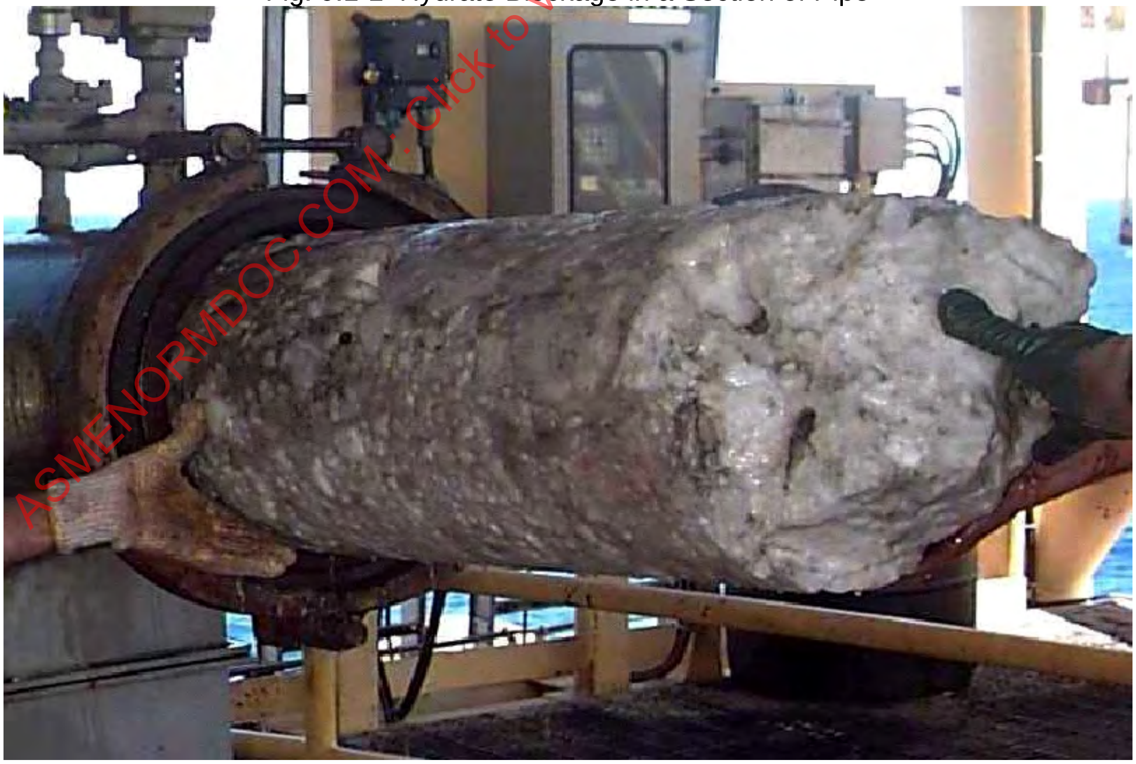


Figure 9.1-1 shows a section of pipe extracted from the field in order to determine the cause of the loss of gas production. It shows a hydrate mass plugging the gas flow line. The image demonstrates the severity of the problems that could potentially occur when hydrates are present.

Hydrates form when the light hydrocarbon components contact liquid water at a temperature and pressure condition that cause the formation of a new crystalline phase. At the phase transition condition, water and dissolved gas molecules rearrange themselves in the liquid phase and initiate the nucleation of nascent hydrate crystals. The rearrangement produces a cage-like framework of water molecules that surround small gas molecules. Once hydrate formation begins, the hydrate phase acts as a mass transfer sink for small molecules from the gas and liquid phase. The decomposition of hydrate crystals is, to an extent, the reverse of the formation process. It occurs when stabilized hydrate cages decompose and release trapped gas components.

Hydrates occur in two common crystalline forms, Type I and Type II. The stability of a given hydrate crystal depends on the temperature, pressure, and gas composition. Much is known about the basic chemistry and equilibrium properties of hydrates. Pressure–temperature graphs are used to define the approximate operating conditions at which hydrate crystals form. Figure 9.1-2 provides an example of the P-T hydrate phase boundary for the formation of pure methane hydrate. To the left of the curve methane hydrate will form, and to the right of the curve methane hydrate will not form or will decompose if present in the flow.

Natural gas, gas condensate, and oil mixtures form hydrates at higher temperature and lower pressure conditions than pure methane. They form both Type I and Type II hydrate crystals. For operating conditions at or below the ice point, e.g., surface flow lines in cold climates, both hydrates and ice will coexist in the flow line. The hydrate phase diagram provides an approximate guide that identifies the operating conditions necessary to support hydrate control and remediation methods.

Hydrates form on the surface of free liquid water, on condensed water droplets on the pipe wall, and at any hydrocarbon liquid-water-gas interface. They are readily transported downstream by the gas and liquid phase under steady and unsteady state flow conditions. Fluid flow causes hydrate crystals to deposit and agglomerate into larger hydrate masses that become the precursors to the development of flow line blockages. The presence of condensate in the flow appears to accelerate the formation and distribution of hydrates. Methanol is the most common chemical treatment applied for hydrate control. It acts by inhibiting water's ability to form stable hydrate cages and causes a shift in the hydrate phase diagram to lower temperatures. Hydrate control chemicals also include ethylene glycol and proprietary anti-agglomerate or kinetic based inhibitors.

The hydrate phase diagram does not provide a complete picture of the physical behavior of hydrates in a flow line. It describes neither the complex transport behavior of hydrates nor important rate-dependent issues including hydrate formation/dissolution rates, hydrate blockage formation properties, and other considerations.

Fig. 9.1-2 Pressure–Temperature Phase Boundary Conditions for Methane Hydrate

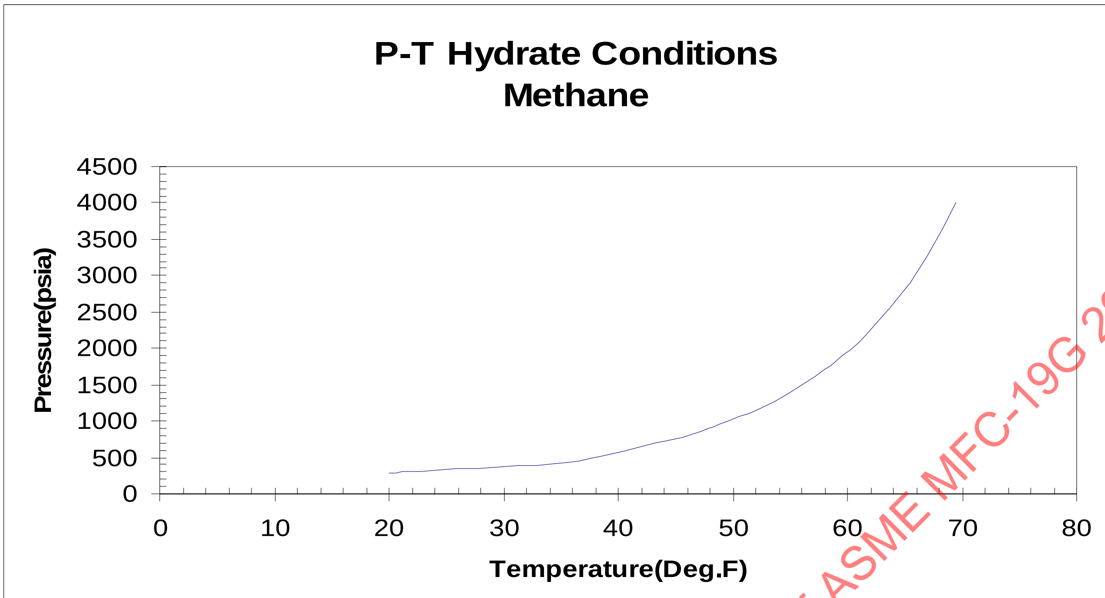


Fig. 9.1-3 Cross-Sectional View of Hydrates in a Flow Stream

Figure 9.1-3 is a cross-sectional view through a high-pressure viewing window into a test line that contains three phases; a natural gas phase, a liquid water phase, and a crystalline hydrate phase. A light source illuminates the bottom of the image through the liquid water phase. The middle portion of the image shows a mass of hydrate crystals that are flowing on the surface of the liquid water phase. Hydrates have agglomerated on the liquid flow stream in the middle of the image against the view glass. The top portion of the image shows hydrate crystals that have washed onto the upper third of the view

port window. Figure 9.1-3 illustrates one of the mechanisms that may occur in the early development of a hydrate blockage at low flow-rate conditions.

Safety concerns are paramount when dealing with hydrates. Removing or decomposing hydrates in the flow line is a time-consuming process and is particularly difficult when access to the flow line is restricted, e.g., hydrate constrictions and blockages in well bores and sub-sea flow lines. It is common for multiple hydrate blockages to occur. The presence of hydrate constrictions and blockages can cause large differential pressures to build up in a flow line. This can cause hydrate masses to act as projectiles, which are capable of damaging flow control and measurement equipment. Poor control of hydrate decomposition may cause the rapid release of a significant amount of high-pressure gas, which increases the risk of a catastrophic failure.

Research is underway to understand the effect of hydrate behavior in flow lines and their subsequent affect on wet gas and multiphase measurement. Much remains to be understood before comprehensive industry guidelines can be prepared on the effect of hydrates on wet gas and multiphase measurement. The industry needs to develop reliable full-scale data on meter performance when hydrates are present in their flow lines and meters. Currently, hydrates are a concern to producers, operators, owners and regulating authorities. Loss of production time, loss of flow lines, and inaccurate measurements means that significant capital is at risk.

(b) General Contamination, Scaling and Salts. Hydrates and ice are not the only solid particle problem faced by engineers metering wet gas flows. Many industrial wet gas flows are not clean and general debris and contaminates are carried with the fluids (e.g., geothermal steam, coal bed methane, or wet natural gas well flows). This adversely affects all pipe components, and flowmeters can be particularly susceptible to this. Wear from erosion can be a problem as can be the buildup of contaminants on a meter's wetted surfaces. Figure 9.1-4 shows the substantial buildup of contaminates on an orifice plate after three months of service in a sour wet coal bed methane flow. Clearly such buildup has the potential to adversely affect many meter designs (e.g., by changing meter factors, damaging turbine bearings, blocking pressure ports, depositing on ultrasonic transducer faces or vortex bluff bodies, etc.)

Fig. 9.1-4 Orifice Plate Removed From a Coal Bed Methane Wet Gas Flow After Three Months' Service
(Reproduced with the permission of McCrometer, 3255 West Stetson Ave, Hemet, CA 92545)

Fig. 9.1-5 Sample of Scale Taken From a Wet Gas Meter

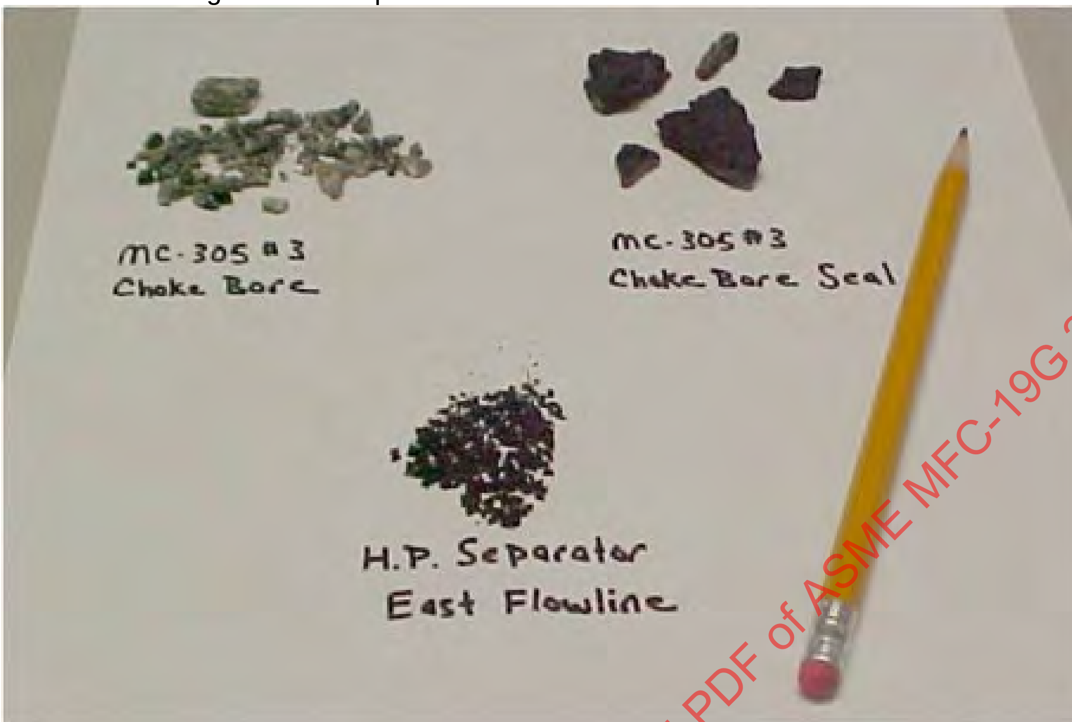


Fig. 9.1-6 Wet Gas Flow Scale Buildup Around a DP-Based Wet Gas Meter

Fig. 9.1-7 Wet Gas Flowmeter After Scale Removed

A similar problem is that of scaling. Scale is usually a mineral compound consisting mainly of calcium or magnesium carbonates or calcium sulfate. The most common problem reported is calcium carbonate deposits. In many cases it is mixed with other substances such as iron or sand grains. The release of carbon dioxide from water can trigger a chemical reaction, which can lead to the deposition of scaling compounds.

Here then, as with hydrates and ice, the core problem is the presence of water. If the meter experiences a reduction in diameter due to a buildup of scale, this change of geometry will cause a corresponding error in its response. Therefore, the result of the scale can be to block sensors and cause a meter to stop operating, or to reduce the flow area and increase the flow velocity, thereby causing error in the measurement. Samples of the scale removed from a subsea low liquid loading wet gas flow are shown in Fig. 9.1-5. This was discovered while investigating the cause of a continuing increase in the flow line imbalance. The scale production in this line was quickly stopped by scale inhibitor injection, but the scale that had already formed could not be removed without a multi-million dollar intervention, and even then there were no guarantees that the flowmeter's original geometry could be restored. Figures 9.1-6 and 9.1-7 show a wet gas meter removed from service after the pipe line had severe scale buildup. Figure 9.1-6 is as the meter was after removal from service, and Fig. 9.1-7 is after the meter had been subsequently cleaned. A pressure port can clearly be seen to have been blocked by scale while in service. The effect of scale is a universal problem in industry, and failure of any wet gas meter device due to scale buildup is not the fault of any meter manufacturer. If a meter fails due to scale buildup, in all probability the competitor's meter would also fail under the same adverse conditions.

Salts building up in pipelines causes similar problems to that of scale buildup. Figure 9.1-8 shows salt buildup in a natural gas production pipeline.

(c) Flow Conditioners. Although it is common practice when space is limited with single-phase flows to apply a flow conditioner to reduce the upstream lengths required to comply with various gas meter

standards, there are reported problems with applying flow conditioners to wet gas flows. Liquids may build up in front of the devices producing a skewed flow profile that will result in measurement error.

Stobie [55] reports a wet natural gas flowmeter giving gross errors when a flow conditioner was applied far greater than when it was not installed. The reason was found to be that the conditioner was partially blocked. In this case hydrates had gathered at the flow conditioner and/or formed due to the pressure drop across the conditioner plate and blocked several of the flow conditioner orifices meaning that the flow was more asymmetrical leaving the conditioner than it was entering it. Stobie claimed the

Fig. 9.1-8 Salts Built Up in Natural Gas Production Line

problem could be alleviated by a combination of different solutions. These are

(1) operate above the hydrate point on the pressure–temperature curve (which admittedly is not always possible)

(2) suppress the hydrate formation with injection of suitable chemicals (e.g., methanol or glycol) and heat trace the instrumentation components that contain cavities for hydrates to settle in stagnant fluid (such as DP meter impulse lines). Heat tracing is not yet possible at sub-sea metering locations.

It is clear that such actions are considerably easier if the problem is addressed at the system's initial design stage, as retrofitting is nearly always more expensive and troublesome.

Stobie [55] states, "It is not advocated that flow conditioners be installed in a wet gas environment…unless it is absolutely sure that hydrates cannot be formed or hydrate inhibitors are to be used as a matter of course."

Unfortunately, hydrates forming "ice plugs" during the meter service with wet gas flow is not the only cause of meter failure due to ice plugs in and around the meter. Meters are often hydro-tested with high-pressure water before going into service. It should be thoroughly checked on pressure let down that the meter is completely dried. Meters such as DP meters have components such as impulse lines where capillarity effects can keep water in the line. If this is not removed before installation and the service flow conditions are suitable for ice or hydrate formation, the impulse line could be blocked by an ice or hydrate plug and the meter will fail to operate.

Also note that this flow conditioner problem is not restricted to the hydrate and natural gas production. Scale, salts, and ice would block flow conditioners in natural gas pipelines, and in other wet gas flows such as geothermal steam, the flow often has particulates that can lodge in flow conditioners and this will clearly have the same effect on the meter downstream of the conditioner plate as hydrate blockage in natural gas metering situations.

There could in certain situations be scope for the use of flow conditioners that are specially designed for use in wet-gas applications. Their purpose would be to attempt to redistribute liquids evenly across the flow rather than alter a single-phase flow profile. In this situation the flow conditioner could be called a mixing device. If the process is such that hydrates will not be formed or any other particulates will not be present, this could be considered.

(d) Meter Location, Flooding, and Slug Strikes. As a general rule, wet gas meters should be installed in high points in a pipeline. Even wet gas flows that appear to be steady deposit liquid at low points in the pipeline network and over time the liquid builds up until it presents the wet gas flow with a significant blockage. It is therefore generally not good practice to position a wet gas meter at a low pipe work position, as it may get periodically partially or fully flooded, which would adversely affect the readings. High points in the pipe work system are best for wet gas meters.

Periodically then, with a buildup of liquid at low points in the pipeline network this blockage will be shunted forward by the buildup of gas blocked by the liquid. These "slugs" as they are termed are a hazard to all wet gas and two-phase meters. Slug catchers are often installed in wet gas and multiphase flow pipe networks, but they cannot, of course, always guarantee that a meter will not suffer periodic slug strikes. Engineers need to consider the sturdiness of a meter design if it is to be used for wet gas flow applications. This is one of the reasons orifice plate meters have been less popular for wet gas flow applications in the last two decades. Some meter operators in the natural gas production industry have reported buckled orifice plates after exposure to slugging flow. Figure 9.1-9 shows such a buckled orifice plate.

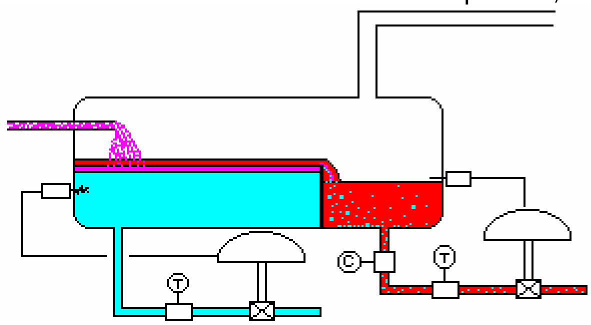
Fig. 9.1-9 Orifice Plate Buckled by a Slug Strike While in Wet Gas Service

It is also a fact that some wet gas meters, regardless of where they are installed, can on occasion be flooded. It is the responsibility of the system design engineers to choose a wet gas meter that is considered to be capable of surviving a slug strike and starting up operation after flooding. If the meter is required to be installed in a remote location it should be capable of start up with no maintenance personnel intervention.

(e) Separator Inefficiency. It is a known fact that in operation many two-phase and multiphase test separators are not correctly sized and suitably maintained or operated to give the lowest measurement uncertainties achievable. Many of the best practice requirements are regularly compromised. Some of the more common problems with the meter installations are outlined below.

Separators can be incorrectly sized for the application (e.g., the two-phase flow has changed over time and no longer matches the design criteria for the separator). The flow rates can be too high

Fig. 9.1-10 Example of Poor Level Control in Three-Phase Separator, Leading to Water in Oil Leg



for the separator's volume to separate the gas and liquid with 100% efficiency. In such conditions, the gas outlet receives wet gas flow and the liquid outlet's bubbly liquid flow. For multiphase separators, poor level control on the liquid separation section can cause the oil phase to contain water. (This is shown in Fig. 9.1-10).

In addition, there are a number of other cautions to the use of test separators. For many years, the test separator has been the only method for conducting well tests in the oil and gas industry. As such, it has been the benchmark for well rate determination, and in many areas of the world, test separator design and its use have been legislated. However, this commonality of use may have tended to hide areas where test separators may not in fact perform as desired, and where measurements made by the single-phase meters on the separator outlets are not as good as declared on the respective meter "nameplate."

(f) Separator Gas Outlet Flowmetering Problems With Orifice Plate Meters. Orifice plate meters are the most popular meter used to meter outlet gas flow from a separator. There are several practical problems with orifice plate meters in gas outlet separator service. The plate geometry can be worn. That is, the sharp edge required by the orifice plate meter can be worn away over time due to lack of inspection, causing noncompliance with the single-phase orifice plate meter standards. The orifice diameter (i.e., the "beta ratio") can be different from the optimum size for the flow in question. There can be installation effects such as short metering runs downstream of pipe components such as 90-deg bends (i.e., less than required for orifice plate meter standards), infrequent calibration of secondary instrumentation (pressure, differential pressure, and temperature), and high errors when running at the low end of differential pressure transmitters' range. Maintenance crews have been known to reinstall the orifice plate back to front after inspections. Furthermore, often only spot samples are taken to find the gas composition in order to determine density, and the presence of any liquid carryover could bias the result.

In general terms, it would be expected that the best measurement uncertainty that could be expected on the gas leg of a test separator would be in the region of ±2% to 3%. However, it is more likely to be the case that an average uncertainty might be in the region of ±3% to 6% with contributions from the some of the above sources. In extreme cases where most or all of the above sources contribute to uncertainties, it has been suggested by some industry sources that it is possible to encounter uncertainties up to ±15%.

This all assumes the gas flow out a separator is dry (i.e., that the separator is 100% efficient). When it is not the flow will be a wet gas flow. With inefficient separators any liquid carryover is often intermittent and so correction for liquid can be difficult to perform accurately. Separators should therefore be designed and operated optimally to avoid significant liquid carryover.

NOTE: When the orifice plate meter does encounter wet gas flows, it is common to install orifice plates in these situations with drain holes located at the base of the orifice plate. Any additional uncertainty incurred as a result of the use of drain holes is usually not accounted for.

Although the orifice plate meter is the most common meter used at the outlet of a separator, many of these problems are not specific to the orifice plate meter and would affect other DP meters and other gas meter performances as well. Therefore, separation is not necessarily the ideal method to meter the gas phase of a wet gas flow.

(g) Separator Liquid Outlet Flowmetering Problems With Turbine Meters. Liquid turbine meters are popular meters for metering the liquid outlet flow from a separator. There are several practical problems with liquid turbine meters in liquid outlet separator service. Over a period of time the physical state of the turbine can deteriorate. The blades can be damaged, the bearings can be worn, and pipe surface can change with erosion, corrosion, and contaminate deposits. The calibration of secondary instrumentation (i.e., pressure and temperature transmitters) can be infrequent. Water cut measurement instrumentation performance (in two-phase separator liquid leg) can lead to uncertainties. Often only spot samples are taken of the fluid in order to determine liquid density and viscosity. Inefficient separators can have gas carryunder, and poor separation performance (emulsions, foaming, etc.) can lead to contamination of phases and changing fluid viscosities outside the meter's calibrated range. Although the turbine meter is a common meter used at the liquid outlet of a separator, many of these problems are not specific to the

turbine meter and would affect other meters as well. Therefore, separation is not necessarily the ideal method to meter the liquid phase of a wet gas flow.

On the liquid side it might be expected that the best measurement uncertainty achievable is 2%. In extreme cases liquid measurement uncertainty could be above ±5%, perhaps as high as ±10%. In the case of a two-phase separator, the performance of the water cut monitor could be a significant source of error, possibly leading to errors of 20% or more on calculated individual oil and water flow rates.

Engineers in the oil and gas industry claim that a correctly sized separator with well maintained meters running in good flow conditions may be metered at best to about ±2% for all phases. However, once outside the "perfect envelope" then performance may fall off dramatically.

(h) Wet Gas Flow Issues With Differential Pressure Meters. DP meters are one of the most popular technologies for metering wet gas flows, whether they are stand-alone devices or a component in a wet gas metering system. It must be noted that DP meters have a flow rate output that has a square root relationship with differential pressure. For example, if the liquid-induced flow rate overreading of a DP meter is 1.4 times the dry gas flow rate, then the square root of the ratio of the actual differential pressure read with the wet gas flow to the differential pressure if the gas phase flowed alone would also be approximately 1.4. (The approximation is due to the second order effect of the expansibility factor being dependent on the differential pressure.) When sizing a wet gas DP meter, the DP transmitter must be chosen such as it will be capable of reading the actual differential pressure produced by the wet gas and not just the theoretical DP if the gas phase flowed alone. The following example highlights the issue. For an arbitrary chosen overreading of 1.4 (i.e., 40%) the actual wet gas differential pressure read would be nearly double (i.e., 1.96) than if the gas phase flowed alone. An explanation of this statement is as follows. The standard generic DP meter mass flow rate equation is:

$$\dot{m}_g = EA_tYC_d\sqrt{2\rho_g\Delta P_g} \quad (42)$$

When a stand-alone DP meter is used with a wet gas flow, the liquid induces an error in the differential pressure read (i.e., ΔP_{tp}). This is almost always a positive error, and the use of this read differential pressure in the generic DP meter mass flow rate equation leads to a gas flow rate positive error or "overreading" of the DP meter. That is:

$$EA_tY_{tp}C_{d\,tp}\sqrt{2\rho_g\Delta P_{tp}} > \left(\dot{m}_g = EA_tYC_d\sqrt{2\rho_g\Delta P_g}\right) \quad (43)$$

Therefore we have:

$$Over-\mathrm{Re}ading = \frac{EA_tY_{tp}C_{d_{tp}}}{EA_tYC_d}\sqrt{\frac{2\rho_g\Delta P_{tp}}{2\rho_g\Delta P_g}} = \frac{Y_{tp}C_{d_{tp}}}{YC_d}\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} \approx \sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} \quad (44)$$

Note that the for single-phase gas flow, the discharge coefficient for DP meters (C_d) is sometimes a function of the Reynolds number, and the expansibility factor (Y) is a function of the read differential pressure (ΔP_g). As single-phase flow DP meters need to know the flow rate to find the Reynolds number the flow rate prediction [i.e., eq. (42)] is iterative. In the case of wet gas flow the incorrect differential pressure is used (ΔP_{tp}) and therefore the iteration of eq. (42) gives a convergence on an incorrect gas mass flow rate ($\dot{m}_{g,Apparent}$), discharge coefficient ($C_{d\,tp}$), and expansibility (Y_{tp}) set. However, in most cases [as in eq. (44)] it is a good approximation to say:

$$\frac{Y_{tp}C_{d_{tp}}}{YC_d} \approx 1 \quad (45)$$

In our example, therefore, we see we have eq. (46), which can be rewritten as eq. (47):

$$Over - Reading \approx \sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = 1.4 \quad (46)$$

$$\Delta P_{tp} \approx (1.4)^2 \Delta P_g = 1.96 \ \Delta P_g \quad (47)$$

Therefore, we see that a DP meter wet gas overreading of 1.4 actually corresponds to a differential pressure wet gas overreading of approximately 1.96.

Failure to take into account the significantly higher DPs that can be caused by liquid presence in the gas flow rate can cause saturation of the DP transmitter and failure of the metering system.

When a DP meter is used in a heavy wet gas (i.e., close to the Lockhart–Martinelli parameter of 0.3) or into the general two-phase flow regime (i.e., $X_{LM} > 0.3$) the possibility of severe slugging increases. Severe slugging can be a problem not only due to potential damage to the primary element of a DP meter but also due to potential damage to the DP transmitters. Passage of slugs can cause sudden large pressure spikes, and this can cause the DP transmitters to drift or even fail. Severe slugging is a flow phenomenon best avoided if possible.

When using a DP meter with a horizontal wet gas flow the common single phase method of taking pressure readings from the averaging of four evenly spaced pressure ports radially around the perimeter of the pipe cross section should not be used. Such a set up would cause flooding of the lower pressure ports and possible errors in the DP read. It is advisable to read the pressure in a horizontal wet gas flow situation from pressure ports located at top dead center of a horizontally mounted DP meter. This gives the best chance for any liquid in the impulse line to overcome any capillary effects and drain due to gravity. If top dead center is not possible for any installation the closest position to this around the top half of the pipe is the appropriate pressure port location.

For situations where the fluid temperature is not as damaging to the DP transmitters, the impulse lines connecting the pressure ports to the differential or static pressure transmitters should be as short as possible in this horizontal mounting. It is also important to have straight impulse lines, or if that is not possible, to have at least no low points in the impulse tubing such as U-bends where liquid could collect and affect the pressure reading. To minimize cooling and local phase changes within impulse lines, it is often beneficial to place the pressure transmitters in a (heated) enclosure.

When the temperature of the fluid is too high for direct contact with the DP transmitter's wetted surfaces (as is typically the case with steam flows) it is usually necessary to insulate the transmitter from the flow by use of liquid insulation. Here the impulse lines are required to vertically drop, so the impulse lines can be filled with a liquid column of a known height. This reduces the temperature of the flow at the DP transmitter. It is important to know the precise height of these liquid columns, because if the two columns are different heights, a correction factor on the DP reading is required. Without this correction gross errors can occur in the DP measurement and therefore the flow rate prediction. A common way for the steam industry to ensure even liquid columns is to use a condensate pot. Nonmandatory Appendix O discusses these issues in more detail.

(i) Differences Among Theory, Laboratory, and Real World Wet Gas Flows. It should be noted that there can be differences between idealized and simplified theories on wet gas flowmeter performance, laboratory wet gas meter performance, and performance of meters used with real world wet gas flows. Theories on wet gas meter performance usually make assumptions and approximations that are not always appropriate in some real world situations. Laboratory tests are by their nature strictly controlled and therefore no unexpected extra variable is (knowingly at least) added. In real wet gas flows, unexpected or untested phenomena can exist to make the situation more complex.

An example of this is thermodynamic effects and phase change throughout a wet gas metering system. Most theoretical discussions of wet gas meter performance assume that there are no significant thermodynamic effects and hence the gas-to-liquid mass flow rate ratio stays approximately constant and the process is effectively isothermal. Most meters tested with wet gas flows in laboratories have stable gas and liquid components, and significant changes in phase through the meter due to pressure and temperature fluctuations do not then generally occur. In real industrial flows this is not always the case. For example, wet steam flows will change dryness fraction/quality if an intrusive meter causes a change in pressure and/or temperature. In the case of a wet natural gas production flow where natural gas, light hydrocarbon liquids (condensates), and water flow together, fluctuations in the thermodynamic conditions throughout the metering system could cause phase change and hence increase the complexity of metering the flow considerably.

Evidence of the importance of thermodynamic effects in a real-world situation is given in Fig. 9.1-11. Here a Venturi meter is metering a wet gas production flow. The picture is taken in a desert shortly after dawn during the winter. The pipe is painted brown. The white substance covering large sections of the pipe is frost that has formed during the colder temperatures at night. The frost has melted (or perhaps never formed in the first place) on sections of the metering system. This suggests that these frost clear pipe sections have a higher temperature than the frost covered sections, which provides evidence that the flow is not isothermal and phase change may be significant. The conclusion from this finding is that wet gas meter users should be aware of the limitations of both the theory and the wet gas test laboratories when faced with these difficult real-world conditions. Gas meter users should therefore anticipate higher uncertainties in the wet gas meter's output in real-world flows than at test laboratories.

(j) Miscellaneous Comments. Wet gas flowmetering systems are relatively new on the market, with most systems in service sized and built to order. The installation should be designed with the requirements of the wet gas flowmeter in mind. The flowmeter could be installed with suitable isolation valves so that the meter can be removed and inspected as required. Because industry has relatively little experience with these systems, there is often a need for reverification of metering systems operated at lower uncertainties. This may necessitate the scheduling of maintenance visits with the express aim of reverifying the wet gas measurement system. As with other routine maintenance programs, scope may exist for the relaxation of calibration frequencies, subject to satisfactory operation of the instrumentation.

Fig. 9.1-11 PDO Wet Gas Venturi Meter With Frost and Frost Clear Sections Showing Thermodynamic Effects as Significant

10 UNCERTAINTY OF A WET GAS METERING SYSTEM

Measurement uncertainty is the estimate of the interval width within which the true value lies. It represents the lack of complete knowledge regarding the measurement process. The lack of knowledge is assumed to be normally distributed and therefore characterized by a mean and standard deviation. The mean is the best estimate of the true value while the standard deviation represents the uncertainty. A confidence level is associated with the final statement of uncertainty. Most measurement uncertainties are stated at 95% confidence, which corresponds to two standard deviations. The relationship between different confidence levels is discussed further at the end of this Section.

Uncertainty analyses for many measurement processes have been well developed for quite some time [56, 57, and 58]. These measurements include mass, length, temperature, and voltage. The uncertainty associated with most typical applications can be estimated based on well-documented methods.

The measurement of flow rate is less straightforward than the measurements discussed above. The calibration process and facilities are generally more complex. Often the meter will exhibit variations in performance and uncertainty over the operating range. A variety of installation effects might affect how a meter performance can change. Finally, the meter might exhibit sensitivity to variations in the thermodynamic properties of the fluid being measured.

Estimating the uncertainty of a wet gas flow measurement is more difficult than the single-phase flow case. At a fundamental level the measurement of wet gas requires the measurement of two fluid flow rates with significant degrees of correlation between the fluid properties of the two phases. Even with relatively straightforward measurement processes, estimating the effect of correlation on the uncertainty is problematic.

Currently there is no generally accepted method to estimate the uncertainty of a wet gas measurement. The material in Nonmandatory Appendix N represents a start towards such a method by presenting several uncertainty scenarios.

In this Report, we recommend that uncertainty levels be provided at 95% confidence intervals. The relationship between percentage confidence interval and uncertainty level is shown in Table 10. If the specified uncertainty for a measurement is stated as 0.5%, 1.0%, 1.5%, and 2% at 68% confidence interval, this means that in a large number of measurements the value of the measured item will fall within this range of accuracy 68% of the time. However, if one requires higher confidence in the measured values (i.e., higher confidence intervals) the stated uncertainty will change as shown in Table 10. As was earlier stated, wet gas literature and equipment specifications can describe uncertainties at different confidence intervals. This guideline recommends that all uncertainties be reported at 95% confidence intervals. Table 10 is supplied to allow conversions so a fairer comparison between products could be made.

EXAMPLE: The specification of a wet gas meter is quoted as “±4% uncertainty in gas flow rate measurement at 90% confidence interval.” The same meter specification will become ±5% uncertainty if the required confidence is 95%. The user should be aware that the value of uncertainty of a meter will be higher) as the level of confidence interval increases.

Table 10 Conversion Factor for Uncertainty at Different Confidence Levels

Stated Confidence Interval, %	Coverage Factor for Normal Distribution	Uncertainty at Confidence Interval (%) = Coverage Factor*Standard Deviation						
68	1	±0.5	±1	±1.5	±2	±2.5	±5	±10
90	1.6	±0.8	±1.6	±2.4	±3.2	±4	±8	±16
95	2	±1	±2	±3	±4	±5	±10	±20
99	2.6	±1.3	±2.6	±3.9	±5.2	±6.5	±13	±26

The technical issues regarding wet gas flowmetering uncertainty are discussed in Nonmandatory Appendix N.

Finally, one last uncertainty related subject should be commented on here. It should be understood that there is as yet no guideline available to industry that addresses the best way to present wet gas/two-phase/multiphase flowmeter test results. There are many different methods in the available literature. It is important to understand that any one set of data that indicates a meter’s performance can be plotted in many different ways. Different data plot methods can make the same data set look quite different. Some methods can give a meter the superficial appearance of having either a better or worse performance than if the data were plotted using other methods. Scheers [114], who gives an excellent overview of similar issues for multiphase flowmeters, emphasizes some points that are also relevant for the case of wet gas flowmeters.

A good way to show wet gas meter test data is to produce several different graph types of the same data set. This allows the data to be seen from different perspectives allowing a more rounded view of the result to be obtained. However, in some of the literature, lack of space or the aim to show a product in only the best light means it is common for data to be presented in one convenient graphical form only. The reader of wet gas meter technical literature is therefore encouraged to take an independent critical view of any plotted data.

NONMANDATORY APPENDIX A[1]
DETAILS INVOLVING THE DEFINITION OF TERMS

A-1 THE ORIGINS OF THE LOCKHART–MARTINELLI PARAMETER

The Lockhart–Martinelli parameter (X_{LM}) used in this Report to define wet gas flow as any two-phase or multiphase flow with a Lockhart–Martinelli parameter value equal to or less than 0.3 has a protracted origin. R. W. Lockhart and R. C. Martinelli were engineers investigating two-phase flow in pipes. In 1949 they published a paper [71] discussing the pressure losses involved with two-phase flow in unit lengths of pipes. The pressure loss prediction was described by a parameter denoted as "*X*" and defined by Lockhart and Martinelli as the square root of the ratio of the liquid flow rate friction pressure drop across a unit length of pipe if the liquid phase flowed alone to the gas flow rate friction pressure drop across a unit length of pipe if the gas phase flowed alone [see eq. (A-1)].

$$X = \sqrt{\frac{\Delta P_{l\,f}}{\Delta P_{g_f}}} \quad \text{(A-1)}$$

Where subscript *f* means friction pressure drop and $\Delta P_{g\,f}$ and $\Delta P_{l\,f}$ are the unit length friction pressure drop of the gas and liquid phases if they flowed alone in the pipe, respectively. That is, the original term "Lockhart–Martinelli parameter" was not developed as a wet gas metering tool but rather a pressure loss predictor for unit lengths of pipe with two-phase flow. The pressure drop along a unit length of pipe of constant pipe area is traditionally called the head loss. The head loss for a unit length of pipe in single-phase flow is calculated by the product of the fluid density and a factor called the major losses (denoted here by the symbol "h_{loss}"). For horizontal flow this major loss is defined by eq. (A-2):

$$\frac{P_1 - P_2}{\rho} = \frac{\Delta P_f}{\rho} = h_{loss} \quad \text{(A-2)}$$

where ΔP_f is the friction pressure drop for either phase's single-phase flow along a unit length of pipe. For turbulent single-phase pipe flows, h_{loss} is calculated by the eq. (A-3):

$$h_{loss} = f_r \frac{L}{D} \frac{\bar{U}^2}{2} \quad \text{(A-3)}$$

Where *L* and *D* are the unit length of pipe and pipe diameter respectively, $\bar{U}$ is the average flow velocity and f_r the "friction factor" that is traditionally found with the "Moody diagram." For the case of wet gas flow it is reasonable to say the liquid velocity if the liquid flowed alone is significantly smaller than the gas velocity if the gas flowed alone, i.e., $\bar{U}_l << \bar{U}_g$. It is also true that the liquid viscosity is an order of magnitude greater than the gas viscosity. That is, $\mu_l >> \mu_g$. Therefore, from the general Reynolds eq. (A-4),

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

$$\mathrm{Re} = \frac{Inertia\ \ Forces}{Viscous\ \ Forces} = \frac{\rho \bar{U} D}{\mu} = \frac{4\dot{m}}{\pi\mu D} \qquad \text{(A-4)}$$

these facts combine to make the Reynolds number of the liquid if the liquid flowed alone to be significantly smaller than the Reynolds number of the gas if it flowed alone, i.e.,:

$$\bar{\mathrm{Re}}_l << \bar{\mathrm{Re}}_g \qquad \text{(A-5)}$$

The friction factor is, for a given pipe roughness, solely a function of the flow's Reynolds number. In this case we are always considering the same pipe with two different flows (i.e., that of the liquid and the gas phases of the wet gas flow flowing alone in that pipe) and therefore the relative pipe roughness is constant. Therefore, from the Moody diagram for any given pipe roughness it can be seen that the difference in Reynolds numbers between the gas and liquid flows means different friction factor values, for example, f_l and f_g for the liquid and gas friction factors, respectively. In fact, for wet gas flow the condition $f_l >> f_g$ exists. Now, from eqs. (A-2) and (A-3) we see that the single-phase major pressure loss in a pipe is

$$\Delta P_f = \rho h_{loss} = \rho f \frac{L}{D}\frac{\bar{U}^2}{2} \qquad \text{(A-6)}$$

The Lockhart–Martinelli parameter was defined by R. W. Lockhart and R. C. Martinelli as square root of the ratio of the liquid flow rate friction pressure drop across a unit length of pipe if the liquid phase flowed alone to the gas flow rate friction pressure drop across a unit length of pipe if the gas phase flowed alone. Therefore, X is actually calculated in the following way:

$$X = \sqrt{\frac{\Delta P_{l_f}}{\Delta P_{g_f}}} = \frac{\sqrt{\rho_l f_l \frac{L}{D}\frac{\bar{U}_l^2}{2}}}{\sqrt{\rho_g f_g \frac{L}{D}\frac{\bar{U}_g^2}{2}}} = \sqrt{\frac{f_l}{f_g}}\frac{\bar{U}_l}{\bar{U}_g}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(A-7)}$$

From conservation of mass:

$$\bar{U} = \frac{\dot{m}}{\rho A} \qquad \text{(A-8)}$$

Therefore, we have eq. (A-9):

$$X = \sqrt{\frac{\Delta P_{l\,f}}{\Delta P_{g_f}}} = \sqrt{\frac{f_l}{f_g}}\frac{\bar{U}_l}{\bar{U}_g}\sqrt{\frac{\rho_l}{\rho_g}} = \sqrt{\frac{f_l}{f_g}}\frac{\dot{m}_l}{\dot{m}_g}\frac{\rho_g}{\rho_l}\sqrt{\frac{\rho_l}{\rho_g}} = \sqrt{\frac{f_l}{f_g}}\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} \qquad \text{(A-9)}$$

i.e.,

$$X = \sqrt{\frac{\Delta P_l}{\Delta P_g}} = \sqrt{\frac{f_l}{f_g}}\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} \qquad \text{(A-10)}$$

Equation (A-10) is the mathematical expression for the original meaning of the Lockhart–Martinelli parameter as published by the engineers R. W. Lockhart and R. C. Martinelli in 1949.

NOTE: In the original paper [71] Lockhart and Martinelli did not use the Moody diagram as discussed above but rather took the Blasius equation form for a friction factor for turbulent flow in smooth pipes (Re < 10^5):

$$f = \frac{0.316}{\mathrm{Re}^{0.25}} \tag{A-11}$$

and stated:

$$f_l = \frac{C_l}{\mathrm{Re}_{lp}^{n}} \tag{A-12}$$

$$f_g = \frac{C_g}{\mathrm{Re}_{gp}^{m}} \tag{A-13}$$

where C_l, C_g, m, and n were unknowns and the superscript *gp* means gas Reynolds number if the gas flowed alone in the pipe. Superscript *lp* means "liquid Reynolds number" if the liquid flowed alone in the pipe. Hence the actual Lockhart–Martinelli parameter was given in [71] as:

$$X^2 = \frac{\mathrm{Re}_{gp}^{m}}{\mathrm{Re}_{lp}^{n}} \frac{C_l}{C_g} \left(\frac{W_l}{W_g} \right) \frac{\rho_g}{\rho_l} \tag{A-14}$$

Where W indicates weight flow so the ratio is the same as the mass ratio. Hence, from eqs. (A-12) and (A-13), eq. (A-14) is equivalent to eq. (A-10).

$$X = \sqrt{\frac{\mathrm{Re}_{gp}^{m}}{\mathrm{Re}_{lp}^{n}} \frac{C_l}{C_g} \left(\frac{W_l}{W_g} \right) \frac{\rho_g}{\rho_l}} = \sqrt{\frac{f_l}{f_g}} \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} \tag{A-15}$$

A widely circulated two-phase flow metering paper was published by Murdock [5] in 1962. In this paper Murdock used a parameter, which has since been denoted by uppercase "*X*" by engineers applying Murdock's work (although Murdock himself never used this symbol). This parameter was used during the discussion for the behavior of orifice plate meters in general two-phase flow. (In more recent times researchers used this same uppercase "*X*" to define other parameters, so for clarity in this document "X_{Murdock}" will denote this "Murdock" parameter.) Murdock described the relative amount of liquid and gas in a given pipe wet gas flow by using the square root of the ratio of the differential pressure that would be read by a given orifice plate meter if the liquid flowed alone to the differential pressure that would be read by the same given orifice plate meter if the gas flowed alone [see eq. (A-16)].

$$X_{Murdock} = \sqrt{\frac{\Delta P_l}{\Delta P_g}} \tag{A-16}$$

Note that ΔP_g and ΔP_l are the differential pressure (DP) meter-induced pressure drops if the gas and liquid phases flowed alone, and they are each the sum of the individual phases' momentum and friction pressure drops between the meter pressure tappings for a horizontal flow. Physically, then, this is the square root of the ratio of the sum of the momentum and friction pressure drop read by the orifice plate meter for when the liquid flows alone and the sum of the momentum and friction pressure drop read by the orifice plate meter for when the gas flows alone. As the standard DP meter equations for gas and liquid are eqs. (A-17) and (A-18), respectively, then eq. (A-16) can also be written as eq. (A-19):

for gas: $$\dot{m}_g = EA_t Y C_d \sqrt{2\rho_g \Delta P_g} \tag{A-17}$$

for liquid: $$\dot{m}_l = EA_t C_d \sqrt{2\rho_l \Delta P_l} \tag{A-18}$$

$$X_{Murdock} = \sqrt{\frac{\Delta P_l}{\Delta P_g}} = \frac{C_{d_{gas}} Y}{C_{d_{liquid}}} \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} \tag{A-19}$$

Note here that $C_{d_{gas}}$ and $C_{d_{liquid}}$ are the particular orifice plate meter discharge coefficients for the Reynolds numbers of the gas and liquid phases flowing alone, respectively, and E is the velocity of approach that is a DP meter geometric constant defined by eq. (A-20):

$$E = \frac{1}{\sqrt{1-\beta^4}} \tag{A-20}$$

Therefore, if (and only if) the phases flowing alone through the orifice plate meter would produce the result

$$\frac{C_{d_{gas}} Y}{C_{d_{liquid}}} \approx 1 \tag{A-21}$$

would this Murdock parameter ($X_{Murdock}$) be equivalent to the definition of the Lockhart–Martinelli parameter [see eq. (4)] contained in this Report.

Chisholm [6,7,8] later published a general two phase flow correlation for orifice plate meters and used a similar (but not identical) parameter denoted again by uppercase “*X.*” This is shown as eq. (A-22), where here we use the subscript “Chisholm.”

$$X_{Chisholm} = \frac{1-x}{x} \sqrt{\frac{\rho_g}{\rho_l}} \tag{A-22}$$

where x is the flow “quality” as described by eq. (16).

$$x = \frac{\dot{m}_g}{\dot{m}_g + \dot{m}_l} \tag{16}$$

Note, then, that Chisholm’s eq. (A-22) can be rewritten as eq. (A-23):

$$X_{Chisholm} = \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} \tag{A-23}$$

Chisholm derived eq. (A-22) similarly to the way Murdock derived eq. (A-16). In fact, eq. (A-22) is essentially the same as eq. (A-16) with the added assumptions that gas expansibility is negligible (i.e., $Y = 1$) and the ratio of the gas and liquid discharge coefficients (for when these phases flow alone) is approximately unity. This in practical terms gives the same result as if it were assumed the superficial single phase flows of gas and liquid had no losses [i.e., the superficial discharge coefficients ($C_{d_{gas}}$ and $C_{d_{liquid}}$) are both unity]. *X* in Chisholm's eq. (A-22) / (A-23) is then simply the square root of the ratio of the liquid inertia if the liquid flowed alone to the gas inertia if the gas flowed alone. Unlike Murdock's eq. (A-19), Chisholm's eq. (A-23) is independent of a DP meter's characteristics, as no discharge coefficients or expansibility terms are required. In fact, this parameter is independent of any meter properties and can be used to describe the liquid to gas content of any wet gas flow regardless of whether a meter is present or not.

This parameter is therefore a very useful nondimensional way of describing the relative amount of liquid in a gas flow. Murdock never called his parameter the Lockhart–Martinelli parameter. Chisholm [7], however, did. This is, in fact, an error by Chisholm as it should be noted that when considering eqs. (A-10), (A-19), and (A-23) they are not the same. That is:

$$\sqrt{\frac{f_l}{f_g}} \neq \frac{C_{d_{gas}} Y}{C_{d_{liquid}}} \neq 1 \qquad \text{(A-24)}$$

For a wet gas flow condition the square root of the ratio of the friction factors of the gas and liquid flows if these phases flowed alone in a straight unit length of pipe in place of a DP meter is not equal to unity. The ratio of a DP meter's product of gas discharge coefficient and expansibility of the gas flow flowing alone to the liquid discharge coefficient of the liquid flow flowing alone is not equal to unity. Also, these two parameters are not equal to each other (except for by coincidence). Therefore, the Lockhart–Martinelli, Murdock, and Chisholm definitions of the parameter *X* are all different.

During the 1990s, two-phase flow research papers began erroneously to call *X* in eq. (A-23) the Lockhart–Martinelli parameter. Some researchers attempted to register this error (e.g., Steven [11] called the parameter derived as eq. (A-23) the "modified Lockhart–Martinelli parameter" to indicate the difference from the original parameter) but their efforts were futile. The term "Lockhart–Martinelli parameter" is now entrenched in the natural gas production industry as meaning eq. (A-23). Therefore, Chisholm's simplification of Murdock's eq. (A-19) — shown as eq. A-23 — has become almost universally known now as the Lockhart–Martinelli parameter, and this therefore is how the "Lockhart–Martinelli parameter" is now defined. That is:

$$X_{LM} = \sqrt{\frac{\text{Superficial Liquid Inertia Force}}{\text{Superficial Gas Inertia Force}}} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(A-25)}$$

A-2 THE GAS DENSIOMETRIC FROUDE NUMBER

The gas densiometric Froude number (Fr_g) is a wet gas flow modification of the standard Froude number (Fr). The standard Froude number is defined as the square root of inertia forces to gravity forces. The gas densiometric Froude number is defined as the square root of the gas inertial force if the gas phase flowed alone to the liquid gravity force ratio. The standard Froude number is calculated by eq. (5). The gas densiometric Froude number is calculated by eq. (6).

$$Fr = \sqrt{\frac{Inertia\ Force}{Gravity\ Force}} \qquad (5)$$

$$Fr_g = \sqrt{\frac{Superficial\ Gas\ Inertia\ Force}{Liquid\ Gravity\ Force}} = \frac{\bar{U}_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} \qquad (6)$$

Where U_{sg} is the superficial gas flow's average velocity, which is calculated by eq. (3).

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} \qquad (3)$$

Equation (6) is now derived for the case of wet gas mist, although the same result can be obtained with other flow patterns.

The gas inertia force if gas flows alone:

$$F_{Gas\ Inertia} = \rho_g U_{sg}^2 D^2 \qquad \text{(A-26)}$$

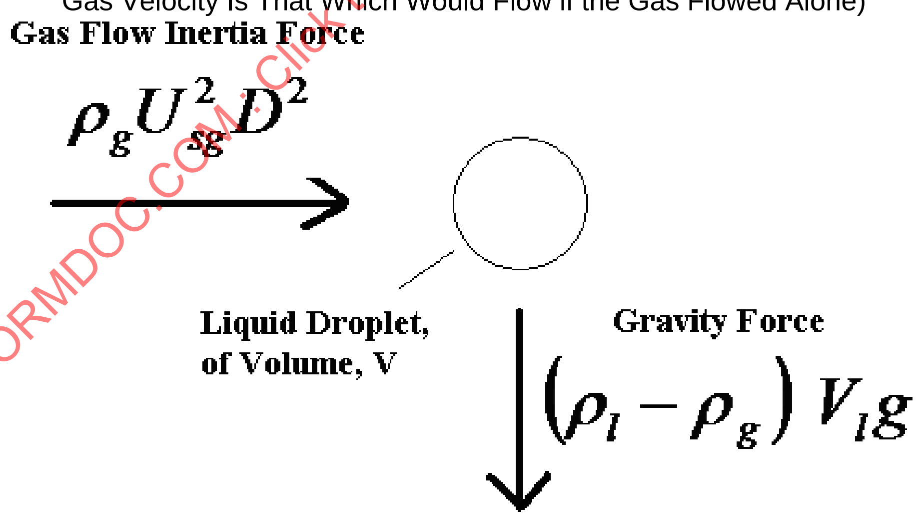
The liquid gravity force

$$F_{Liquid\ Gravity} = (\rho_l - \rho_g) V_l g \qquad \text{(A-27)}$$

where V_l is the volume of a droplet of liquid in a wet gas mist flow (that is, the cubic length dimension, $V_l = \frac{\pi}{6} D^3$).

Figure A-1 shows a diagram of the forces on a droplet.

Fig. A-1 Illustration of the Average Forces on a Droplet (Assuming the Average Gas Velocity Is That Which Would Flow if the Gas Flowed Alone)



Therefore:

$$Fr_g = \sqrt{\frac{Superficial\ Gas\ Inertia\ Force}{Liquid\ Gravity\ Force}} = \sqrt{\frac{\rho_g U_{sg}^2 D^2}{(\rho_l - \rho_g)\left(\frac{\pi}{6} D^3\right) g}} = \sqrt{\frac{6}{\pi}}\frac{U_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} \qquad \text{(A-28)}$$

In dimensional analysis theory all constants can be dropped and lengths are required to be defined. For example, single-phase pipe flow Reynolds number is defined with the inside bore of the pipe. Likewise here, *D* is the inside bore of the pipe. Therefore, eq. (A-28) reduces to eq. (6).

A-3 A MODIFIED WEBER NUMBER (We_{tp})

The standard Weber number is shown as eq. (10). A modified Weber number for wet gas metering is shown in eq. (11).

General definition:
$$\text{We} = \frac{\text{Inertia Forces}}{\text{Surface Tension Forces}} \quad (10)$$

Possible wet gas definition:
$$We_{tp} = \frac{\dot{m}_g^2}{\sigma_l \rho_g D^3} \quad (11)$$

Equation (11) is derived below.

The gas inertia force if gas flows alone:

$$\text{F}_{\text{Gas Inertia}} = \rho_g U_{sg}^2 D^2 \quad \text{(A-29)}$$

The superficial gas velocity is given by eq. (3).

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} \quad (3)$$

Liquid surface tension force:
$$F_{Surface\ Tension} = \sigma_l D \quad \text{(A-30)}$$

A modified Weber number for wet gas flow metering is given below as eqs. (A-31) and (A-32):

$$We_{tp} = \frac{Superficial\ Gas\ Inertia\ Force}{Liquid\ Surface\ Tension\ Forces} \quad \text{(A-31)}$$

$$We_{tp} = \frac{\rho_g U_{sg}^2 D^2}{\sigma_l D} = \frac{\rho_g D^2}{\sigma_l D}\left(\frac{\dot{m}_g}{\rho_g\left(\frac{\pi}{4}D^2\right)}\right)^2 = \left(\frac{16}{\pi^2}\right)\frac{\dot{m}_g^2}{\sigma_l \rho_g D^3} \quad \text{(A-32)}$$

As in dimensional analysis theory, all constants can be dropped and *D* is chosen as the inside bore of the pipe. Therefore eq. (A-32) reduces to eq. (11).

Note that surface tension values for stationary common liquids with an air interface are relatively well known (at atmospheric pressure) compared to interfacial tensions between flowing liquids and assorted gases and hence surface tension is the parameter used here for practical reasons (thereby implicitly making the unproven assumption that the static surface and dynamic interfacial tension values are similar). As yet little wet gas meter research is known to be published that discusses the effects of liquid properties. Steven [15] discusses preliminary DP meter research but much is unknown regarding the liquid effects on wet gas meters and at the time of writing there are proposed research tests by different organizations that include investigating the effect interfacial tension and other liquid property parameters.

A-4 A WORKED EXAMPLE OF THE EFFECT OF USING DIFFERENT LOCKHART–MARTINELLI PARAMETER DEFINITIONS

The following numerical example is reproduction of the example given by Steven [21]. Let us consider a wet natural gas flow in a 4 in. Schedule 80 pipe (i.e., an inside bore diameter of 0.09718 m) of a relative roughness level of 0.001. Let us say in this example the pressure is 50 bara at 300 K, the natural gas density is 38 kg/m^3, the gas viscosity is approximately 0.01124 cP and the gas flow rate is 400 m^3/h. Let us also say that the liquid flow rate is 1.937 kg/s, the liquid density is 800 kg/m^3 and the liquid viscosity is 1.92 cP. Let us further say that the pipe has an ASME MFC-3M compliant orifice plate meter of beta ratio 0.7 with *D* and *D*/2 tappings fitted. What are the original Lockhart–Martinelli parameter [eq. (A-10)], the Murdock parameter [eq. (A-19)] and the Chisholm parameter [eq. (A-23)] values and how do they compare to each other?

The Reynolds number of the gas phase if it flowed alone in the pipe is approximately 4.93e^6. The Reynolds number of the liquid phase if it flowed alone in the pipe is approximately 1.32e^4. From the Moody diagram for a relative roughness of 0.001 the gas friction factor f_g is 0.021 and the liquid friction factor f_l is 0.032.

Therefore, the original Lockhart–Martinelli parameter is:

$$X = \sqrt{\frac{\Delta P_{l_f}}{\Delta P_{g_f}}} = \sqrt{\frac{f_l}{f_g}}\,\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = 0.123 \qquad \text{(A-10)}$$

From ASME MFC-3M, we see that the orifice plate meter's discharge coefficient at a Reynolds number of 4.9e^6 is approximately 0.608 and at a Reynolds number of 1.32e^4 is approximately 0.639. The orifice plate expansibility equation given is shown as eq. (A-33) here:

$$Y = 1 - \left(0.351 + 0.256\beta^4 + 0.93\beta^8\right)\left[1 - \left(\frac{p_2}{p_1}\right)^{1/\kappa}\right] \qquad \text{(A-33)}$$

where β is the orifice plate beta ratio, κ is the isentropic exponent (which is approximately 1.3 for natural gas), and $\frac{p_2}{p_1}$ is the ratio of the downstream-to-upstream pressure tapping. The approximate throat tap pressure (p_2) can be found by taking the difference between the upstream pressure and the approximated differential pressure found by assuming the expansibility to be unity and rearranging eq. (A-17):

$$\Delta P_g \approx \frac{1}{2\rho_g}\left(\frac{\dot{m}_g}{EA_tC_d}\right)^2 \qquad \text{(A-17a)}$$

This approximate differential pressure value is 36,653 Pa. The throat pressure is therefore

$$P_2 = P_1 - \Delta P_g = 4{,}963{,}347 \;\; Pa$$

So we have $\frac{p_2}{p_1}$ as 0.993. The corresponding expansibility factor from eq. (A-33) is therefore 0.9974. Therefore, Murdock's parameter from eq. (A-19) is

$$X_{Murdock} = \sqrt{\frac{\Delta P_l}{\Delta P_g}} = \frac{C_{d_{gas}} Y}{C_{d_{liquid}}} \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} = 0.0949$$

Chisholm's parameter is calculated from eq. (A-23). That is,

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} = 0.1$$

Therefore, for the same wet gas flow conditions we have

$$X = 0.123\,,\quad X_{Murdock} = 0.0949\,,\quad X_{LM} = 0.1$$

As we promote X_{LM} as the appropriate value (due to the argument above) we see that the original Lockhart–Martinelli parameter is 23% higher and the Murdock parameter for this particular DP meter is 5.1% lower. Hence the difference is not trivial.

A-5 CORRELATION SENSITIVITY TO LOCKHART–MARTINELLI PARAMETER ERRORS

It should be noted that the error in the gas flow rate prediction when using a gas DP meter wet gas correlation when there is an error in the Lockhart–Martinelli parameter definition is, while still significant, not as serious as the errors in the Lockhart–Martinelli parameter itself due to the sensitivity of the correlations to the Lockhart–Martinelli parameter. To show this point let us continue the above worked example by applying Chisholm's orifice plate wet gas correlation [8]. Nonmandatory Appendix G explains that Chisholm stated that for an orifice plate meter the over-reading of the gas flow rate when the flow is wet gas can be computed by the following equation:

$$Over-Reading = \frac{\dot{m}_{g_{Apparent}}}{\dot{m}_g} = \sqrt{1 + \left\{\left(\left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} + \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}}\right) X_{LM}\right\} + \left(X_{LM}\right)^2} \qquad \text{(A-33)}$$

where $\dot{m}_{g_{Apparent}}$ is the gas flow rate prediction that is not corrected for the liquid induced error. [Equation (A-33) is obtained from eq. (G-13) of Nonmandatory Appendix G.] With a gas-to-liquid density ratio of 0.0475 the associated over-reading predictions are

$X = 0.123$ gives OR = 15.6%,
$X_{Murdock} = 0.0949$ gives OR = 12.1%,
$X_{LM} = 0.1$ gives OR = 12.7%

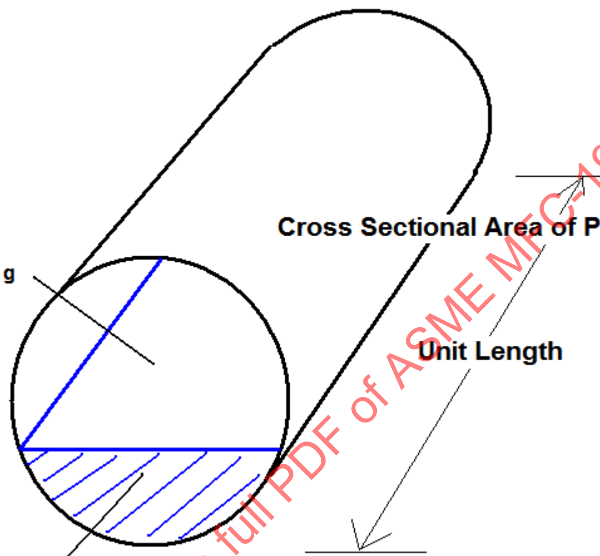
In this example Chisholm, of course, used $X_{LM} = 0.1$ so the Chisholm correlation is saying the over-reading (or gas flow rate predictions positive error) is 12.7%. However, an incorrect choice of definition can cause an error. In this case a -0.6% underestimation (and a corresponding undercorrection) for

$X_{Murdock} = 0.0949$ and a 2.9% overestimation (and a corresponding overcorrection) for $X = 0.123$. This is a bias, not an uncertainty.

The prediction of a DP meter's wet gas correlation is therefore affected by which parameter is being used. Different meters and correlations will have different sensitivities to varying Lockhart–Martinelli parameter values. (A DP meter designs sensitivity to errors in Lockhart–Martinelli parameter are described by Steven [13] for cases where the Lockhart–Martinelli parameter is in error due to incorrect liquid mass flow estimates. However, the description of the effect generic errors on the Lockhart–Martinelli parameter have on DP meter wet gas correlations is also valid for this argument.) For Chisholm's equation, used for the example above, the gas flow rate prediction gets more sensitive to errors in the Lockhart–Martinelli parameter as the gas-to-liquid density ratio reduces. It should also be noted that the Venturi meter has significantly greater sensitivity than the orifice plate, so a similar example with a Venturi would cause larger gas flow rate prediction errors. It should also be noted that these are errors and not uncertainties. Chisholm's correlation is said to give the overreading to an uncertainty of 2% so small differences may not be noticeable, but the added error in this case is avoidable. In many real wet gas flowmetering cases the error could be above 2%.

NONMANDATORY APPENDIX B[1]
DIFFERENCE BETWEEN THE GAS VOLUME FRACTION AND THE GAS TO TOTAL VOLUME RATIO PER UNIT LENGTH OF PIPE IN STEADY FLOW

Fig. B-1 Stratified Flow Pattern in a Pipe



B-1 GENERAL

The gas volume fraction (GVF) is defined as the ratio of the gas volume flow rate to the total volume flow rate [see eq. (12)]. That is:

$$\text{GVF} = \frac{\dot{Q}_g}{\dot{Q}_g + \dot{Q}_l} \qquad (12)$$

Unfortunately, the name "gas volume fraction" has led to some confusion in the industry as this is sometimes mistakenly thought to be the actual ratio of the volume of the gas to the pipe volume along a unit section of pipe in two-phase flow. Calling this term the "gas to pipe volume ratio" this is shown in eq. (B-1).

$$Gas\ to\ Pipe\ Volume\ Ratio = \frac{V_g}{V_g + V_l} = \frac{A_g L}{A_g L + A_l L} = \frac{A_g}{A_g + A_l} \qquad \text{(B-1)}$$

where
A_g = gas cross sectional area
A_l = liquid cross sectional area
L = unit length of constant area
V_g = gas volume in a unit volume of constant area pipe

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[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

V_l = liquid volume in a unit volume of constant area pipe

The gas volume fraction is not the same parameter as the gas-to-pipe volume ratio unless the average gas velocity and average liquid velocity are the same. That is, the "slip ratio" denoted by "S_R" here, which is defined as the ratio of the average gas and liquid velocities [see eq. (21)] must be unity for the GVF, and the gas-to-pipe volume ratio parameters to be equal.

$$S_R = \frac{\bar{U}_g}{\bar{U}_l} \tag{21}$$

An explanation of this statement is as follows:

$$\dot{Q} = A\bar{U} \tag{B-2}$$

where eq. (B-2) is the general volume flow rate equation for any phase. Therefore, eq. (12) can be written as eq. (B-3):

$$GVF = \frac{\dot{Q}_g}{\dot{Q}_g + \dot{Q}_l} = \frac{A_g\bar{U}_g}{A_g\bar{U}_g + A_l\bar{U}_l} = \frac{A_g}{A_g + \left(\frac{\bar{U}_l}{\bar{U}_g}\right)A_l} = \frac{A_g}{A_g + \left(\frac{1}{S_R}\right)A_l} \tag{B-3}$$

whereas

$$Gas\ to\ Pipe\ Volume\ Ratio = \frac{V_g}{V_g + V_l} = \frac{A_g L}{A_g L + A_l L} = \frac{A_g}{A_g + A_l} \tag{B-1}$$

Therefore, the GVF is equal to the gas-to-pipe volume ratio only when the slip ratio is unity. Unfortunately, in most two-phase flows this is not the case. In fact, the slip value, while very difficult to measure in practice, is known to often be considerable, so an approximation of unity is in the vast majority of cases not valid. This is the technical reason devices that estimate the phase fractions/liquid volume-to-pipe volume ratios (such as capacitance meters) cannot be used in conjunction with a single-phase meter to make a wet gas meter system (i.e., a meter that can meter both liquid and gas phases simultaneously) without sophisticated flow modeling to predict the slip. That is, there is a common misconception that such devices measure the GVF. If the GVF could be independently measured in some way the value would be available for input into eq. (22) and hence for known gas and liquid densities the Lockhart–Martinelli parameter would be known.

$$X_{LM} = \frac{1-(GVF)}{(GVF)}\sqrt{\frac{\rho_l}{\rho_g}} \tag{22}$$

This, in turn, would allow the use of any single-phase meter wet gas correlation based on the Lockhart–Martinelli parameter to predict the gas and liquid flow rates. However, capacitance meters estimate the gas-to-pipe volume ratio — not the GVF. Hence, the simple combination of a single-phase meter with a wet gas correlation and device that estimates the gas-to-pipe volume ratio does not produce a wet gas meter. There are systems on the market that use combinations of phase fraction devices and DP meters to produce multiphase meters that work in the wet gas flow region but these use sophisticated proprietary flow modeling techniques and multiple data sets to predict slip.

It is also noteworthy that there is a very rarely used parameter in existence that indicates the difference between GVF and void fraction. It was quoted by Chisholm in his derivation of eq. (G-12) and it is called the Armand coefficient (C_A). This is shown in eq. (B-4).

$$C_A = \frac{\alpha_g}{GVF} \quad \text{(B-4)}$$

Note that by substituting eqs. (B-3), (18), and (19),

$$\alpha_g = \frac{A_g}{A} \quad (18) \qquad \alpha_l = 1 - \alpha_g = \frac{A_l}{A} \quad (19)$$

into eq. (B-4) we get:

$$C_A = \alpha_g + \frac{\alpha_l}{S_R} \quad \text{(B-5)}$$

For wet gas flows, it is typically correct to say $\alpha_g >> \alpha_l$ and $S_R > 1$, especially at higher GVF values. Therefore, a reasonable approximation for many high GVF wet gas flows is $C_A \approx \alpha_g$. Therefore, in some lower loading wet gas flows $C_A \approx \alpha_g \approx 1$.

Chisholm approximated the Armand coefficient to unity from his available experimental evidence. That is, Chisholm effectively stated

$$GVF \approx \alpha_g \approx 1 \quad \text{(B-6)}$$

In some practical cases this is a reasonable assumption as the difference in GVF and void fraction can be small. However, it cannot be guaranteed to be an appropriate assumption in all cases. It is of interest to continue the worked example started in Nonmandatory Appendix H here to show the difference in the GVF and void fraction values. This work was originally produced by Steven [20] and reproduced here with permission.

Chisholm created a slip model [8] (which is derived in Nonmandatory Appendix H) which stated for stratified flow the slip of a wet gas flow was

$$S_R = \frac{\bar{U}_g}{\bar{U}_l} = \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} \quad \text{(H-202)}$$

(It should be noted that there are several slip models in the literature and that they should be applied according to the limits of the models assumptions and the suitability of the model to the application in question. We make no claim on the validity of any particular slip model in this Report.) As in the example, the gas density is 38 kg/m^3 and the liquid density is 800 kg/m^3. The slip is according to Chisholm's slip model:

$$S_R = \frac{\bar{U}_g}{\bar{U}_l} = \left(\frac{800}{38}\right)^{\frac{1}{4}} = 2.14$$

That is, according to Chisholm, the average gas velocity is 2.14 times that of the average liquid velocity. Taking eqs. (18), (B-4), and (B-5), we can derive eq. (B-7):

$$\alpha_g = \frac{1}{1 + \left\{ S_R \left(\frac{1}{GVF} - 1 \right) \right\}} \qquad \text{(B-7)}$$

Therefore, according to Chisholm's slip model in our example, with 400 m^3/hr of gas and 8.72 m^3/hr of liquid, we have a GVF of 0.979,

$$GVF = \frac{\dot{Q}_g}{\dot{Q}_g + \dot{Q}_l} = \frac{400}{400 + 8.72} = 0.979$$

And the void fraction is:

$$\alpha_g = \frac{1}{1 + \left\{ S_R \left(\frac{1}{GVF} - 1 \right) \right\}} = \frac{1}{1 + \left\{ 2.14 * \left(\frac{1}{0.979} - 1 \right) \right\}} = 0.956$$

Hence the hold up is from eq. (19):

$$\alpha_l = 1 - 0.956 = 0.0439$$

The Armand coefficient (C_A) is:

$$C_A = \frac{\alpha_g}{GVF} = \frac{0.956}{0.979} = 0.977$$

The Armand coefficient is unity when there is no slip, i.e., $S_R = 1$. The GVF and the void fraction are not the same parameter. This particular example uses the flow conditions as the example in Nonmandatory Appendix A and a small difference in GVF and void fraction is noticeable. This then proves the difference in the parameters. In this particular example the difference is in fact relatively small but it is important to note that other flow conditions and other slip models can have greater differences between the parameter values.

NONMANDATORY APPENDIX C[1]
INCOMPATIBILITY OF DIFFERENT SUGGESTED WET GAS DEFINITIONS

C-1 GENERAL

There has historically been several wet gas definitions across different industries. The best known of these have been a Lockhart–Martinelli parameter (X_{LM}) limit of no greater than either 0.3 or 0.35, a minimum quality (x) of any limit down to zero, or a quality greater than 0.5 and a gas volume fraction (GVF) of 0.9 or above. The most common of these appears to be $X_{LM} \le 0.3$, x $\ge$ 0.5 and a GVF $\ge$ 0.9. It is important to understand that these definitions are not equivalent. In fact they can be very different. For example, consider eqs. (22) through (24).

$$X_{LM} = \frac{1-(GVF)}{(GVF)}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} \tag{22}$$

$$GVF = \frac{\sqrt{\frac{\rho_l}{\rho_g}}}{X_{LM} + \sqrt{\frac{\rho_l}{\rho_g}}} = \frac{\left(\frac{\rho_l}{\rho_g}\right)}{\left(\frac{\rho_l}{\rho_g}\right) + \left(\frac{1-x}{x}\right)} \tag{23}$$

$$x = \frac{1}{1 + X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1 + \left(\left(\frac{1-(GVF)}{(GVF)}\right) * \frac{\rho_l}{\rho_g}\right)} \tag{24}$$

If we were to assume equivalence and set the maximum liquid loading at both X_{LM} = 0.3 and x = 0.5, then eqs. (22) through (24) dictate that only one gas-to-liquid density ratio and one GVF can exist for both these maximum wet gas limits to be met simultaneously. In fact, setting any two of these three different definitions automatically sets the third and the gas-to-liquid density ratio. That is:

For setting X_{LM} = 0.3 and x = 0.5 then the GVF and density ratio are then fixed:

$$X_{LM} = 0.3$$
$$x = 0.5$$
$$GVF = 0.917$$
$$L/G\ DR = 11.11$$

For setting GVF = 0.9 and x = 0.5 then the Lockhart–Martinelli parameter and density ratio are then fixed:

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

$$X_{LM} = 0.333$$
$$x = 0.5$$
$$GVF = 0.9$$
$$L/G\ DR = 9.0$$

For setting X_{LM} = 0.3 and GVF = 0.9 then the quality and density ratio are then fixed:

$$X_{LM} = 0.3$$
$$x = 0.552$$
$$GVF = 0.9$$
$$L/G\ DR = 7.29$$

Figure C-1 graphically shows the differences between the Lockhart–Martinelli parameter wet gas definition (i.e., $X_{LM} \leq 0.3$), quality wet gas definition (i.e., $x > 0.5$), and the GVF wet gas definition (i.e., GVF > 0.9). The graph shows the relationship between X_{LM} and x at different gas-to-liquid density ratios. (Note "Dense Phase Conditions" is when the pressure is such that the gas-to-liquid density ratio is unity. In industry it is an extremely rare occurrence for dense phase conditions to be approached. An application to the right of this line indicating gas-to-liquid density ratios greater than unity is not realistic. A Lockhart–Martinelli parameter of less than 0.3 is seen as below the red horizontal line. A quality of greater than 0.5 is to the right of the green vertical line. GVFs greater than 0.9 are below the blue curve. Hence the three different well-known definitions for a border between wet gas flow and general two-phase flow are marked on the graph and it is clearly seen in Fig. A-1, which magnifies these borders that each definition can call certain flow conditions wet gas flow that the other two definitions call general two-phase flow. There is only a relatively small area in Fig. C-2 under which all three definitions agree the flow is a wet gas flow. Hence we only accept one definition for wet gas flow (i.e., $X_{LM} \leq 0.3$).

Fig. C-1 Relationship Between X_{LM}, x, and GVF and the Gas-to-Liquid Density Ratio

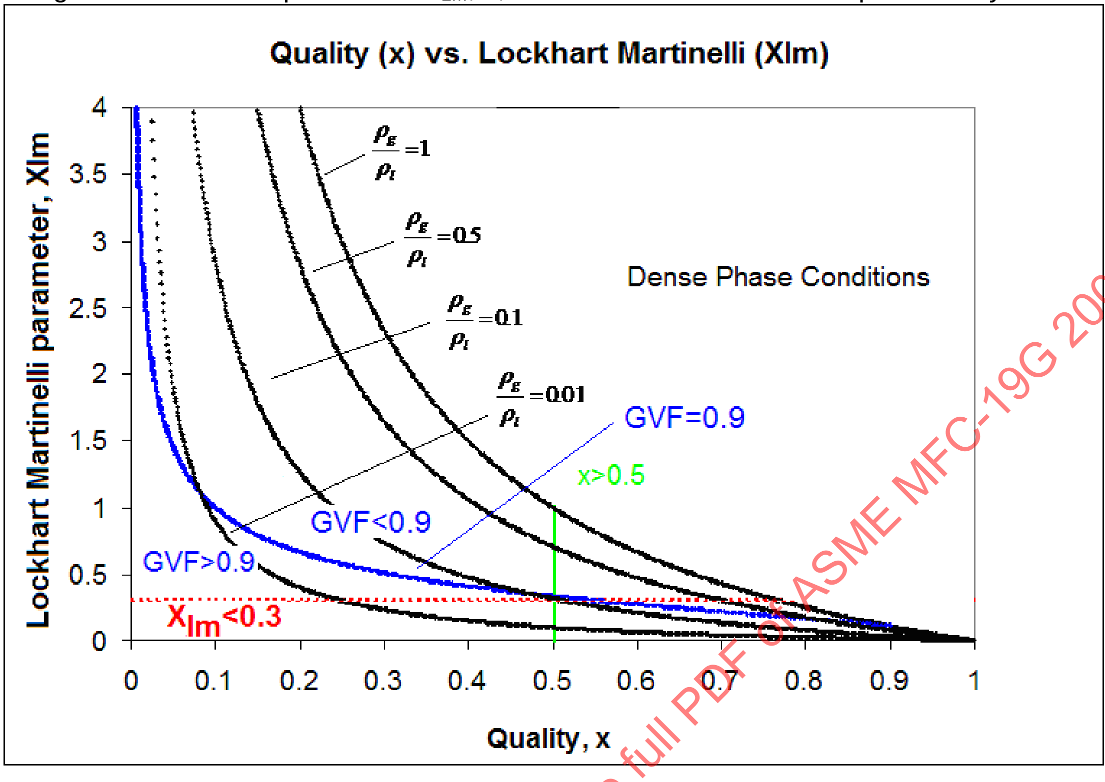
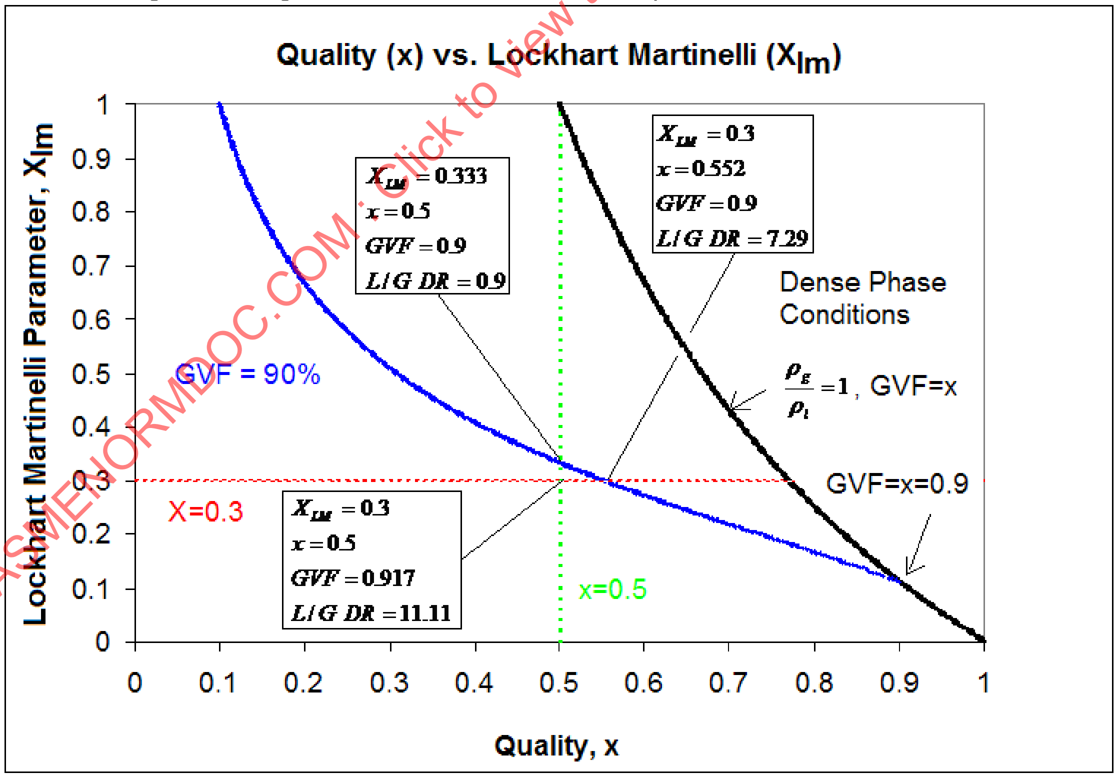


Fig. C-2 Magnified View of the Relationship Between X_{LM}, x, and GVF



NONMANDATORY APPENDIX D[1]
EQUATIONS AND GRAPHS FOR CONVERSIONS OF WET GAS FLOW PARAMETERS

D-1 GENERAL

In real-world situations, engineers seldom know the Lockhart–Martinelli parameter directly. The information on what quantity of liquid is flowing with a gas flow comes in many different ways. It is necessary for an engineer to relate this information to the Lockhart–Martinelli parameter and useful to relate it to the other well-known parameters for comparison. These calculations are done with eqs. (22) to (27). That is:

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\dot{Q}_l}{\dot{Q}_g}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{1-(GVF)}{(GVF)}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{(LVF)}{1-(LVF)}\sqrt{\frac{\rho_l}{\rho_g}} \qquad (22)$$

$$GVF = \frac{1}{1+\left(\frac{m_l}{m_g}*\frac{\rho_g}{\rho_l}\right)} = \frac{1}{\left(\frac{\dot{Q}_l}{\dot{Q}_g}\right)+1} = 1-(LVF) = \frac{\sqrt{\frac{\rho_l}{\rho_g}}}{X_{LM}+\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{\left(\frac{\rho_l}{\rho_g}\right)}{\left(\frac{\rho_l}{\rho_g}\right)+\left(\frac{1-x}{x}\right)} \qquad (23)$$

$$x = \frac{1}{1+\left(\frac{\dot{m}_l}{\dot{m}_g}\right)} = \frac{1}{1+\left(\frac{\rho_l}{\rho_g}\frac{\dot{Q}_l}{\dot{Q}_g}\right)} = \frac{1}{1+X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1+\left(\left(\frac{1-(GVF)}{(GVF)}\right)*\frac{\rho_l}{\rho_g}\right)} = \frac{1}{1+\left(\left(\frac{(LVF)}{1-(LVF)}\right)*\frac{\rho_l}{\rho_g}\right)} \qquad (24)$$

$$\frac{\dot{m}_l}{\dot{m}_g} = X_{LM}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{\rho_l}{\rho_g}\frac{\dot{Q}_l}{\dot{Q}_g} = \frac{1-(GVF)}{GVF}\frac{\rho_l}{\rho_g} = \frac{LVF}{1-LVF}\frac{\rho_l}{\rho_g} = \frac{1-x}{x} \qquad (25)$$

$$\frac{\dot{Q}_l}{\dot{Q}_g} = X_{LM}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{\rho_g}{\rho_l}\frac{\dot{m}_l}{\dot{m}_g} = \frac{1-(GVF)}{GVF} = \frac{LVF}{1-(LVF)} = \left(\frac{1-x}{x}\right)\frac{\rho_g}{\rho_l} \qquad (26)$$

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

$$LVF = 1-(GVF) = \frac{X_{LM}\sqrt{\frac{\rho_g}{\rho_l}}}{1+X_{LM}\sqrt{\frac{\rho_g}{\rho_l}}} = \frac{\left(\frac{1-x}{x}\right)}{\left(\frac{1-x}{x}\right)+\frac{\rho_l}{\rho_g}} = \frac{\frac{\rho_g}{\rho_l}}{\frac{\rho_g}{\rho_l}+\frac{\dot{m}_g}{\dot{m}_l}} = \frac{\frac{\dot{Q}_l}{\dot{Q}_g}}{1+\frac{\dot{Q}_l}{\dot{Q}_g}} \quad (27)$$

Note that all gas volume flows are at flow conditions.

The application of these equations [(22) to (27)] will for any given two-phase flow definition relate the given condition and gas-to-liquid density ratio to the other common terms (i.e., one of the six common methods of describing the two-phase flow). However, it can be preferable to use graphs for reasons of speed and convenience. The graphical representation of this data is now discussed.

Figure D-1 shows the quality relationship with the gas-to-liquid density ratio and the Lockhart–Martinelli parameter.

Figure D-2 shows the GVF relationship with the gas-to-liquid density ratio and the Lockhart–Martinelli parameter.

Figure D-3 shows the LVF relationship with the gas-to-liquid density ratio and the Lockhart–Martinelli parameter. It should be noted that $LVF = 1-(GVF)$ and therefore Figs. D-2 and D-3 have overlapping constant lines. The only difference is Fig. D-2 shows these lines as constant GVF lines whereas Fig. D-3 shows these lines as constant LVF lines. The relative values of the constant parameters are dictated by the $LVF = 1-(GVF)$ relationship.

Figure D-4 shows constant liquid mass flow rate to gas mass flow rate ratio lines plotted on gas-to-liquid density ratio to the Lockhart–Martinelli parameter.

Fig. D-1 Constant Quality (x) Lines Plotted on Gas-to-Liquid Density Ratio to the Lockhart–Martinelli Parameter

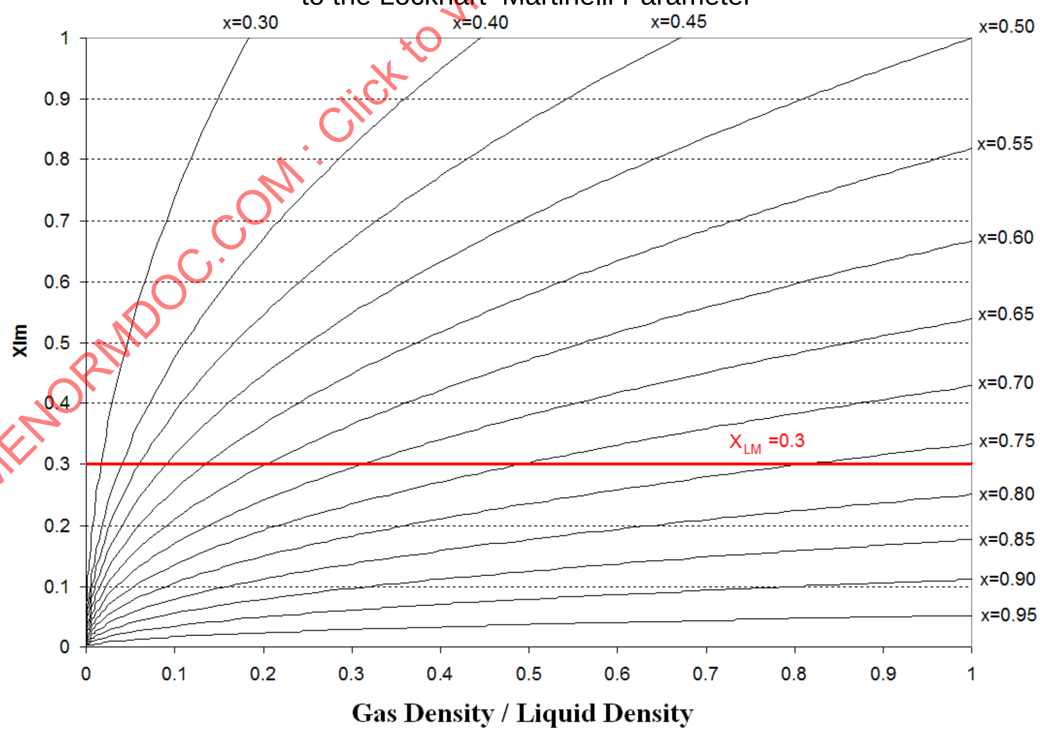


Fig. D-2 Constant GVF Lines Plotted on Gas-to-Liquid Density Ratio to the Lockhart–Martinelli Parameter

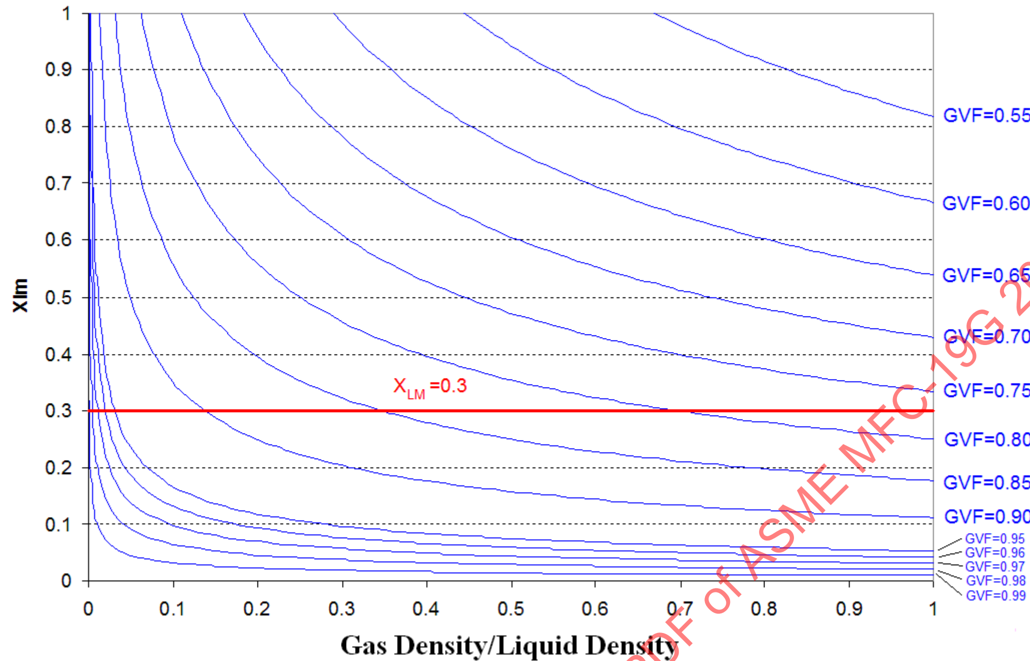


Fig. D-3 Constant LVF Lines Plotted on Gas-to-Liquid Density Ratio to the Lockhart–Martinelli Parameter

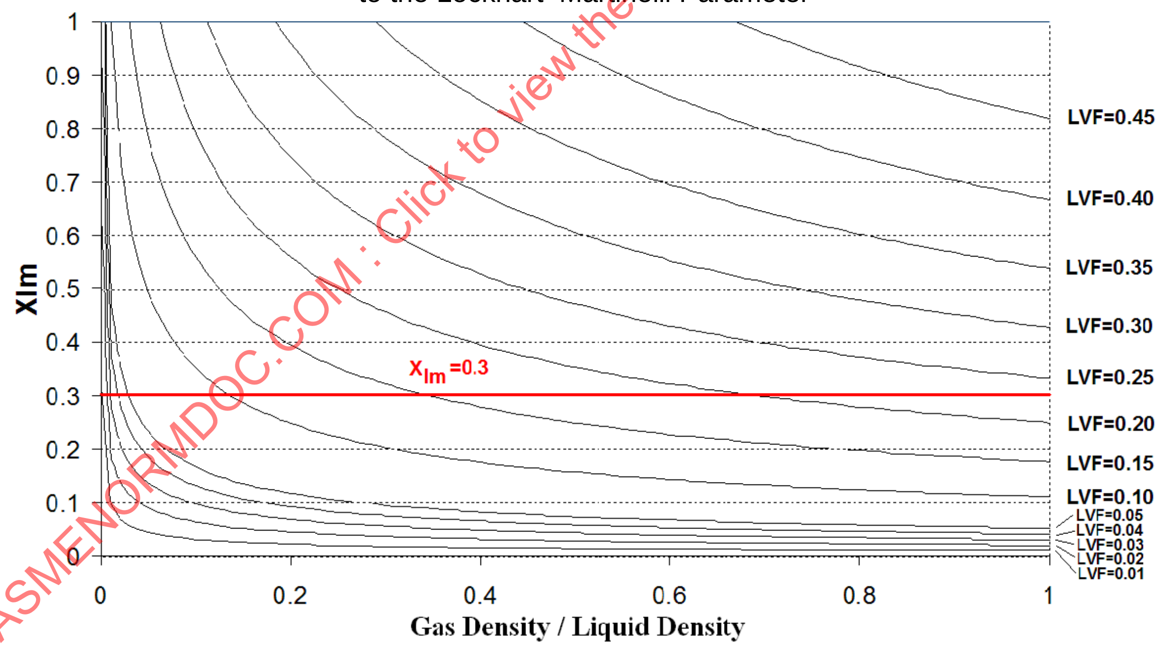


Fig. D-4 Constant Liquid Mass Flow Rate to Gas Mass Flow Rate Ratio Lines Plotted on Gas-to-Liquid Density Ratio to the Lockhart–Martinelli Parameter

(where M l/g indicates the liquid-to-gas mass flow ratio, i.e. $\dot{m}_l / \dot{m}_g$)

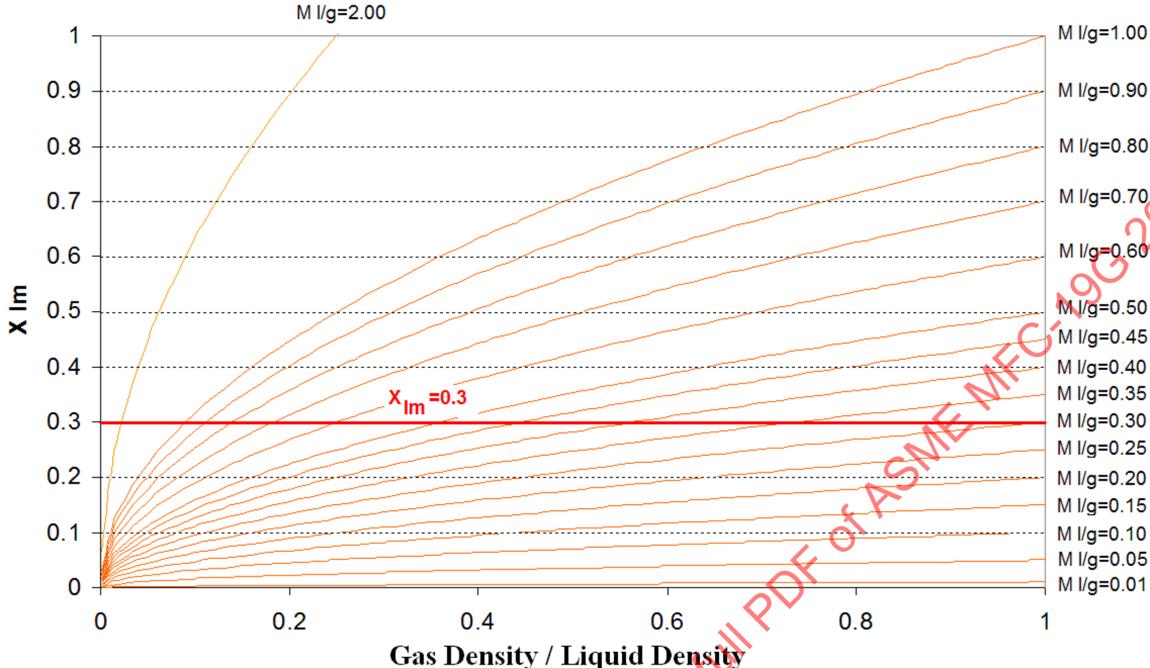
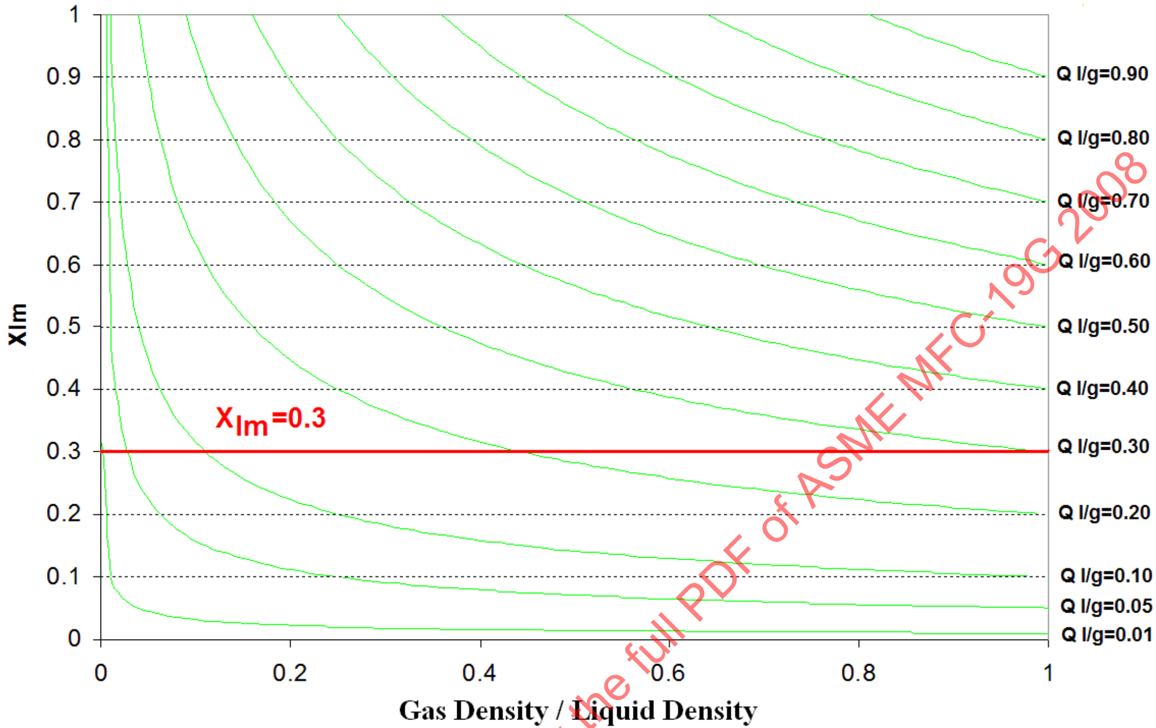


Figure D-5 shows lines of constant liquid volume flow rate to gas volume flow rate ratio plotted on gas-to-liquid density ratio to the Lockhart–Martinelli parameter.

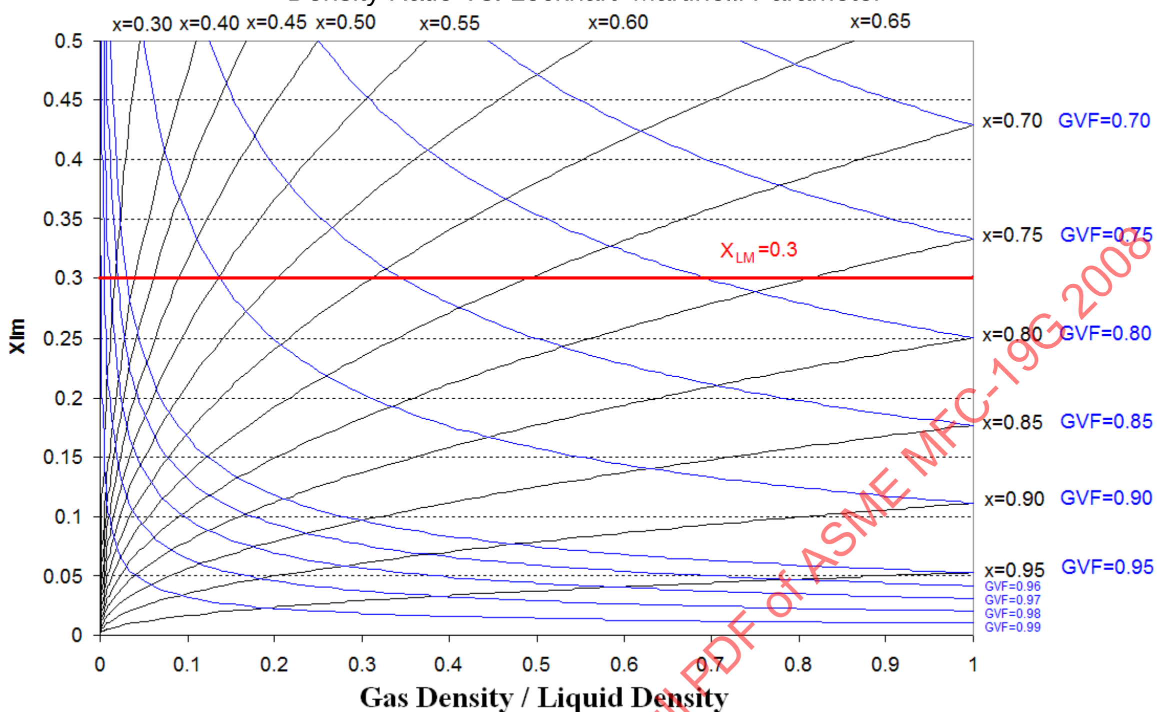
Fig. D-5 Constant Liquid Volume Flow Rate to Gas Volume Flow Rate Ratio Lines Plotted on Gas-to-Liquid Density Ratio to the Lockhart–Martinelli Parameter

(where Q l/g indicates the liquid-to-gas mass flow ratio, i.e., $\dot{Q}_l / \dot{Q}_g$)



From the graphs represented in Figs. D-1 through D-5 for any given value of gas-to-liquid density ratio and one description of the liquid content of a wet gas [i.e., one value from the list of quality (*x*), gas volume fraction (GVF), liquid volume fraction (LVF), liquid-to-gas mass flow rate or liquid-to-gas volume flow rate] the Lockhart–Martinelli parameter, X_{LM}, can be found from the relevant graph and then the remaining parameters can be found from the other graphs. In fact, except to maintain clarity there is no need for separate graphs as these individual graphs can be combined into one graph that shows direct relationships. For example, Fig. D-6 shows both the constant quality (*x*) and constant GVF lines on the same gas-to-liquid density ratio versus Lockhart–Martinelli parameter graph.

Fig. D-6 Both the Quality (x), and the Gas Volume Fraction (GVF), on the Gas-to-Liquid Density Ratio Vs. Lockhart–Martinelli Parameter



D-2 WORKED EXAMPLE

A wet natural gas flow is known to have a gas density of 70 kg/m^3 and a liquid (condensate) density of 700 kg/m^3. The gas-to-liquid density ratio is therefore 0.1. If the gas volume fraction is stated to be 0.95, then from eq. (14):

$$X_{LM} = \frac{1-(GVF)}{(GVF)}\sqrt{\frac{\rho_l}{\rho_g}} = \frac{1-0.95}{0.95}\sqrt{\frac{700}{70}} = 0.1664$$

and from eq. (16):

$$x = \frac{1}{1+\left(\left(\frac{1-(GVF)}{(GVF)}\right)*\frac{\rho_l}{\rho_g}\right)} = \frac{1}{1+\left(\left(\frac{1-0.95}{0.95}\right)*\left(\frac{700}{70}\right)\right)} = 0.655$$

or from eq. (14):

$$x = \frac{1}{1+X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1+\left(0.1664*\sqrt{\frac{700}{70}}\right)} = 0.655$$

This result can also be found with no calculation required, if not to four decimal places, by use of Fig. D-6. (Note that due to flow condition prediction and wet gas meter uncertainties it is rare in real world applications that these parameters have meaningful numbers after the second decimal place.) Figure D-

7 shows the crossing point of the gas-to-liquid density ratio of 0.1 and the GVF of 0.95 is seen to be at the y-axis (i.e., Lockhart–Martinelli parameter) value of 0.1644 and on the constant quality line of 0.655.

Figure D-6 shows just the GVF and quality on the graph but all of the parameters can be plotted together. The first full graph shown (Fig. D-8) is for a considerably wider two-phase range than wet gas flow. This is shown because an initial investigation required by meter users will be to check if the flow is a two-phase or a wet gas flow. Only if it is found that the Lockhart–Martinelli parameter is less or equal to 0.3 will the more detailed wet gas graph shown as Fig. D-9 be required. In fact this plot is further magnified in Fig. D-10 as the vast majority of industrial applications do not have gas-to-liquid densities greater than 0.5. (The most common high gas-to-liquid density ratio in industry is probably high pressure saturated steam. For example at 210 bar (with a saturation temperature of 369.8^{0}C) the steam density is approximately 201 kg/m^3. At that condition water has a density of approximately 469 kg/m^3. That is a gas-to-liquid density of approximately 0.43.) Most applications have gas-to-liquid density ratios less than 0.15. Figure D-11 shows the graph magnified to a gas-to-liquid density ratio less than 0.2.

Fig. D-7 Worked Example of Using the Gas-to-liquid Density Ratio Vs. Lockhart–Martinelli Parameter to Predict the Different Wet Gas Flow Liquid Loading Parameters

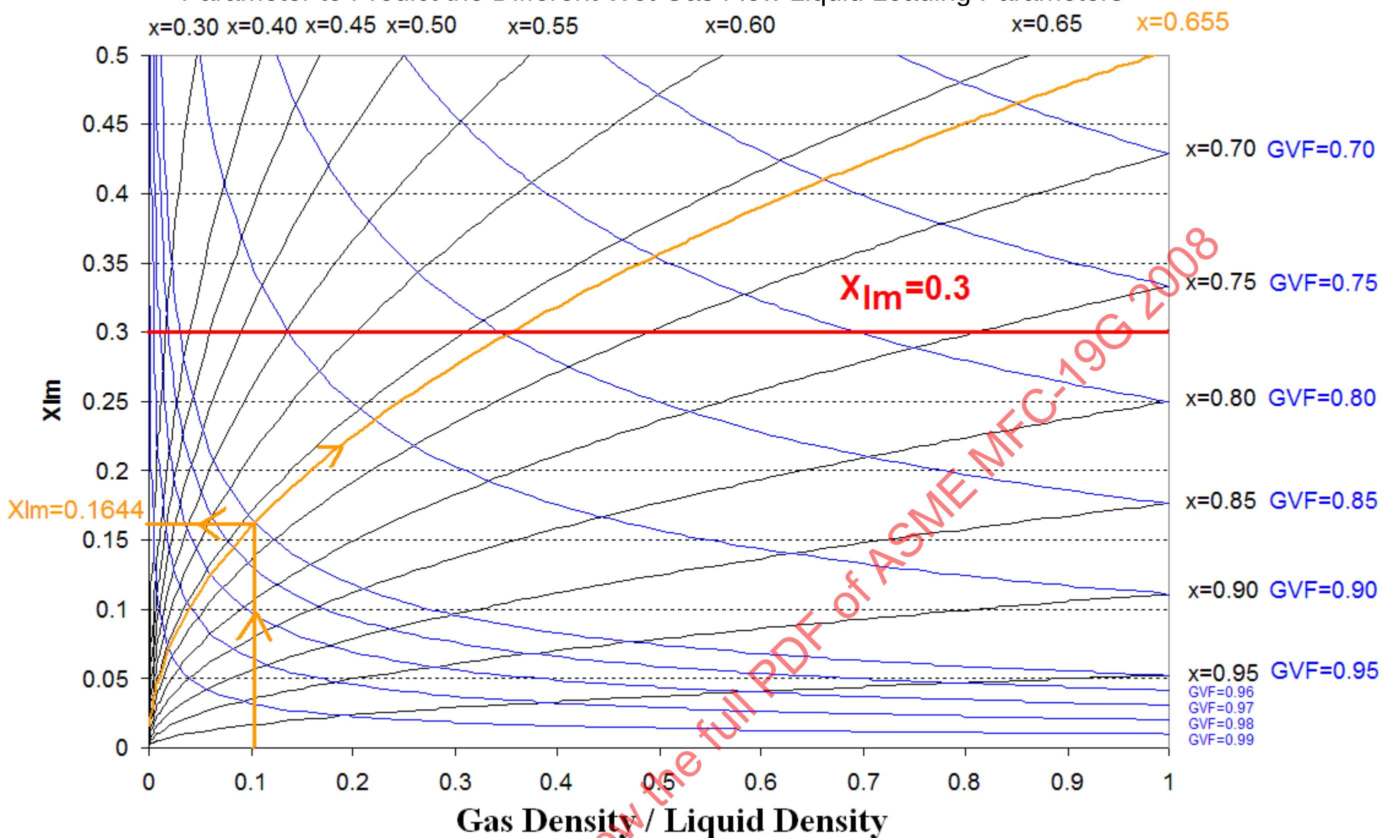


Fig D-8 Graph to Check If Flow Is Wet Gas Flow or General Two Phase Flow

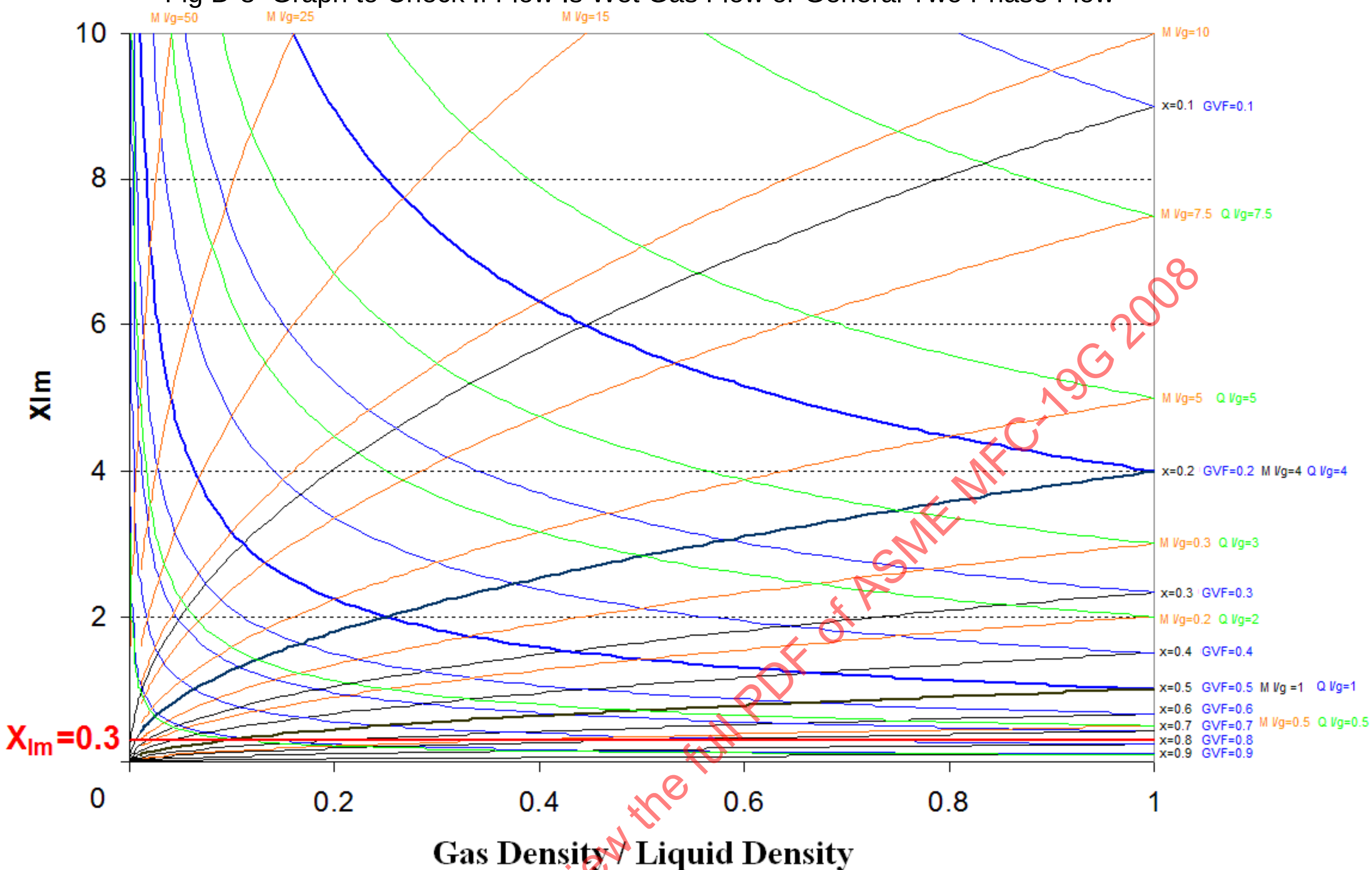


Fig. D-9 General Wet Gas Flow Liquid-to-Gas Flow Ratio Parameter Relationships

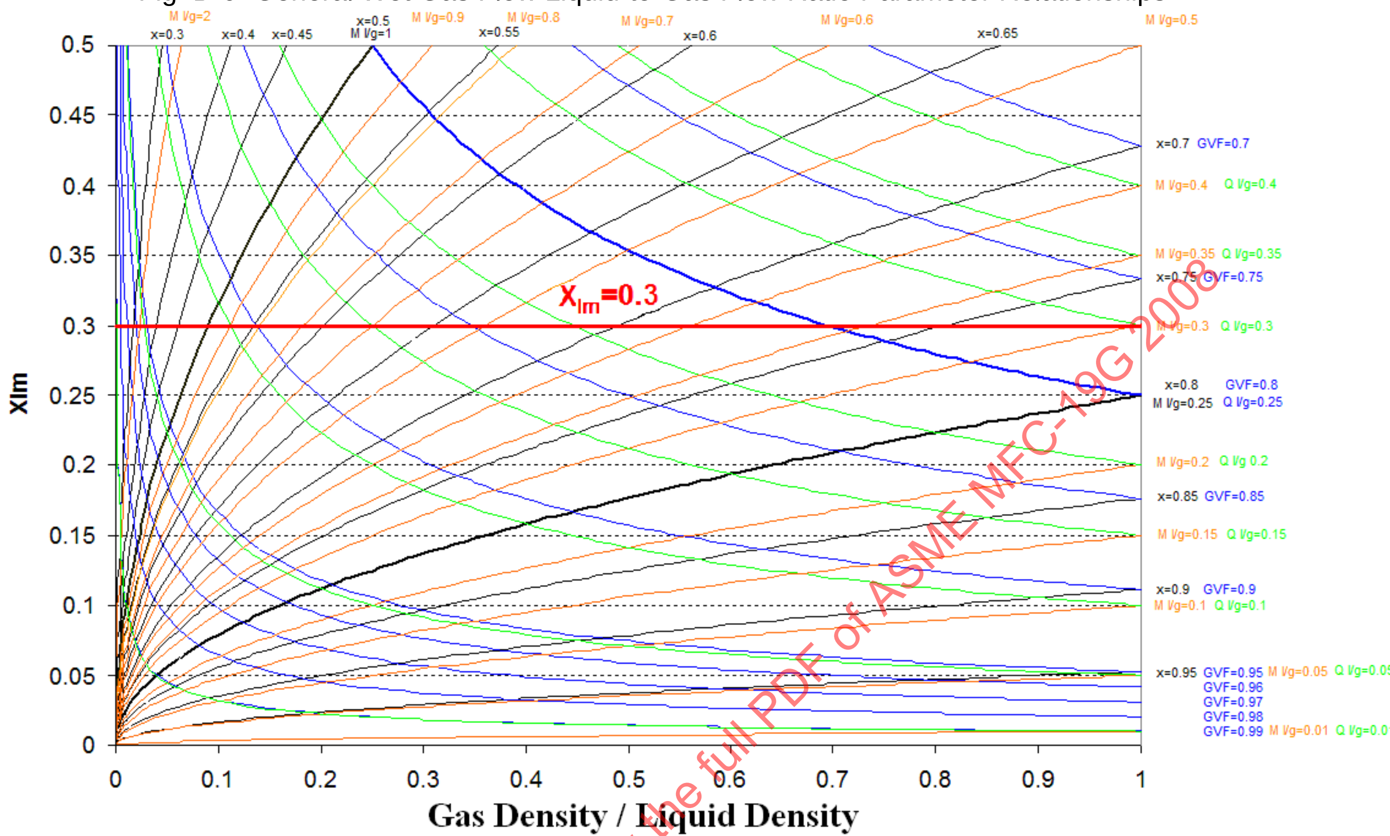


Fig. D-10 Magnified Graph of Wet Gas Flow Liquid-to-Gas Flow Ratio Parameter Relationships

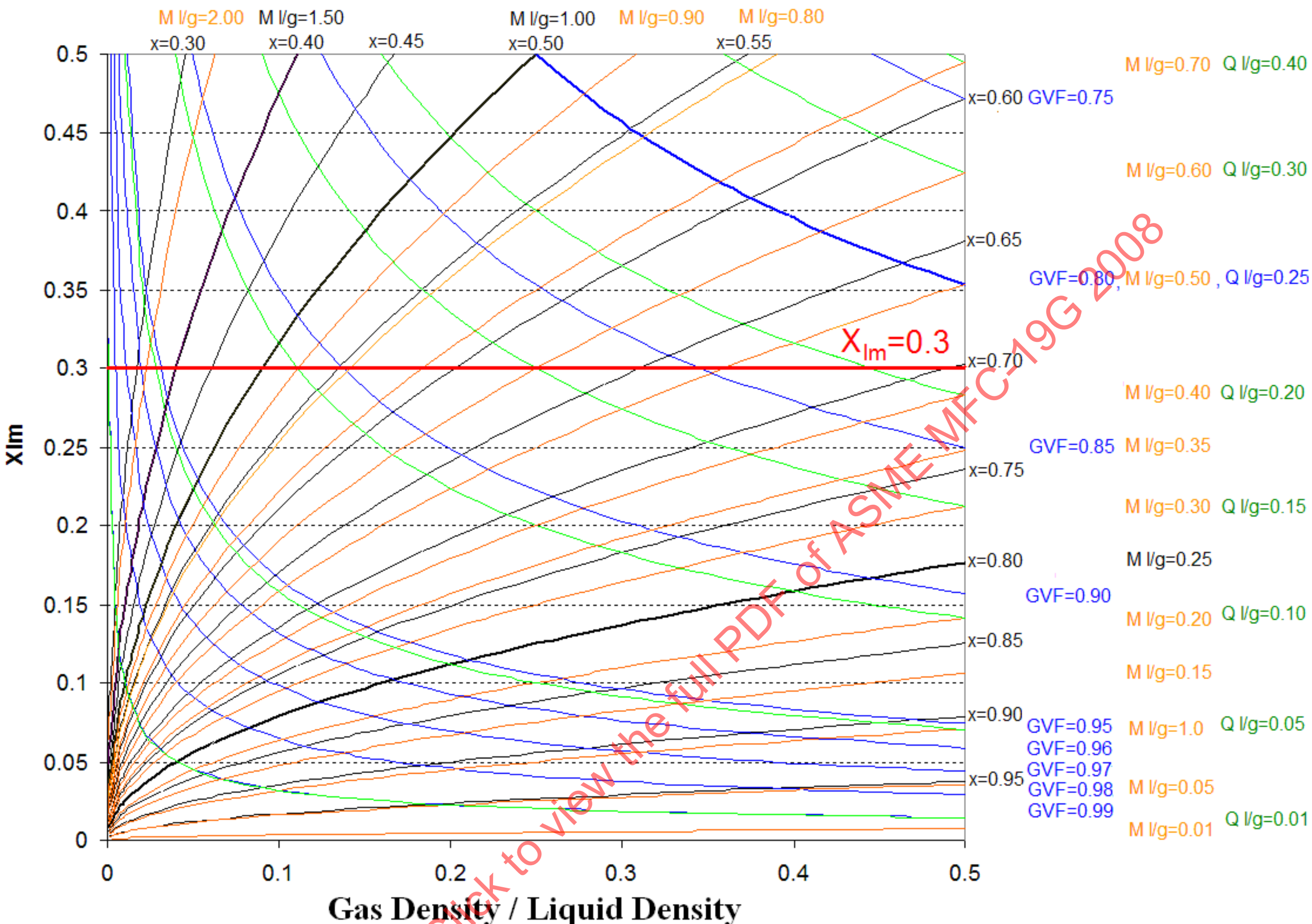
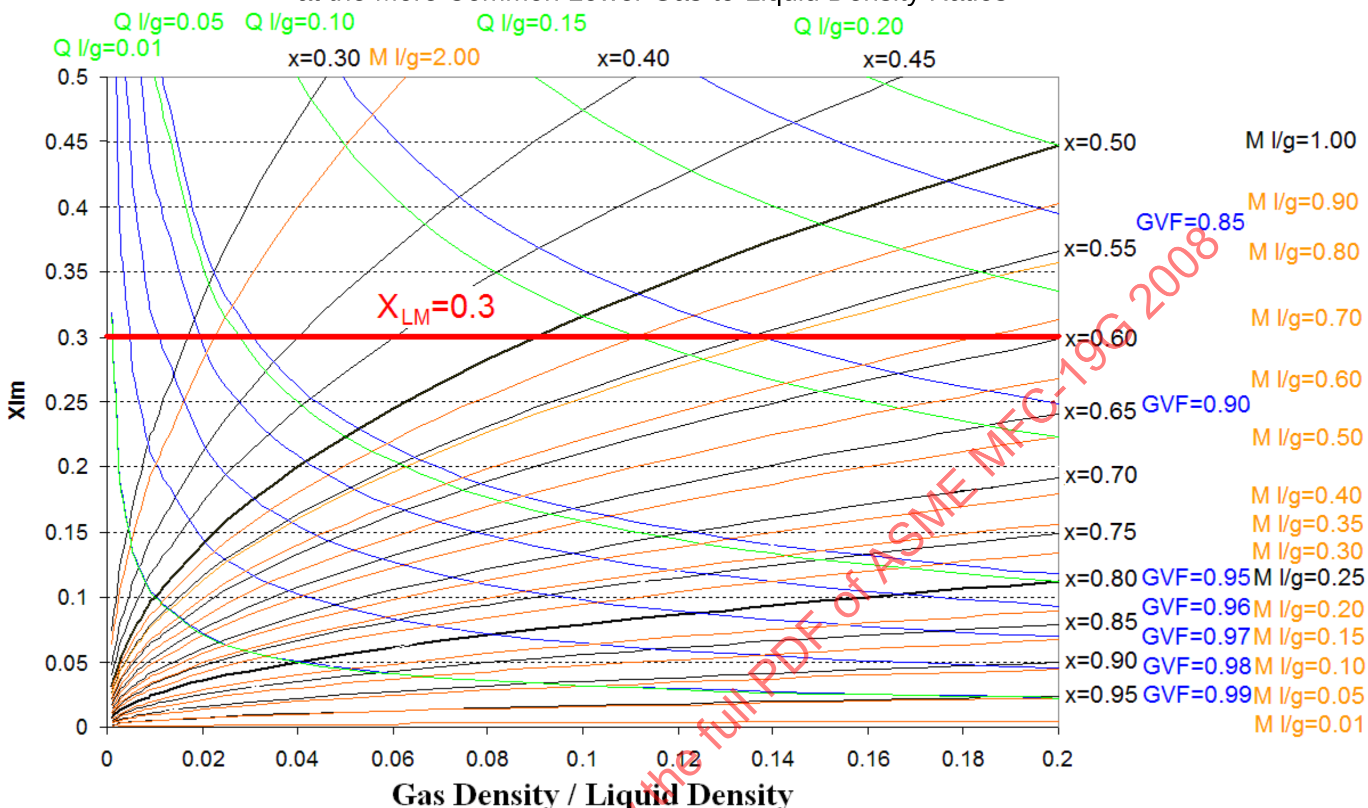


Fig. D-11 Magnified Graph of Wet Gas Flow Liquid-to-Gas Flow Ratio Parameter Relationships at the More Common Lower Gas-to-Liquid Density Ratios



NONMANDATORY APPENDIX E[1]
API WET GAS DEFINITIONS

E-1 GENERAL

The American Petroleum Institute (API) released a report [2] in 2004 where wet gas flow was defined and characterized. This wet gas flow definition used the Lockhart–Martinelli parameter as the primary parameter but was referenced to the gas volume fraction (GVF). It is understood to be the case, that although the API report [2] states the definition, it is not as of the time of writing of this report officially ratified by API. The definition in question is based on the practical experience of the oil and gas industry with regard to meter selection for different scenarios. As this Report is a generic industry publication (and therefore not solely based on the oil and gas industry) and no one wishes to link the definition of wet gas flow to meter technology limitations, this API definition is not promoted by ASME. Nevertheless, this API definition is common knowledge within the industry and as a review of wet gas flow meter technology, this Report needs to address it.

Three different ranges of the Lockhart–Martinelli parameter were identified with regard to wet gas metering systems by the API [2]. Figure E-1 shows an API map for classifying wet gas flows, on the basis of superficial velocities for gas and liquid and the Lockhart–Martinelli parameter. The three types of wet gas regions defined by API are shown on this map. The examples plotted on the map were for 200 bar and 10 bar (to indicate the gas-to-liquid density ratio effect).

E-2 API TYPE 1 WET GAS

Type 1 is seen as a wet gas flow where $X_{LM} \leq 0.02$ indicates a relatively small amount of liquid. This type is a typical liquid-to-gas flow ratio limit for which single-phase gas differential pressure meters are operated without malfunctioning or giving substantial errors that would require correlations to correct. That is, the range could cause a relatively small and possibly acceptable gas flow rate error. It is the responsibility of any meter user to judge whether the liquid-induced error is acceptable for their application. It should be noted however, that non-DP gas flow meters do not necessarily have the same reaction to wet gas flow as a DP meter and therefore the range may be less appropriate in other meter cases.

It is not common for wet gas flow meters (that is, a system that meters both gas and liquid flows) to be used for type 1 flows due to the limited amount of liquid present in the flow. The small quantity of liquid causes the liquid flow rate estimation to have a high uncertainty and in many (but not all) cases if the liquid is not a substantial quantity of the total flow there is no imperative to meter it.

It should be noted that the boundary between types of wet gas flow in Fig. E-1 is dependent on the composition of the liquid and the pressure (which affects the density of the gas). The dotted and solid lines in Fig. E-1 illustrate these effects when two alternative Type 1 boundaries are produced by a set gas and liquid at pressures of 150 psi (10 bar) and then 3,000 psi (200 bar). As the density of the gas changes with pressure the set Lockhart–Martinelli parameter line representing a constant 0.02 value shifts.

For the majority of industrial situations, Type 1 wet gas corresponds to a typical range of GVF ≥99.8%. This type of wet gas typically consists of less than 0.2% liquid by volume. The primary interest in this type of wet gas metering is often to measure the gas content of the flow and the liquid content of the flow is often of less importance, although accurate knowledge of the liquid content may be desirable to develop more accurate gas readings, especially in fiscal metering applications. There are cases however, where the liquid flow rate is desired to be known accurately as the primary production fluid.

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

E-3 API TYPE 2 WET GAS

Type 2 is seen by API [2] as wet gas flow in the range $0.02 < X_{LM} \le 0.3$. This type of wet gas has liquid flow rates higher than Type 1. API states this range was chosen as wet gas flow meters typically suit this range. It is a common range at (natural gas) production wellheads, commingled pipelines, and well testing applications. With these API Type 2 wet gas flows it is often required to meter both phase flow rates. Single phase flow meters can be used to meter these API Type 2 wet gas flow but in order to use these meters an understanding of the meter's response to liquid loading is usually required (i.e., a wet gas correlation) along with the ability to obtain liquid flow rates from an independent source. Otherwise a wet gas meter design is required (i.e., a system that meters simultaneously both gas and liquid flows).

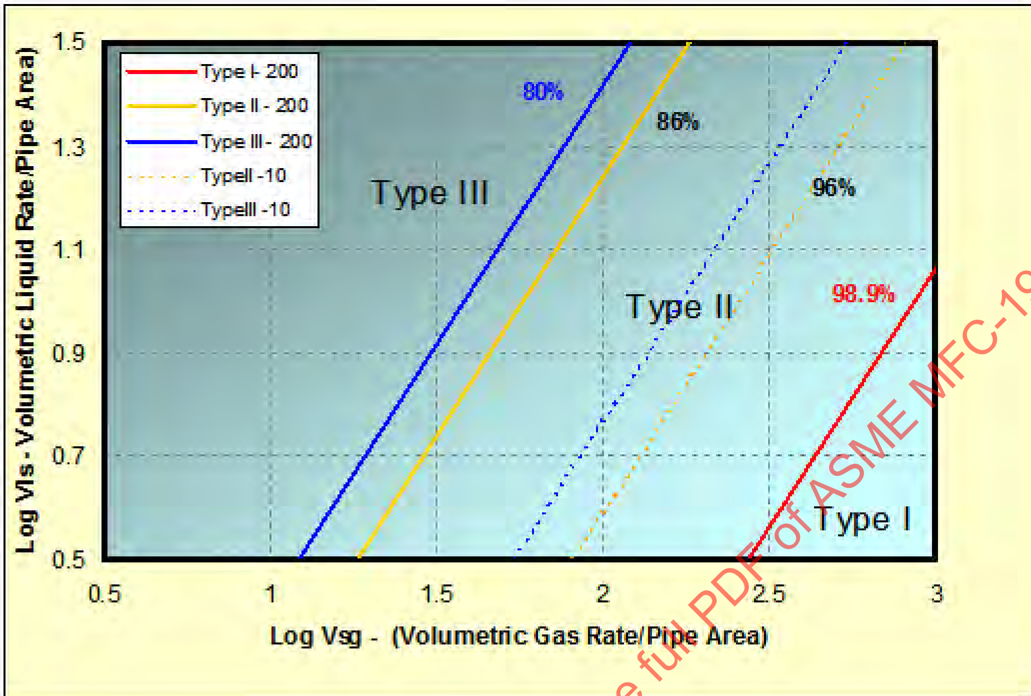
API Type 2 wet gas flow is defined as the region above API Type 1 in Fig. E-1 and constrained to the liquid content limited by $X_{LM} \le 0.3$. The position of the $X_{LM} \le 0.3$, boundary in Fig. E-1 is dependent for a given gas and liquid type on the pressure (i.e., it is dependent on the gas-to-liquid density ratio).

E-4 API TYPE 3 WET GAS

Type 3 is seen by API [2] as wet gas flow in the range $X_{LM} > 0.3$. API Type 3 is not considered in this Report to be wet gas flow. This region is beyond that which most single-phase gas meters (with wet gas correlations) or wet gas meters usually operate. API Type 3 is considered in this Report to be a general two-phase flow. Typically if a flow requires metering that is an API Type 3 flow a "multiphase meter" is required as there is too high a liquid loading for most wet gas metering technologies (although individual technologies have different stated maximum liquid loading limits).

It should be noted that the API Type 3 has no upper boundary of liquid content and is therefore not a precise definition. This is similar to the Norwegian Society for Oil and Gas Measurement's "Handbook of Multiphase Metering" [1] where they state "… generally wet gas is defined as gas/liquid systems with a Lockhart–Martinelli parameter smaller than approximately 0.3 in. The reason for the API and the Norwegian Society for Oil and Gas Measurement's imprecise definition is that, in the particular problem of oil and gas production measurement, the definition is aimed at the various wet gas and multiphase products where the limits of their working range can vary around the Lockhart–Martinelli parameter value of 0.3.

Fig. E-1 API Map for Classifying Wet Gas Streams
(Reproduced with the permission of the American Petroleum Institute. All rights reserved.)
Note: Three types of "wet gas" regions are defined on the basis of gas/liquid superficial velocity, GVF, and Lockhart–Martinelli parameter.



This Report is aimed at general industry and therefore mentions the API types for reference only. The wet gas definition in this Report is simply $X_{LM} \leq 0.3$. It is not technically possible to give direct equivalent limits in terms of other wet gas parameters such as GVF, LVF, flow quality (x), etc. as they are all related by the gas-to-liquid density ratio as indicated by eqs. (22) through (27) and the figures in Nonmandatory Appendix D. (As the definition of wet gas flow is a foundation on which this Report relies Nonmandatory Appendices C and D discuss this point in detail.)

NONMANDATORY APPENDIX F[1]
WET GAS FLOW CONDITION SAMPLE CALCULATIONS

F-1 GENERAL

Most wet gas meter manufacturers describe the specifications and performance of the meters in terms of the Lockhart–Martinelli parameter. Occasionally the gas volume flow rate, the quality, the Liquid volume flow rate, the liquid-to-gas mass flow rate, or the liquid-to-gas volume flow rate are also used. Most engineers in general industry have little experience with wet gas flows and describe wet gas (or general two phase) conditions in many different ways.

For the industries that deal most with wet gas flows, i.e., the oil and gas industry and the steam-related industries, there are however common ways of giving this information. In the oil and gas industry often flow condition specifications are expressed as gas (mass or volume) flow and liquid (mass or volume) flow. Either the gas and liquid flow densities are supplied or the pressure, temperature, and the gas molecular weight are offered with the types of liquid flowing. In the steam industry quality (i.e., dryness fraction) is usually quoted.

In order to examine the possible metering methods it is necessary to convert these data sets into the wet gas terms such as the Lockhart–Martinelli parameter and then the gas-to-liquid density ratio, gas densiometric Froude number, etc. to verify that any wet gas flow correlation for a particular meter will be directly applicable or require significant extrapolations.

The following are examples of real wet gas flow enquiries received by meter manufacturers and how the varied data sets supplied were then processed into the wet gas flow parameters required to categorize the flow in question. These examples are chosen to show typical common practical problems and allow comments on the different situations. All names of companies and locations of the flows have been removed.

F-2 EXAMPLE 1: A WET NATURAL GAS PRODUCTION FLOW WET GAS METERING INQUIRY

The information supplied to the meter manufacturer was

4-in. meter, extra thick wall, internal diameter, (ID) 2.884 in.
Gas Flow Rate: 7 MMSCFD (i.e., 7 million standard cubic feet per day)
Pressure: 650 psig (i.e., 45.8 bara)
Temperature: 70^0F (i.e., 294.3 K)
Gas Molecular Weight: 19.7 mW
Liquid Flow Rate Information:
Water: 5 barrels / MMSCFD at a density of 996 kg/m^3
Hydrocarbon Liquid: 10 barrels / MMSCFD at a density of 780 kg/m^3

The calculation procedure for wet gas flow condition analysis was:

At a pressure of 45.8 bara and a temperature of 294.3 K a natural gas with a molecular weight of 19.7 has an approximate density of 40.2 kg/m^3.

The density of the liquid mixture is the total combined liquid mass per unit volume occupied by the liquid phase. It is assumed that the two liquid components are fully mixed and a homogenous liquid phase is flowing. The density of this homogeneous liquid phase can be calculated in the following way:

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

$$\rho_{l_{homogenous}} = \frac{\rho_{l_1} \rho_{l_2}}{\rho_{l_2} x_l + \rho_{l_1} (1 - x_l)}$$

where

$$x_l = \frac{\dot{m}_{l_1}}{\dot{m}_{l_1} + \dot{m}_{l_2}}$$

(For proof of this statement see Nonmandatory Appendix J, para. J-3.)
therefore:

1 U.S. barrel ≡ 42 U.S. gallons ≡ 0.1589873 m^3

The volume flow rate of water is 5 barrels/MMSCFD. Therefore volume flow rate of water is 0.7949 m^3/MMSCFD. Therefore for a gas flow rate of 7 MMSCFD the volume flow rate of water is 5.564 m^3/day. Therefore, with a water density of 996 kg/m^3 the water mass flow rate is 5 542.3 kg /day.

The volume flow rate of liquid hydrocarbon is 10 barrels / MMSCFD. Therefore volume flow rate of liquid hydrocarbon is 1.589873 m^3/MMSCFD. Therefore, for a gas flow rate of 7 MMSCFD the volume flow rate of liquid hydrocarbon is 11.129 m^3/day. Thus, with a liquid hydrocarbon density of 780 kg/m^3 the liquid hydrocarbon mass flow rate is 8 680.7 kg /day.

therefore:

$$x_l = \frac{\dot{m}_{l_{water}}}{\dot{m}_{l_{water}} + \dot{m}_{l_{hydrocarbon}}} = \frac{5542.3}{5542.3 + 8680.7} = 0.39$$

so:

$$\rho_{l_{homogenous}} = \frac{\rho_{l_{water}} \rho_{l_{hydrocarbon}}}{\rho_{l_{hydrocarbon}} x_l + \rho_{l_{water}} (1 - x_l)} = \frac{996 * 780}{(780 * 0.39) + (996 * (1 - 0.39))} = 852.1 \; kg / m^3$$

Hence, the total liquid flow rate is 5 542.3 kg/day of water and 8 680.7 kg/day of liquid, which is a total liquid flow of 14 223 kg /day. (Also note that the total volume flow is 15 barrels/MMSCFD, i.e., 2.385 m^3/MMSCFD. At a gas flow rate of 7 MMSCFD the total liquid volume flow rate is 16.695 m^3/day. At a homogeneous density of 852.1 kg/m^3 this is a total liquid mass flow rate of 14 224 kg/day. The difference of 1 kg is rounding errors).

Therefore, with a gas density of 40.2 kg/m^3 the gas-to-liquid density ratio is

$$\frac{\rho_g}{\rho_{l_{homogenous}}} = \frac{40.2}{852.1} = 0.0472$$

At the flow conditions 7 MMSCFD is 165 160 kg/day.

Therefore, Lockhart–Martinelli Parameter is

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} = \frac{14223}{165160} \sqrt{\frac{40.2}{852.1}} = 0.0187$$

The superficial gas velocity (i.e., the average velocity of the gas if the gas phase flowed alone in the pipe) is

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} = \frac{1.91}{40.2 * 4.2145e-3} = 11.3 \ m/s$$

where

$$\dot{m}_g = 165160 \ kg/day = 1.91 \ kg/s$$

and

$$A = \frac{\pi}{4} D^2 = \frac{\pi}{4} (0.07325)^2 = 4.2145e-3 \ m^2$$

as inlet diameter D = 2.884 in. = 0.07325 m.

Therefore, the gas densiometeric Froude number is

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}} \sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{11.3}{\sqrt{9.81 * 0.07325}} \sqrt{\frac{40.2}{852.1 - 40.2}} = 2.97$$

Therefore, the wet gas flow can be defined in the following way:
Natural Gas / Liquid Hydrocarbon / Water Flow
Meter inlet area: 2.884 in./0.07325 m
Lockhart–Martinelli parameter: 0.0187
Gas-to-liquid density ratio: 0.0472
Gas densiometric Froude number: 2.97

The other common wet gas parameters are not required but for completeness are calculated below:

from eq. (23)

$$GVF = \frac{\sqrt{\frac{\rho_l}{\rho_g}}}{X_{LM} + \sqrt{\frac{\rho_l}{\rho_g}}} = \frac{\sqrt{\frac{852.1}{40.2}}}{0.0187 + \sqrt{\frac{852.1}{40.2}}} = 0.996$$

from eq. (24)

$$x = \frac{1}{1 + X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1 + \left(0.0187 * \sqrt{\frac{8521}{40.2}}\right)} = 0.921$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = X_{LM}\sqrt{\frac{\rho_l}{\rho_g}} = 0.0187 * \sqrt{\frac{852.1}{40.2}} = 0.0861$$

from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = X_{LM}\sqrt{\frac{\rho_g}{\rho_l}} = 0.0187 * \sqrt{\frac{40.2}{852.1}} = 0.0041$$

from eq. (27)

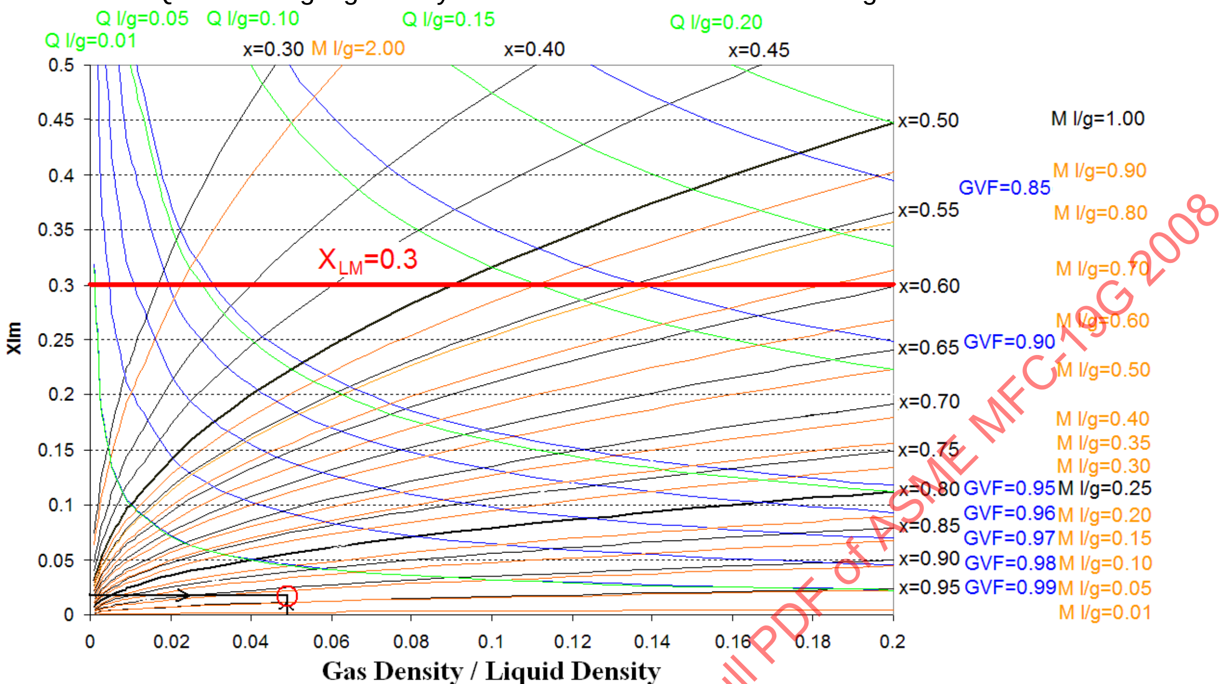
$$LVF = 1 - (GVF) = 1 - 0.996 = 0.004$$

F-2.1 Comments on Example 1

This method of describing gas flow in terms of “MMSCFD” (i.e., millions of cubic feet if the gas phase flowed at standard atmospheric conditions) and liquid flow in terms of barrels per MMSCFD is very common in the oil and gas industry. This information must come with the actual flow conditions in terms of pressure and temperature as well as the gas and liquid properties as this information must be converted to find the actual flow conditions the meter will experience. The change from standard to actual gas volume flow rates should be done by a reputable fluid property calculation method. This is out with the scope of this document and in the above example the actual gas density and mass flow rate is simply stated as known from the given information. From this common method of describing the two-phase flow no further analysis can be done until the actual fluid property calculations are complete. For this type of case, only when this is done, can the two-phase flow parameter graphs be utilized to show a quick graphical review of the flow condition. With the calculation of the gas-to-liquid density ratio and the Lockhart–Martinelli parameter these graphs could be used to show predictions of the other wet gas parameters without having to apply eqs. (29), (31), (32), (42), and (44). Figure D-11 is reproduced here with this gas-to-liquid density ratio to Lockhart–Martinelli parameter indicated by a red circle in Fig. F-1. It is seen that by following the constant parameter lines all the predictions from eqs. (29), (31), (32), (42), and (44) can be approximated at a glance from this graph.

It is common for wet natural gas flows to have water and liquid hydrocarbon flows quoted. At the time of writing there is little information in the public domain on the effect of different liquid properties on wet gas flows. The only references known to us are some of the most recent technical papers at the time of writing and they show that varying liquid properties can have a moderate effect on a DP meter wet gas response (i.e., Reader Harris et al. [17,18] and Steven et al. [19,20]). However, currently when calculating the wet gas flow conditions with more than one liquid component industry tends to assume full mixing of these liquids and an average of the liquid densities.

Figure F-1 Magnified Graph of Wet Gas Flow Liquid-to-Gas Flow Ratio Parameter Relationships With the Point in Question Highlighted by a Red Circle to Allow the Reading Off of All Wet Gas Parameters



The Lockhart–Martinelli parameter (X_{LM}) of 0.0187 is very low. This flow is considered a wet gas flow rather than a general two-phase flow as $X_{LM} \leq 0.3$. However, it is also $X_{LM} \leq 0.02$, and some engineers claim that this level of liquid loading is small enough that meter users could consider using single-phase gas meter technologies with no wet gas correction ability, to get the gas flow reading with only a small increase in uncertainty caused by the liquid presence. It is often thought to be not worthwhile correcting any liquid-induced error at very low Lockhart–Martinelli parameter levels as this error will be no more than the correction uncertainty. In actual practice users decide from the application requirements and knowledge of the wet gas meter performance of the particular meter being considered, if a correction is beneficial or whether the additional uncertainty introduced by the wet gas is acceptable.

The meter size (at 2.884 in. inlet diameter) and the flow conditions of a Lockhart–Martinelli parameter of 0.0187, a gas-to-liquid density ratio of 0.0472 and a gas densiometric Froude number of 2.97 are all well within the typical wet gas test facility's capabilities. The only problem to an engineer in this case is that the liquid is a mix of water and hydrocarbon liquid and with the latest research suggesting that the liquid properties can, under some conditions at least, influence the response of meters to a wet gas flow, a decision will have to be made on whether to test at one liquid property or insist on testing with the expected liquid mix. This will show any liquid property effect but there are, at the time of writing and for the foreseeable future, few test facilities that can test gas/hydrocarbon liquid/water flows at or close to real natural gas production conditions. Testing wet gas flows with various liquid mixes is time consuming and expensive. Finally, note that even for many real wet gas flows with gas and single component liquids there is often one or more parameter out with the test matrix of most or all wet gas flow test facilities (e.g., meter diameter, pressure, flow rate, etc.) and extrapolation of the test results is an unavoidable result. The fact that there is currently no easy answer to this dilemma is indicative of the fact that this technology is still in its infancy.

F-3 EXAMPLE 2: WET NATURAL GAS PRODUCTION METERING INQUIRY

The following information was supplied to a meter manufacturer by an oil company inquiring about the metering of a 6-in. Schedule 160 two-phase flow that was stated to be a wet gas flow.

Pressure 166 bara
Temperature 25°C
Gas Density 212.7 kg/m^3
Gas Flow Rate: 26 750 kg/hr
Liquid Density: 1 005 kg/m^3
Liquid Flow Rate: 69 008.3 kg/hr

This data set allowed the direct calculation of the gas-to-liquid density ratio and the Lockhart–Martinelli parameter. That is:

$$\frac{\rho_g}{\rho_l} = \frac{212.7}{1005} = 0.212$$

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{69{,}008}{26{,}750}\sqrt{\frac{212.7}{1005}} = 1.187$$

Clearly here this is not a wet gas flow but a general two-phase flow as X_{LM} >>0.3. Wet gas technologies are not appropriate for such flows, and a multiphase meter will need to be used for this flow condition.

The other common wet gas parameters are not required but for completeness are calculated below:

from eq. (23)

$$GVF = \frac{\sqrt{\frac{\rho_l}{\rho_g}}}{X_{LM} + \sqrt{\frac{\rho_l}{\rho_g}}} = \frac{\sqrt{\frac{1005}{212.7}}}{1.187 + \sqrt{\frac{1005}{212.7}}} = 0.645$$

from eq. (24)

$$x = \frac{1}{1 + X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1 + \left(1.187 * \sqrt{\frac{1005}{2127}}\right)} = 0.279$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = X_{LM}\sqrt{\frac{\rho_l}{\rho_g}} = 1.187 * \sqrt{\frac{1005}{212.7}} = 2.59$$

from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = X_{LM}\sqrt{\frac{\rho_g}{\rho_l}} = 1.187 * \sqrt{\frac{212.7}{1005}} = 0.546$$

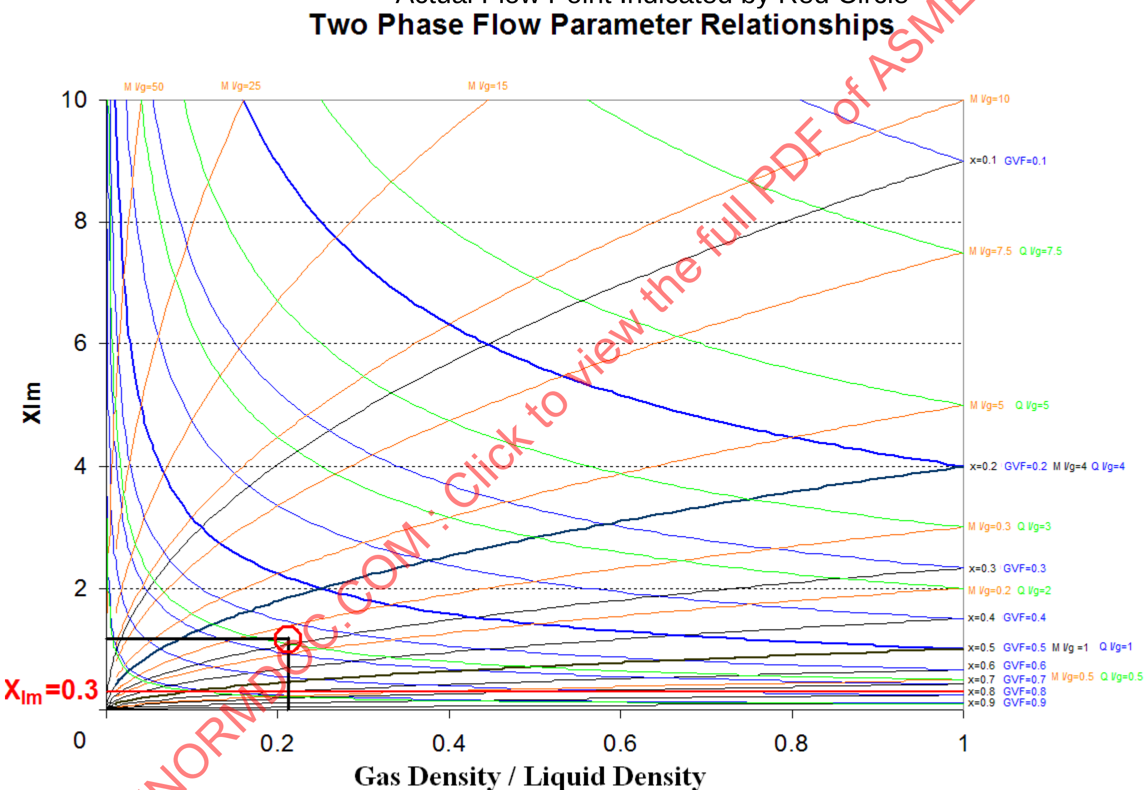
from eq. (27)

$$LVF = 1 - (GVF) = 1 - 0.647 = 0.353$$

F-3.1 Comment on Example 2

This example shows that it is not always obvious to engineers what flow conditions are wet gas flows and which are general two-phase flows without calculating the wet gas parameters. This two-phase condition is not close to being categorized as a wet gas flow (as $X_{LM} >> 0.3$) and yet the meter manufacturer was asked if a wet gas meter was appropriate. Use of the two-phase flow parameter graphs should help users judge if they need a multiphase or wet gas meter technology. Figure F-2 shows the general two-phase flow parameter map (i.e., Fig. D-8) with the actual flow condition being indicated by the center of the red circle. It is seen in Fig. F-2 that the GVF is approximately 0.65 [actual GVF value calculated by application of eq. (23) is 0.645] and hence in this two-phase flow there is more gas by volume than liquid. This can cause confusion and lead some engineers to call this wet gas flow. It is not according to the definition in this Report. The term wet gas flow strictly means flows with $X_{LM} \leq 0.3$.

Fig. F-2 General Two-Phase Flow Parameter Map With Actual Flow Point Indicated by Red Circle



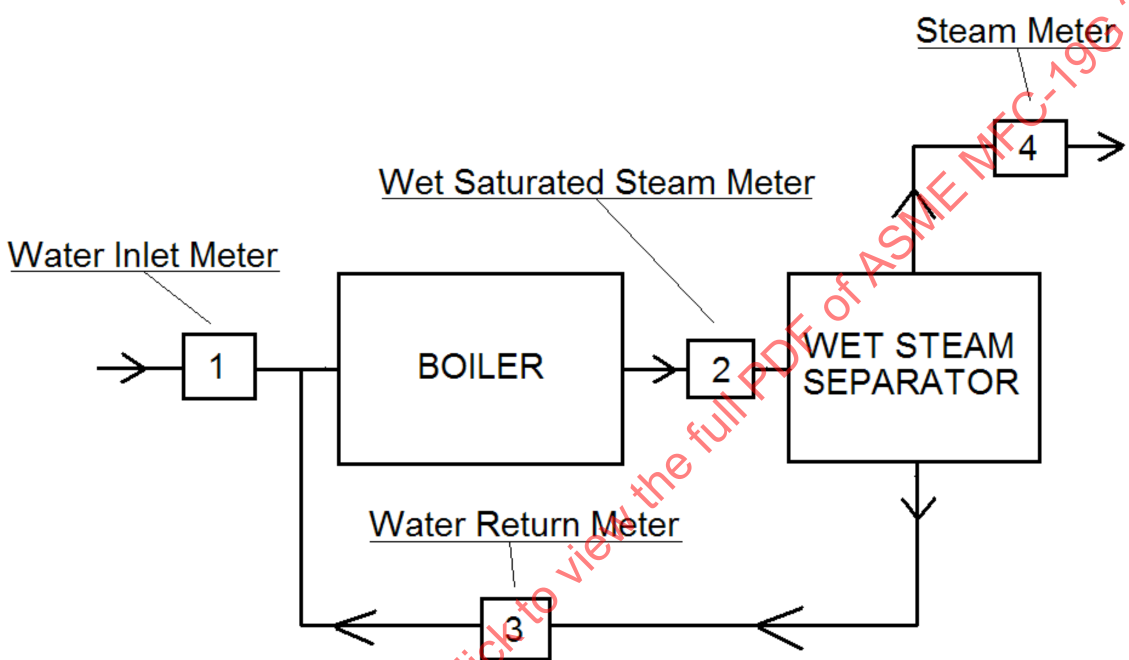
A final comment on example 2 is it should be noted that the pressure is so high that the gas-to-liquid density ratio is well in excess of the available wet gas test facilities. Therefore, for wet gas flow conditions with such high gas-to-liquid density ratios, caution should be taken when choosing a meter that has wet gas correlations based on data sets with significantly lower maximum gas-to-liquid density ratios. The user should check the correlation does not diverge with this gas-to-liquid density ratio extrapolation. If in doubt the user should seek the advice of the meter manufacturer.

F-4 EXAMPLE 3: A STEAM QUALITY CHECK INQUIRY

F-4.1 Explanation

A steam flow is injected into an oil well to reduce the oil viscosity and therefore increase oil production. The steam was produced by an on-site boiler heating a water supply. The boiler was rated to produce a quality of 65% for the given flow rate. The two-phase flow out of the boiler was separated by a separator vessel and the water phase was returned to the boiler inlet pipe and the steam was injected into the oil well (see Fig. F-3). The system designers inquired to a meter manufacturer if it was possible to confirm the quality of the wet saturated steam exiting the boiler and if it was possible to check if the steam exiting the separator was dry steam.

Fig. F-3 Steam Injection System



The following is a description of the calculation procedure. Four possible metering points are shown numbered in the schematic diagram. Analysis:

The flow through meter one is water ($\dot{m}_{l1}$.). This can be metered as a single-phase flow with low uncertainty. The flow through meter two is steam and water mix ($\dot{m}_{tp2}$). The flow through meter three is water ($\dot{m}_{l3}$). This can be metered as a single-phase flow with low uncertainty. The flow through meter four is either single-phase steam or a flow of steam and water ($\dot{m}_4$).

For steady state flow and from mass continuity:

$$\dot{m}_{l1} + \dot{m}_{l3} = \dot{m}_{tp2} \tag{F-1}$$

and

$$\dot{m}_4 = \dot{m}_{tp2} - \dot{m}_{l3} \tag{F-2}$$

Therefore, it can be seen by substitution of eq. (F-1) into eq. (F-2) that we get the obvious steady state result:

$$\dot{m}_{l1} = \dot{m}_4$$

As required by mass continuity for a steady state system.

With the water inlet flow and the returning liquid (water) flow from the separator being metered the total mass flow rate through the boiler is therefore known if not the steam to water mass ratio. Knowing the total mass flow rate allows a DP meter to give a steam quality/dryness fraction prediction. Note most DP meter wet gas performances have been quoted to be functions of the Lockhart–Martinelli parameter, gas to density ratio and gas densiometric Froude number. The gas flow rate predictions are in the form:

$$\dot{m}_g = \frac{\dot{m}_{uncorrected\ gas\ reading}}{f\left(X_{LM}, \frac{\rho_g}{\rho_l}, Fr_g\right)} \quad \text{(F-3)}$$

However, note that:

$$x = \frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g} \quad \text{(16)}$$

that is:

$$\dot{m}_g = x\left(\dot{m}_l + \dot{m}_g\right) = x\,\dot{m}_{total} \quad \text{(F-4)}$$

and

$$X_{LM} = \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} \quad \text{(22)}$$

and from eq. (7) and mass continuity [eq. (F-5)]:

$$\dot{m}_g = \rho_g A U_{sg} \quad \text{(F-5)}$$

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{\dot{m}_g}{A\sqrt{gD}}\sqrt{\frac{1}{\rho_g\left(\rho_l - \rho_g\right)}} = \frac{x\,\dot{m}_{total}}{A\sqrt{gD}}\sqrt{\frac{1}{\rho_g\left(\rho_l - \rho_g\right)}} \quad \text{(F-6)}$$

Therefore, in this case, as the total mass flow through the meter ($\dot{m}_{total}$) is known (i.e., $\dot{m}_{tp2}$ in this examples terminology) due to eq. (F-1) (with $\dot{m}_{l1}$ and $\dot{m}_{l3}$ being known from single-phase meters), the only unknown in eq. (F-3) is the steam quality/dryness fraction (x). That is:

$$x\left(\dot{m}_{l1} + \dot{m}_{l3}\right) = \frac{\dot{m}_{uncorrected\ gas\ reading}}{f\left(\frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}}, \frac{\rho_g}{\rho_l}, \frac{x\left(\dot{m}_{l1} + \dot{m}_{l3}\right)}{A\sqrt{gD}}\sqrt{\frac{1}{\rho_g\left(\rho_l - \rho_g\right)}}\right)} \quad \text{(F-7)}$$

Depending on the correlation chosen the quality term may be separated out to allow a direct calculation, or, if this is not algebraically possible, the quality term could be found by an iterative procedure. The uncertainty of this prediction is wholly dependent on the uncertainties of the inputs and the correlation chosen.

In this worked example the given flow conditions are

4 in. schedule 80, i.e., 3.826 in. (i.e., 0.09718 m)
Flow assumed to be steady.
Pressure: 90 bara (saturation temperature is therefore 578.45 K).

Total water flow rate supplied ($\dot{m}_{l1}$): 33 000 kg/hr

Liquid return from separator ($\dot{m}_{l3}$): 19 250 kg/hr

Uncorrected orifice plate meter steam flow rate reading downstream of the exit of the boiler,

$\dot{m}_{g2_{apparent}} = 37\,919 \;\; kg/hr$.

Uncorrected orifice plate meter steam flow rate reading for steam injection line,

$\dot{m}_{g4_{apparent}} = 31\,883 \;\; kg/hr$.

The orifice plate meters are applied with the simple two-phase orifice plate meters' Chisholm equation.

The total mass through the boiler ($\dot{m}_{tp2}$) is found with eq. (F-1)

$$\dot{m}_{l1} + \dot{m}_{l3} = \dot{m}_{tp2} = 33{,}000 + 19{,}250 = 52{,}250 \;\; kg/hr$$

The steam tables state that at a saturation condition of 90 bara and 578.45 K the steam and water densities are

$$\rho_g = 48.8 \;\; kg/m^3 \text{ and } \rho_l = 703.8 \;\; kg/m^3$$

The gas-to-liquid density ratio is therefore 0.069. The orifice plate at the outlet to the boiler is reading 37 919 kg/hr. That is, the orifice plate meter that is set up to read a single-phase steam flow rate and is actually experiencing a two-phase flow of total mass 52 252 kg/hr, is reading 37 919 kg/hr of steam. This value is denoted here as $\dot{m}_{g2_{apparent}}$. Therefore: $\dot{m}_{g2_{apparent}} = 37{,}919 \;\; kg/hr$.

The Chisholm equation (see Nonmandatory Appendices G and H) can be written as:

$$x\dot{m}_{total} = \dot{m}_{gas} = \frac{\dot{m}_{gas\ apparent}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} = \frac{\dot{m}_{gas\ apparent}}{\sqrt{1 + \left(C\left(\frac{1-x}{x}\right)\sqrt{\frac{\rho_g}{\rho_l}}\right) + \left(\left(\frac{1-x}{x}\right)\sqrt{\frac{\rho_g}{\rho_l}}\right)^2}} \qquad \text{(G-11)}$$

where

$$C = \left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} + \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} \qquad \text{(G-12)}$$

This equation can be solved for the quality of the two-phase steam exiting the boiler (x_2) by iteration of eq. (G-13). That is:

$$x_2\left(\dot{m}_{l1}+\dot{m}_{l3}\right)-\frac{\left(\dot{m}_{gas\ apparent}\right)_2}{\sqrt{1+\left(C\left(\frac{1-x_2}{x_2}\right)\sqrt{\frac{\rho_g}{\rho_l}}\right)+\left(\left(\frac{1-x_2}{x_2}\right)\sqrt{\frac{\rho_g}{\rho_l}}\right)^2}}=0 \qquad \text{(F-8)}$$

where $\left(\dot{m}_{gas\ apparent}\right)_2$ is the uncorrected gas (i.e., steam) flow rate from meter 2 at the exit of the boiler and C is found to be 2.46 from eq. (G-12). Therefore, if the Chisholm equation is correct the quality is 0.6 (i.e., 60%). This indicates that the boiler is not producing as high a quality steam as had been assumed by the system operators.

The steam produced by the boiler ($\dot{m}_{2_g}$) is:

$$\dot{m}_{2_g} = x_2 m_{tp_2} = 0.6 * 52{,}250 = 31{,}350 \ \ kg/hr$$

And the quantity of water out of the boiler ($\dot{m}_{2_l}$) is:

$$\dot{m}_{2_l} = (1-x_2)m_{tp_2} = (1-0.6)*52{,}250 = 20{,}900 \ kg/hr$$

It will be noted that the liquid (or water) flow prediction at the boiler outlet by the Chisholm equation is 20 900 kg/hr and the meter reading on the recirculating line (meter 3) is quoted as 19 250 kg/hr. This indicates that the separator is not 100% efficient and the flow being injected is not dry steam but a wet steam. (The observant may have noticed at the start of this example that the inlet flow rate read by meter 1 is not the same as meter 4, which is reading the system's outlet flow. If the flow is steady this indicates that one or more of the meters are not metering the flow correctly. In this situation an obvious reason is that the outlet flow is wet steam flow and not the assumed single-phase steam flow.) From continuity and an assumption of steady state flow the mass flow through the steam meter is known to be the same as that entering the system, i.e., 33 000 kg/hr. Therefore, the total mass flow being metered by meter 4 must be 33 000 kg/hr. The steam being produced in the boiler, flown through meter 4 and injected into the well is found to be 31 350 kg/hr. (This statement assumes that the system is well lagged and there are short distances between the components thereby assuring no significant phase change from changes in thermodynamic conditions between meter 2 and meter 4). The difference between the liquid return flow and the calculated liquid flow out of the boiler is therefore 20 900 – 19 250 = 1 650 kg/hr. This suggests an injection steam quality (x_4) of:

$$x_4 = \frac{33{,}000-1{,}650}{33{,}000} = \frac{32{,}340}{33{,}000} = 0.95$$

i.e., a 95% steam quality

This can be checked by applying the Chisholm equation to meter 4 in the steam injection line. Assuming no significant change of pressure or temperature between the boiler outlet and meter 4 the phase densities are as before. The single-phase steam meter reads an uncorrected steam flow

($\left(\dot{m}_{gas\ apparent}\right)_4$) of 31 883 kg/hr. Equation (F-9) is the Chisholm equation arranged for the steam quality iteration at meter 4. Note from mass continuity the total mass flowing through meter 4 (and therefore out of the system) is the same as the mass through meter 1 (and into the system), $\dot{m}_{l1}$.

$$x_4\left(\dot{m}_{l1}\right)-\frac{\left(\dot{m}_{gas\ apparent}\right)_4}{\sqrt{1+\left(C\left(\frac{1-x_4}{x_4}\right)\sqrt{\frac{\rho_g}{\rho_l}}\right)+\left(\left(\frac{1-x_4}{x_4}\right)\sqrt{\frac{\rho_g}{\rho_l}}\right)^2}}=0 \qquad \text{(F-10)}$$

Therefore, this check confirms that the injection steam is not dry and the injection quality is 95%.

F-4.2 Comments on Example 3

This example shows that it is not just the unprocessed natural gas production lines that require wet gas metering. Other applications and other industries (especially all industries using steam often face the challenge of metering wet gas flows).

It is often not realized that for a known total mass flow (such as inlet water to a boiler) wet gas correlations such as Chisholm's orifice plate two-phase correlation can be used to derive a prediction of the flow quality/dryness fraction. This flow quality/dryness fraction prediction uncertainty is directly dependent on the applicability of the wet gas correlation chosen. As stated in example one there is little information available on the effects of different liquid properties on wet gas meters. The early research (1950s to 1970s) into two-phase/wet gas metering was largely concentrated on steam flows. By the late 1970s the steam industry research had ceased. After two decades of research inactivity the natural gas production industry started research into wet gas metering in the 1990s and used the archive steam research as the foundation for further developments. However, whereas originally the natural gas production industry borrowed ideas and developments from the steam industries, now the situation has reversed and the steam industry has a renewed interest in wet gas metering due to the new technologies being developed by the natural gas production industry offering the potential for an increase in efficiency and cost cutting in steam flow industrial applications. However, the trouble with this situation is nobody is sure of how applicable wet steam data is to natural gas/water/liquid hydrocarbon flows and vice versa. Direct checking is hugely hindered by the fact that virtually all wet steam laboratories with the ability to hold the reference values to a low uncertainty have been decommissioned over the last three decades and so all that is available are the old wet steam data sets and correlations. It is an open question how much confidence should be given to decades-old wet steam data that is not traceable.

The problem today is a small but still significant amount of natural gas and liquid hydrocarbon data and correlations are available, but with no wet steam flow laboratories with good reference data to check them with, it is not known how representative these correlations would be if applied to a wet steam flow. Currently, due to lack of alternatives, industry applies either the old steam correlations or the new natural gas and liquid hydrocarbon correlations depending on the industry, company, and individual engineer. The above example utilized the Chisholm correlation. In each unique application the user could possibly chose to use a different DP meter type and a different correction factor. This problem highlights the need of a modern wet steam test facility that could conduct research to tackle this problem.

For practice the conditions at the boiler exit and steam injection meters are plotted on the wet gas plot to predict the other wet gas parameters [see Fig. (F-4)]. As before the flow condition points are highlighted at the center of the red circles. The calculations for the boiler exit meter are also shown below. Again, it should be noted that the calculation and graph results match each other. The flows are seen to be well within the wet gas flow region. It is noteworthy that Chisholm [8] states the orifice plate wet gas correction eq. (G-11) is formed for X_{LM} <1 and the majority of the data was for 2.5-in. to 4-in. flows. Therefore, the Chisholm equation is found to be a reasonable choice of correlation in this case.

from eq. (22)

$$X_{LM} = \frac{1-0.6}{0.6}\sqrt{\frac{48.8}{703.8}} = 0.176$$

from eq. (23)

$$GVF = \frac{\left(\frac{703.8}{48.8}\right)}{\left(\frac{703.8}{48.8}\right)+\left(\frac{1-0.6}{0.6}\right)} = 0.956$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = \frac{1-0.6}{0.6} = 0.667$$

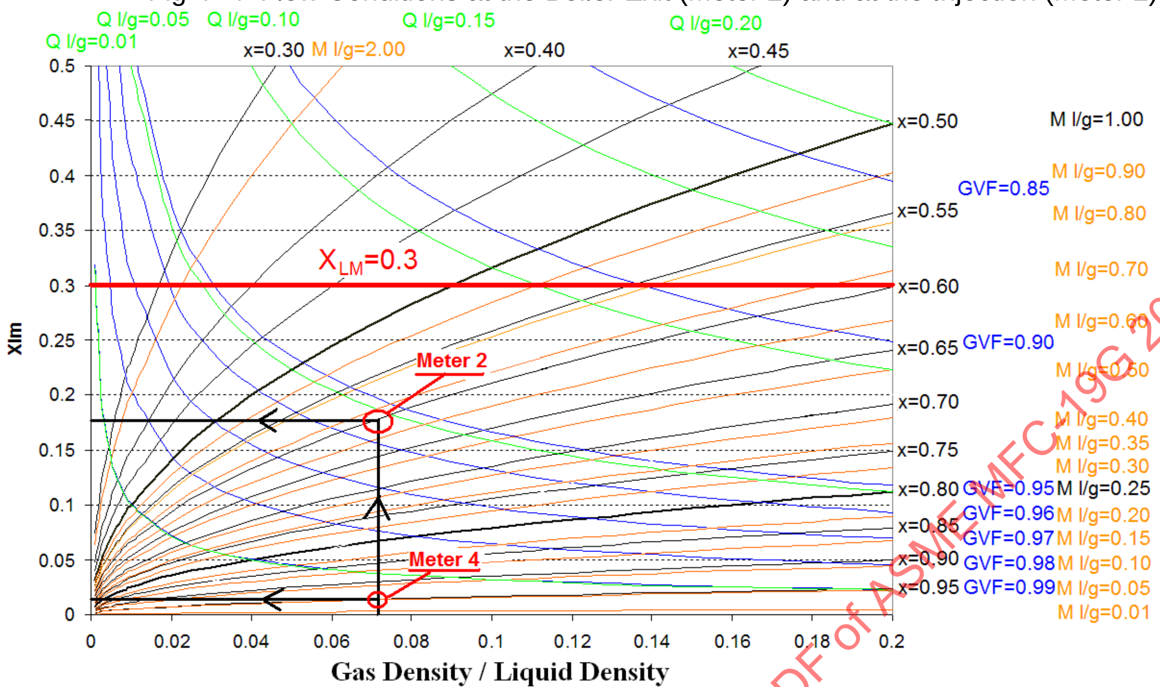
from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = \left(\frac{1-0.6}{0.6}\right)\frac{48.8}{703.8} = 0.0462$$

from eq. (27)

$$LVF = \frac{\left(\frac{1-0.6}{0.6}\right)}{\left(\frac{1-0.6}{0.6}\right)+\frac{703.8}{48.8}} = 1-0.956 = 0.044$$

Fig. F-4 Flow Conditions at the Boiler Exit (Meter 2) and at the Injection (Meter 2)



F-5 EXAMPLE 4: PRESENTATION OF FLOW DATA TO METER MANUFACTURERS AND THE DANGER OF SATURATION OF DP TRANSMITTERS

Many pieces of information are typically supplied to metering engineers that are irrelevant to the wet gas flow calculations. However, there are certain pieces of information that must be supplied for wet gas flow metering analysis to be possible. It is also stated by meter manufacturers that there are two common problems related to the supply of data that often occur due to lack of wet gas flow knowledge from the meter user. These problems are shown in this example. The relevant data, taken from a long list of data describing a proposed offshore platform wet natural gas flow handed to the meter engineer, was as follows:

Line Size: 6 in. Schedule 80 pipe work, i.e., 5.761 in./0.1463 m
Gas Mass Flow: Minimum 6,950 kg/hr, Maximum 69 500 kg/hr
Liquid Mass Flow: Minimum 3,125 kg/hr, Maximum 31 250 kg/hr
Pressure steady at 100 bara
Gas Density: 82.4 kg/m^3
Liquid Density: 864 kg/m^3

This was an unusually detailed data set and reduces the work of the engineer analyzing the flow by giving the mass flows and densities directly. Also, the liquid was quoted as being a mix of water and condensate but the average density was given. However, this data set also shows a typical ambiguity in terms of gas and liquid mass flow rate ranges. A common trait in oil and gas production data sets is seen here. The data set gives maximum and minimum mass flow rates but does not declare if the gas and liquid maximum and gas and liquid minimum mass flow rates are pairs or if the maximum of one phase may flow with the minimum of the other. It is not always obvious to the meter engineer which combinations go together. One common situation is that the maximum gas flow rate is at maximum production and the high flow rate draws up the maximum liquid flow with it. Due to production demands the flow may be throttled back to lower gas velocities and the liquid drawn out with it is thereby also reduced. Hence, in this case the maximum gas and liquid flow rates and the minimum gas and liquid flow rates are pairs. However, it is also common for natural gas production companies to inquire about wet gas meters by quoting the expected maximum and minimum of phase flow rates the meter will see in the life of the well. Here, it is common for wells to start life producing a relatively dry gas (i.e., maximum gas flow rate and minimum liquid flow rate). As the well ages the gas flow reduces and the water being drawn increases while pressure in the well drops causing the heavier hydrocarbon gases to convert to light hydrocarbon liquid. Therefore, the minimum gas flow rate and maximum liquid flow rate of the well occur concurrently. Another common situation is for an existing pipeline with one wet gas flow to be in service when a later developed reservoir production flow is "tied back" to this older pipeline and the flows commingled. In this case with out extra information regarding the other wet gas flow it is not possible to derive which of the stated gas and liquid flow rates are pairs. Hence, unless stated with the data set the wet gas metering analyst does not know which gas and liquid flow rates are pairs." As it is not the meter manufacturer's responsibility or position to know what the flow conditions are predicted to be it is the responsibility of the meter user to state this. Currently, this is often not done and the analysis has to take account of all possible conditions. This can lead to hugely inaccurate wet gas flow condition predictions. The following example shows this:

The actual conditions were found later to be that the maximum phase flow rates and minimum phase flow rates are pairs. (That is, maximum gas and liquid flow rates and minimum gas and liquid flow rates were together.) The actual expected Lockhart–Martinelli parameters were therefore:

Maximum Flow Conditions: $X_{LM} = \frac{31{,}250}{69{,}500}\sqrt{\frac{82.4}{864}} = 0.139$

Minimum Flow Conditions: $X_{LM} = \frac{3{,}125}{6{,}950}\sqrt{\frac{82.4}{864}} = 0.139$

However, if the maximum gas and minimum liquid flow rates were paired:

$$X_{LM} = \frac{3{,}125}{69{,}500}\sqrt{\frac{82.4}{864}} = 0.0139$$

If the maximum liquid and minimum gas flow rates were paired:

$$X_{LM} = \frac{31{,}250}{6{,}950}\sqrt{\frac{82.4}{864}} = 1.389$$

Hence, the actual conditions expected are in the middle of the wet gas flow range where wet gas meters are applicable. If incorrect (but reasonable) assumptions were made an incorrect Lockhart–Martinelli parameter range of 0.0139 (i.e., almost dry — where a single-phase meter could be used) to 1.389 (i.e., where no wet gas meter will work and a multiphase meter is required) would have been predicted.

A potentially costly error could be made in these types of cases. Such errors could lead to a flow being labeled multiphase (with the added complexity and cost of multiphase systems then coming into play on marginal field propositions) or labeled a wet gas where the wet gas systems being specified will in fact not work in a flow that is actually multiphase.

In this case the actual Lockhart–Martinelli parameter is 0.139 and the gas-to-liquid density, superficial velocity, and gas densiometeric Froude number were found:

$$\frac{\rho_g}{\rho_l} = \frac{82.3}{864} = 0.0952$$

$$\bar{U}_{sg\,\max} = \frac{\dot{m}_g}{\rho_g A} = \frac{\left(\frac{69{,}500}{3600}\right)}{82.3 * 1.681e-2} = 13.95 \ m/s$$

where

$$A = \frac{\pi}{4} D^2 = \frac{\pi}{4}(0.1463)^2 = 1.681e-2 \ m^2$$

Therefore the maximum gas densiometeric Froude number is

$$Fr_{g_{\max}} = \frac{U_{sg}}{\sqrt{gD}} \sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{13.95}{\sqrt{9.81 * 0.1463}} \sqrt{\frac{82.3}{864 - 82.3}} = 3.77$$

and for the minimum flow case:

$$\bar{U}_{sg\,\min} = \frac{\dot{m}_g}{\rho_g A} = \frac{\left(\frac{6{,}950}{3600}\right)}{82.3 * 1.681e-2} = 1.39 \ m/s$$

$$Fr_{g_{\min}} = \frac{U_{sg}}{\sqrt{gD}} \sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{1.39}{\sqrt{9.81 * 0.1463}} \sqrt{\frac{82.3}{864 - 82.3}} = 0.377$$

All these parameters were within the wet gas test loop ranges and hence the wet gas correlations were considered trustworthy.

In this case a DP meter was chosen. If the gas flowed alone the maximum gas flow was predicted to give a DP reading (ΔP_g) of 27.1 kPa (i.e., approximately 109 in. WC). This was the maximum DP the user expected from their preliminary calculations. A DP transmitter of 0-125 in. WC was specified. However, the actual DP reading (ΔP_{tp}) was in fact much higher as they had not accounted for the liquid effect. The liquid loading of $X_{LM} = 0.139$ gave an overreading of approximately 20%. For all DP meters, the flow rate prediction is directly related to the square root of the read differential pressure. The actual ΔP_{tp} should have been predicted by the following equation:

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = 1.2$$

where ΔP_g is the predicted differential pressure when the predicted gas flow rate flows alone. Therefore:

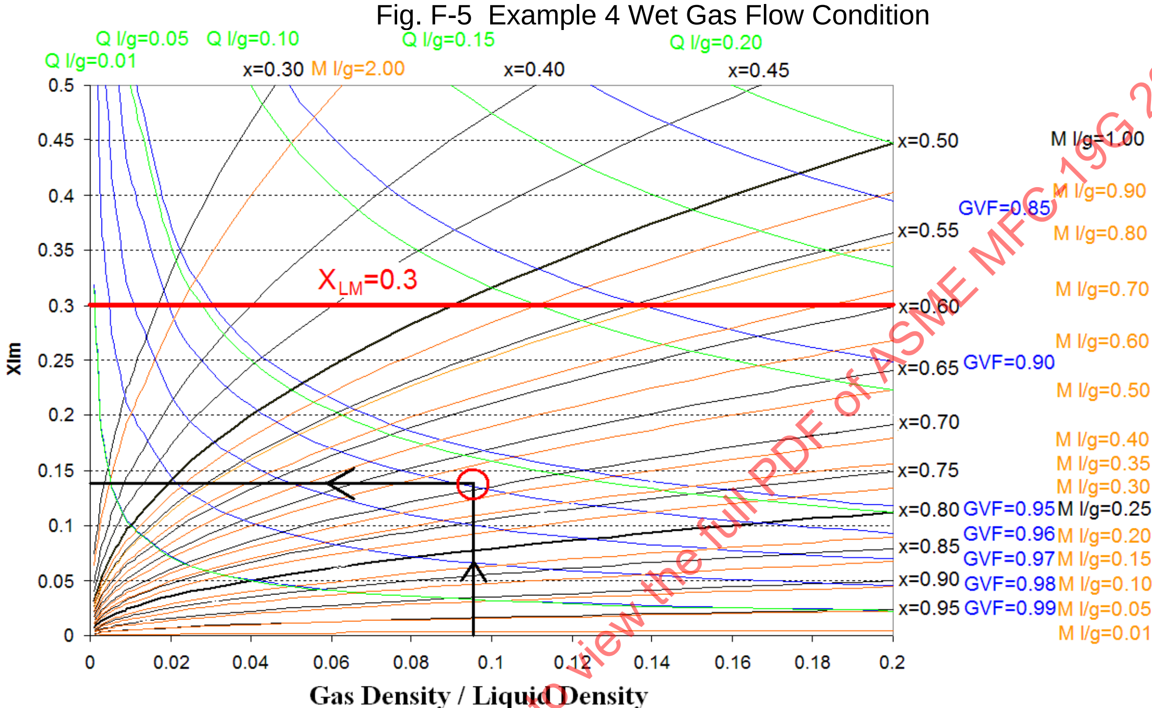
$$\Delta P_{tp} = (1.2)^2 * \Delta P_g = 1.44 * 109''WC = 157 \ ''WC$$

In this case the use of a 0-125 in. WC DP transmitter would have resulted in the saturation of the DP transmitter and the failure of the metering system to read any flow information.

Engineers should always choose the DP transmitter after considering the maximum expected wet gas DP.

Again for practice from the known gas-to-liquid density ratio and the known liquid-to-gas mass flow rate ratio the other wet gas parameters are calculated and read off the wet gas flow map (see Fig. F-5). As required, the graph and equations are seen to give the same information.

Fig. F-5 Example 4 Wet Gas Flow Condition



The initial information gave a gas flow rate of 69 500 kg/hr and a liquid flow rate of 31 250 kg/hr. The liquid-to-gas mass ratio (M l/g in Fig. D-5) is therefore approximately 0.45. At the gas-to-liquid density ratio of 0.0953 the flow condition is shown as the center of the red circle in Fig. D-5. The other parameters can be read off the figure and they are seen to match the calculations below:

from eq. (22)

$$X_{LM} = \frac{31{,}250}{69{,}500}\sqrt{\frac{82.4}{864}} = 0.139$$

from eq. (23)

$$GVF = \frac{1}{1+\left(\frac{m_l}{m_g} * \frac{\rho_g}{\rho_l}\right)} = \frac{1}{1+\left(\frac{31{,}250}{69{,}500} * \frac{82.3}{864}\right)} = 0.959$$

from eq. (24)

$$x = \frac{1}{1+\left(\frac{31{,}250}{69{,}500}\right)} = 0.690$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = \frac{31{,}250}{69{,}500} = 0.4496$$

from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = \frac{\rho_g}{\rho_l}\frac{\dot{m}_l}{\dot{m}_g} = \frac{82.3*31{,}250}{864*69{,}500} = 0.0428$$

from eq. (27)

$$LVF = \frac{\frac{\rho_g}{\rho_l}}{\frac{\rho_g}{\rho_l}+\frac{\dot{m}_g}{\dot{m}_l}} = \frac{\left(\frac{82.3}{864}\right)}{\left(\frac{82.3}{864}\right)+\left(\frac{69{,}500}{31{,}250}\right)} = 0.0411$$

F-6 EXAMPLE 5: APPLICATIONS WITH PARAMETERS OUT WITH THE TEST CENTER LIMITS

F-6.1 Explanation

A natural gas production company was retrofitting an offshore platform with the aim of maximizing the gas production rate. The two designs being considered had different pipe sizes. A 10 in. Schedule 160 or a 6 in. Schedule 160 (i.e., the existing infrastructure) were being considered. The meter manufacturer was requested to comment on the wet gas metering conditions of both possibilities.

Predicted maximum flow conditions on a natural gas platform:

Line size: 6 in. Schedule 160 pipe work, i.e., 5.187 in./ 0.1317 m
or 10 in. Schedule 160 pipe work, i.e., 8.500 in. / 0.2159 m
Pressure steady at 53 bara
Temperature steady at 88 ^{0}C.
Gas molecular weight: 19.5 mW
Gas mass flow: 3.68 Sm^3/day
Liquid “mass”[2] flow: 3 650 barrels/day (i.e., 3650 x 0.1589872 m^3/day = 580.3 m^3/day)
Gas density: 40.1 kg/m^3
Liquid density: 684.1 kg/m^3

Analysis for either design (i.e., parameters independent of meter size):

At the given flow conditions the gas mass flow is found to be 23.3 kg/s.

[2] It is common for industry to confuse mass and volume quotes. This is a typical example of a volume flow rate [i.e., barrels (which is a volume measurement) per unit time] being erroneously called a mass flow rate.

The liquid volume flow (quoted as mass flow) was 3 650 barrels/day, which is, when converting to SI units from the above quoted information, 4.59 kg/s.

Therefore:

$$X_{LM} = \frac{4.6}{23.3}\sqrt{\frac{40.1}{684.1}} = 0.048$$

(Note that here the ratio of mass flow rates in the Lockhart–Martinelli parameter calculation was done by ratios of kilograms per second. In the previous example kilograms per hour was used. In nondimensional number calculations the units chosen are not important as long as the units chosen remain constant throughout a particular calculation procedure.)

The gas-to-liquid density ratio is:

$$\frac{\rho_g}{\rho_l} = \frac{40.1}{684.1} = 0.0586$$

Therefore, the Lockhart–Martinelli parameter and the gas-to-liquid density ratio are both well within the wet gas test loop test envelopes.

For the 6-in. Schedule 160 meter case, the superficial gas velocity is:

$$U_{sg} = \frac{\dot{m}_g}{\rho_g A} = \frac{23.3}{40.1 * 0.01363} = \ 42.6 \ \ m/s$$

For the 6-in. Schedule 160 meter case, the gas densiometric Froude number is:

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{42.6}{\sqrt{9.81 * 0.1317}}\sqrt{\frac{40.1}{684.1 - 40.1}} = 9.35$$

The gas velocity for the existing 6 in. Schedule 160 is extremely high. It is higher than the maximum typically allowed by the operators. The gas densiometric Froude number is also extremely high. The value of 9.35 is far in excess of any published data on wet gas meter performance and far in excess of the maximum value reachable by the majority of wet gas test facilities. No published wet gas data was known to exist for such a flow condition and no published wet gas correlation (i.e., correction for a gas meter with a known liquid mass flow rate) or no wet gas meter was known to be proven at these extreme conditions. There could be no guarantee that any metering method would give the quoted system uncertainty due to large extrapolation of the correlation data sets being required. Most wet gas correlations are not guaranteed to not diverge at large extrapolations of various parameters.

For the 10-in. Schedule 160 meter case, the superficial gas velocity is:

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} = \frac{23.3}{40.1 * 0.0366} = \ 15.9 \ \ m/s$$

For the 10-in. Schedule 160 meter case, the gas densiometric Froude number is:

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{15.9}{\sqrt{9.81 * 0.2159}}\sqrt{\frac{40.1}{684.1 - 40.1}} = 2.73$$

For the 10 in. option the gas velocity for the existing 10 in. Schedule 160 was a common value typical of such a pipe size. The gas densiometric Froude number was well within the known test data envelopes. The major problem was that no data was publicly available to prove that wet gas data from a smaller meter could be applied to a 10 in. meter and that therefore the user could not rely on these existing correlations or systems to give their quoted uncertainties. Problems with regard to extrapolating data sets to different diameter meters have been discussed by Wood [44] and Steven [19, 21].

The solution decided upon was that as there was no known publicly available 10-in. wet gas meter data of any sort or no known publicly available 6-in. wet gas data with these extremely high gas densiometric Froude numbers both meter sizes needed calibrated. This was seen as the only way to avoid the situation of an extrapolation of any existing wet gas correlation causing the possibility of divergence and gross errors. Therefore, as there was no calibration saving in choosing one option over another to save on retrofit costs, the customer decided to utilize the existing infrastructure. The flow rates were finally reduced somewhat to comply with safety regulations, and the 6-in. schedule 160 meter was wet gas calibrated.

F-6.2 Comments on Example 5

The problem with wet gas calibration is that no existing test center can calibrate meters across the extensive ranges of wet gas conditions encountered by industry. A major restriction in wet gas meter technology is therefore the limits of the available wet gas flow test loops. With industry requiring a huge range of flow conditions metered it is not practical (or for that matter possible) for test centers to cover all possible ranges. The large flow rates, diameter and pressure ranges, and the varied types of liquid flowing with gases means that the test centers chose to design a test system according to their financial budget and what they predict is the most likely range of interest to the majority of the market. This restriction, along with the fact that most wet gas data is not held in the public domain, means that very often wet gas metering inquiries received by meter manufacturers are for flow conditions outside of the test data available to that manufacturer, often outside of the flow conditions of any test data set held in confidence by any organization and it is not uncommon for the flow in question to be beyond the current test envelope of any existing test center. Example 5 describes such a situation.

In this case the calibration still reached a maximum gas densiometric Froude number considerably higher than the manufacturer or meter the user had experience with. It was found that a difference in the gas mass flow prediction of greater than 10% existed between actual results and those predicted by the large extrapolation of the meter's published wet gas correlation.

The conclusion is for any wet gas meter application, large extrapolations produce large uncertainties and for assurance of accurate metering, wet gas calibration at the expected wet gas flow conditions is required.

The related wet gas flow conditions to a gas-to-liquid density ratio of 0.0586 and a Lockhart–Martinelli parameter of 0.048 are calculated below and read from Fig. F-6.

Note that the results of this analysis show that the flow is a wet gas flow. However, note that the analysis above still claims that wet gas flow meters cannot be applied and expected to work with low uncertainty without calibration. It should be understood that a flow condition with $X_{LM} \leq 0.3$ does not guarantee any wet gas meter will automatically work without calibration.

from eq. (23)

$$GVF = \frac{\sqrt{\frac{684.1}{40.1}}}{0.048 + \sqrt{\frac{684.1}{40.1}}} = 0.9885$$

from eq. (24)

$$x = \frac{1}{1 + 0.048\sqrt{\frac{684.1}{40.1}}} = 0.835$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = 0.048\sqrt{\frac{684.1}{40.1}} = 0.198$$

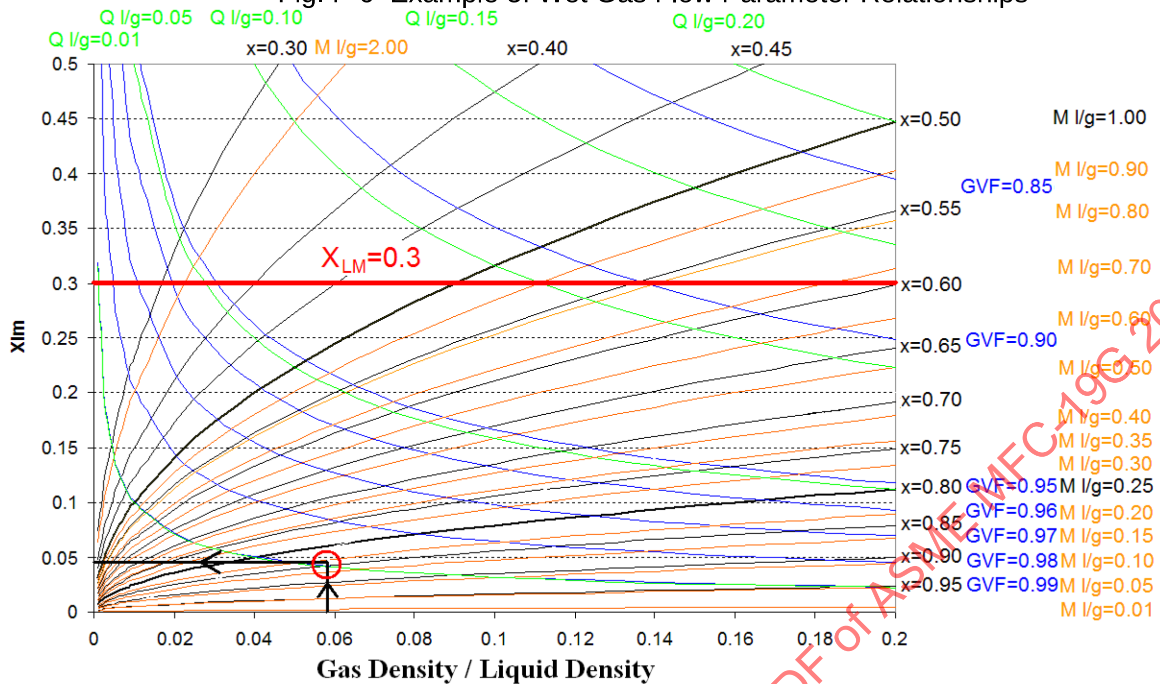
from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = 0.048\sqrt{\frac{40.1}{684.1}} = 0.0116$$

from eq. (27)

$$LVF = \frac{0.048\sqrt{\frac{40.1}{684.1}}}{1 + 0.048\sqrt{\frac{40.1}{684.1}}} = 0.0115$$

Fig. F-6 Example 5: Wet Gas Flow Parameter Relationships



F-7 EXAMPLE 6: FIELD PRODUCTION TRAJECTORY AND ITS IMPACT ON METERING

An important, but often neglected aspect of wet gas meter selection is the potential impact of the field production trajectory, multi-zone production, and future tieback.

The following data shows an example of a production profile at the start-up, mid-life, and end-life of a fictitious field. The subsequent calculations translate the production data into the corresponding wet gas flow dimensionless parameters for the start up, mid-life, and end of life. The intent is to show how, over the lifetime of a sub-sea pipe line, the two-phase flow conditions at a meter installation can change considerably due to reservoir depletion, producing from different levels/layers within the reservoir and addition of tie-backs upstream of the metering point. This production trajectory for a pipeline affects the flow pattern and the Lockhart–Martinelli parameter significantly. The expected meter performance can then be ascertained over the range of conditions it will encounter, and such considerations need to be taken into account when selecting the most appropriate meter. It is possible that for the full life prediction of the flow, no one meter on the market will work successfully throughout the lifetime of the well.

Line Size: 4-in. Schedule 160 pipe work, i.e., 3.438 in./0.08732 m. Therefore:

$$A = \frac{\pi}{4} D^2 = \frac{\pi}{4}(0.08732)^2 = 0.00599 \ m^2$$

Sample meter condition profile:

<u>Start of Life:</u>
Gas Mass Flow: 4.85 kg/s
Liquid Mass Flow: 0.15 kg/s
Gas Density: 88 kg/m^3
Liquid Density: 797 kg/m^3

<u>Middle of Life:</u>
Gas Mass Flow: 1.43 kg/s
Liquid Mass Flow: 0.463 kg/s
Gas Density: 80.1 kg/m^3
Liquid Density: 836 kg/m^3

<u>End of Life:</u>
Gas Mass Flow: 0.84 kg/s
Liquid Mass Flow: 1.60 kg/s
Gas Density: 41.3 kg/m^3
Liquid Density: 901 kg/m^3

Examining each condition in turn:

Start of Life Expected Conditions:

$$\frac{\rho_g}{\rho_l} = \frac{88}{797} = 0.11$$

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{0.15}{4.85}\sqrt{\frac{88}{797}} = 0.01$$

The superficial gas velocity (i.e., the average velocity of the gas if the gas phase flowed alone in the pipe) is:

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} = \frac{4.85}{88 * 0.00599} = 9.2 m/s$$

Therefore, the gas densiometeric Froude number is:

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{9.2}{\sqrt{9.81 * 0.08732}}\sqrt{\frac{88}{797 - 88}} = 3.5$$

It is therefore found that the initial flow expected is a wet gas flow as $X_{LM} < 0.3$. However, with X_{LM} =0.01 there is only a small quantity of liquid and in this situation it may, depending on the application, be acceptable to meter the gas flow with a single-phase gas meter and accept the liquid induced error as an increase in the gas flow rate uncertainty. At this low liquid loading, wet gas meters will give high uncertainty estimations of the liquid flow rate. The density ratio and the gas densiometric Froude number are high for the available test facilities but still within the range of what could be tested.

Middle of Life Expected Conditions:

$$\frac{\rho_g}{\rho_l} = \frac{80.1}{836} = 0.096$$

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{0.46}{1.43}\sqrt{\frac{80.1}{836}} = 0.1$$

The superficial gas velocity (i.e., the average velocity of the gas if the gas phase flowed alone in the pipe) is:

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} = \frac{1.43}{80.1 * 0.00599} = 2.99 m/s$$

Therefore the gas densiometeric Froude number is:

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}} \sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{2.99}{\sqrt{9.81 * 0.087325}} \sqrt{\frac{80.1}{836 - 80.1}} = 1.05$$

By the meter's mid life the wet gas characteristics of this flow have changed considerably compared to the start up flow conditions. The wet gas flow is now much wetter with X = 0.1 indicating that a more substantial quantity of liquid is now flowing. Metering this flow would require a wet gas meter or, if the liquid flow rate or liquid-to-gas flow rate ratio is available, a single-phase gas meter with a correlation that is useable at these flow conditions. The gas-to-liquid density ratio and the gas densiometeric Froude number values are within the available wet gas test facility range.

End of Life Expected Conditions:

$$\frac{\rho_g}{\rho_l} = \frac{41.3}{901} = 0.046$$

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g} \sqrt{\frac{\rho_g}{\rho_l}} = \frac{1.6}{0.84} \sqrt{\frac{41.3}{901}} = 0.41$$

The superficial gas velocity (i.e., the average velocity of the gas if the gas phase flowed alone in the pipe) is:

$$\bar{U}_{sg} = \frac{\dot{m}_g}{\rho_g A} = \frac{0.84}{41.3 * 0.00599} = 3.4 m/s$$

Therefore, the gas densiometeric Froude number is:

$$Fr_g = \frac{\bar{U}_{sg}}{\sqrt{gD}} \sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{3.4}{\sqrt{9.81 * 0.087325}} \sqrt{\frac{41.3}{901 - 41.3}} = 1.59$$

At the end of the flow's life X_{LM} = 0.41. As X_{LM} > 0.3 the flow is now a general two-phase flow and not considered a wet gas in this Report. Here wet gas meters or single-phase gas meters with wet gas correlations for known liquid information may or may not be capable of metering this flow. A multiphase meter may be required. Even though the gas-to-liquid density ratio and the gas densiometeric Froude number values are well within the available wet gas test facility range, the liquid quantity may be too high for wet gas flow meter technologies.

A point of interest here is that the changes in the flow conditions over the life of the well and the inclusion of tie-backs not only cause a light liquid load wet gas flow to change to a moderate liquid load wet gas flow and then to a general two-phase flow but also lead to the flow pattern changing over time.

Figure F-7 shows the three conditions plotted on the Shell Flow Pattern Map. For clarity the drawings of the flow patterns are superimposed on the various flow pattern regions.

At startup the Shell Flow Pattern Map predicts the flow will be an annular mist flow (denoted as "mist" flow on the map). As the well ages towards the mid-life conditions, the map is predicting that the flow pattern will gradually shift into a transition zone between annular mist flow to stratified flow. At the mid-life condition the map predicts the flow pattern will be in a transitional area between stratified and annular mist. As the well continues to age the flow pattern remains in this transition zone of these two distinct flow patterns and at the end of life condition the flow is also getting close to slug flow conditions.

Calculating the other wet gas parameters for each condition for practice:

Start of well life:

from eq. (23)

$$GVF = \frac{1}{1+\left(\frac{\dot{m}_l}{\dot{m}_g} * \frac{\rho_g}{\rho_l}\right)} = \frac{1}{1+\left(\frac{0.15}{4.85} * \frac{88}{797}\right)} = 0.997$$

from eq. (24)

$$x = \frac{1}{1+\frac{\dot{m}_l}{\dot{m}_g}} = \frac{1}{1+\left(\frac{0.15}{4.85}\right)} = 0.97$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = \frac{0.15}{4.85} = 0.03$$

from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = \frac{\rho_g}{\rho_l}\frac{\dot{m}_l}{\dot{m}_g} = \frac{88}{797} * \frac{0.15}{4.85} = 0.003$$

from eq. (27)

$$LVF = \frac{\left(\frac{\rho_g}{\rho_l}\right)}{\left(\frac{\rho_g}{\rho_l} + \frac{\dot{m}_g}{\dot{m}_l}\right)} = \frac{\left(\frac{88}{797}\right)}{\left(\frac{88}{797} + \frac{4.85}{0.15}\right)} = 0.003$$

Middle of well life:

from eq. (23)

$$GVF = \frac{1}{1+\left(\frac{\dot{m}_l}{\dot{m}_g} * \frac{\rho_g}{\rho_l}\right)} = \frac{1}{1+\left(\frac{0.46}{1.43} * \frac{80.1}{836}\right)} = 0.97$$

from eq. (24)

$$x = \frac{1}{1+\frac{\dot{m}_l}{\dot{m}_g}} = \frac{1}{1+\left(\frac{0.46}{1.43}\right)} = 0.756$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = \frac{0.46}{1.43} = 0.323$$

from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = \frac{\rho_g}{\rho_l}\frac{\dot{m}_l}{\dot{m}_g} = \frac{80.1}{836} * \frac{0.46}{1.43} = 0.031$$

from eq. (27)

$$LVF = \frac{\left(\frac{\rho_g}{\rho_l}\right)}{\left(\frac{\rho_g}{\rho_l} + \frac{\dot{m}_g}{\dot{m}_l}\right)} = \frac{\left(\frac{80.1}{836}\right)}{\left(\frac{80.1}{836} + \frac{1.43}{0.46}\right)} = 0.03$$

End of well life:

From eq. (23)

$$GVF = \frac{1}{1+\left(\frac{\dot{m}_l}{\dot{m}_g} * \frac{\rho_g}{\rho_l}\right)} = \frac{1}{1+\left(\frac{1.6}{0.84} * \frac{41.3}{901}\right)} = 0.92$$

from eq. (24)

$$x = \frac{1}{1 + \frac{\dot{m}_l}{\dot{m}_g}} = \frac{1}{1 + \left(\frac{1.6}{0.84}\right)} = 0.34$$

from eq. (25)

$$\frac{\dot{m}_l}{\dot{m}_g} = \frac{1.6}{0.84} = 1.92$$

from eq. (26)

$$\frac{\dot{Q}_l}{\dot{Q}_g} = \frac{\rho_g}{\rho_l}\frac{\dot{m}_l}{\dot{m}_g} = \frac{41.3}{901} * \frac{1.6}{0.84} = 0.088$$

from eq. (27)

$$LVF = \frac{\left(\frac{\rho_g}{\rho_l}\right)}{\left(\frac{\rho_g}{\rho_l} + \frac{\dot{m}_g}{\dot{m}_l}\right)} = \frac{\left(\frac{41.3}{901}\right)}{\left(\frac{41.3}{901} + \frac{0.84}{1.6}\right)} = 0.081$$

The flow conditions through the life of the meter are plotted on Fig. F-8. It is seen that the flow starts as almost dry, becomes wetter, and finally crosses the boundary of what this Report calls the limit of wet gas flow into general two-phase flow. The flow path sketched between the stated flow conditions is assumed and therefore not precise.

The metering engineers in these situations have a judgment call to make. A relatively inexpensive dry gas meter will operate satisfactorily for the initial flow conditions as the flow is not very wet, but as the flow becomes wetter the dry gas meter will become less accurate and finally give little or no useable information. A wet gas meter will be considerably more expensive, and for the initial flow conditions it will not offer much advantage over the single-phase flow meter as the liquid loading is too small for the liquid prediction to be of a relatively low uncertainty. However, as the mid-life flow condition is reached and the flow is wetter, a wet gas meter would be the most appropriate system to be in use. Wet gas meters, though, typically do not work far into the general two-phase flow region (i.e., $X_{LM} > 0.3$). For these flow conditions general two-phase meters (i.e., the meters generally called multiphase meters) are required. Multiphase meters are typically very expensive compared to single-phase meters and more expensive than most wet gas meters. For this extra expense no advantage over single-phase or wet gas meter systems would be gained for the initial flow, little advantage would be gained mid life over a less expensive wet gas meter and finally a multiphase meter would operate better than the other options at the end of the meter's life, but it only shows an advantage at the lowest gas flow (i.e., money) production at the end of the meter's productive life. Therefore, the problem is in these cases,[3] the most expensive meter is required for the case of the smallest revenue flow. These decisions are as much business

[3] It should be noted that there are many two-phase flows in industry where the entire life of the flow is a general two-phase flow and multiphase meters are successfully employed from start up to end of the systems life.

financial management decisions as technical decisions; therefore, further discussion on this topic is beyond the scope of this Report.

This example shows that it is often not possible to select one wet gas flow meter for the entire life of a wet natural gas production metering position as the flow conditions and flow patterns can change considerably over time and change beyond the operating range of any one wet gas metering technology. Scheers [72] gives an independent discussion on the types of wet gas/two-phase/multiphase flow patterns a typical meter installation may experience during the lifetime of a well trajectory.

Fig. F-7 Changing Flow Patterns During the Meter Vs. Life

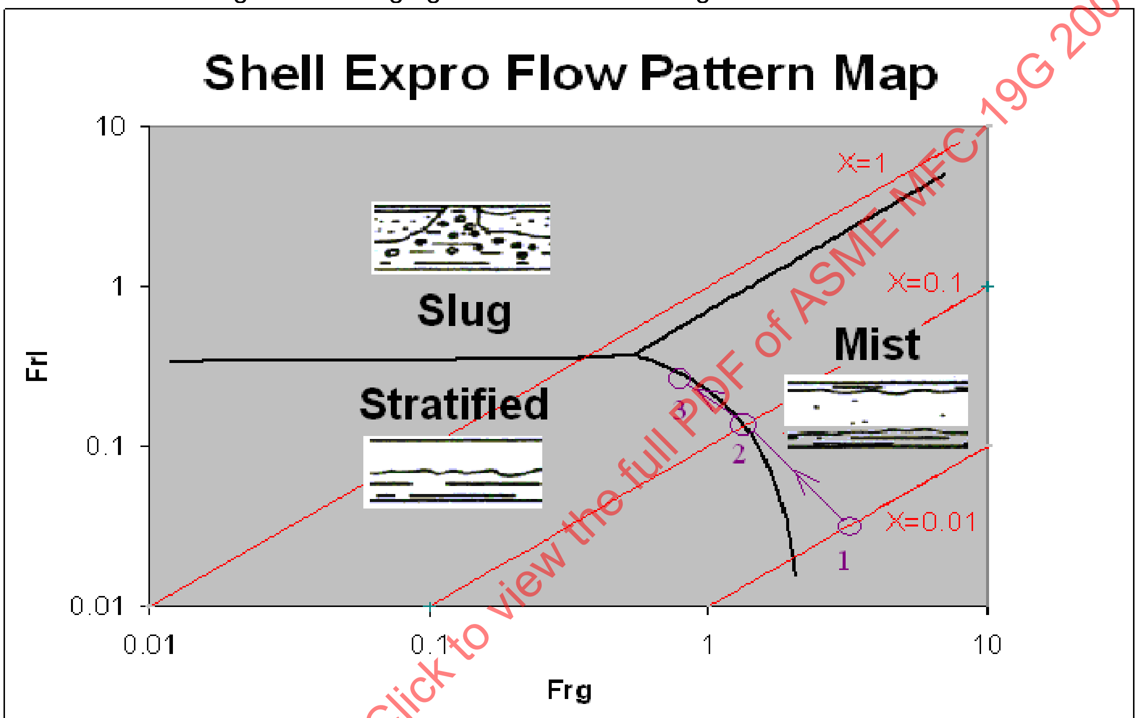
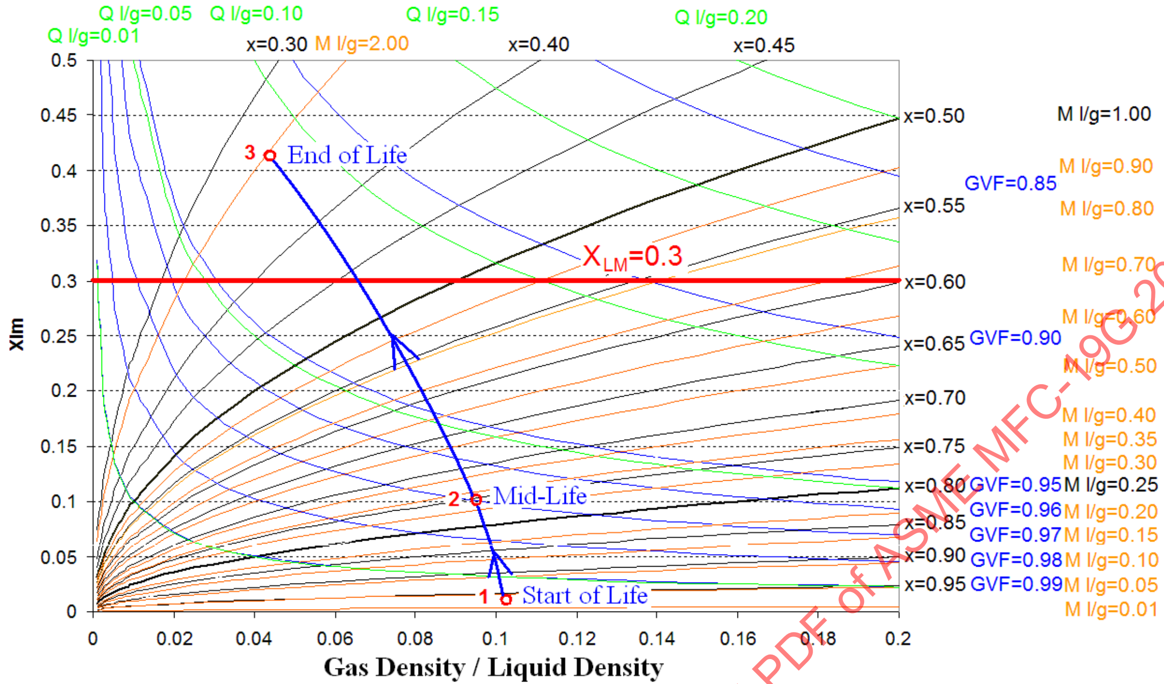


Fig. F-8 Changing Two-Phase Flow Conditions Throughout the Life of a Natural Gas Production Well



NONMANDATORY APPENDIX G[1]
DIFFERENTIAL PRESSURE METER WET GAS CORRELATIONS

G-1 INTRODUCTION

Differential pressure (DP) meters are at the time of writing generally accepted to be the most commonly used generic single-phase gas meter type used with wet gas flows. There are several wet gas reports in the public domain that discuss particular DP meter design performances with wet gas flow. Several of these papers present wet gas correlations for particular DP meters. That is, many of these papers offer correction factors for particular geometry DP meters that, for a known liquid flow rate or liquid-to-gas flow rate ratio, will correct the liquid-induced error on the gas meters gas flow prediction. A few of these papers discuss the performance of published wet gas correlations when tested independently. The information discussed here is available to the public through journal papers, conference papers, and vendor press releases.

DP meter wet gas correlations are generally created from experimental data that is obtained mainly from wet gas tests centers and occasionally from field data. There are only a few industrial grade test centers, and these have relatively limited ranges of wet gas flow conditions in comparison to the varied demands of industry. Therefore, many single-phase DP meters when used as wet gas meters (when the liquid flow rate is known or estimated) have their reaction to the wet gas predicted by extrapolation of the correlations found by these limited test centers and field data. The suitability of extrapolating DP meter wet gas correlations is an open question in industry. The following is a discussion on DP meter wet gas correlations published and their stated performances.

Not enough scientific knowledge yet exists in the subject of wet gas flow metering to create a standard and ASME cannot take responsibility for the validity of the claims made in the published literature reproduced below. ASME does not approve or disapprove of any method. All meter performances discussed describe their operation with steady wet gas flows with no appreciable phase change.

Of the differential pressure (DP) meter devices regularly used by industry several designs (but by no means all designs) have published wet gas flow research papers. All DP meters operate using the principles of the conservation of mass and energy. This leads to the universal single-phase DP meter gas flow eq. (G-1).

$$\dot{m}_g = EC_d Y_g A_t \sqrt{2\rho_g \Delta P_g} \qquad \text{(G-1)}$$

where

A_t = area of the minimum cross-sectional area sometimes called the "throat"

C_d = the discharge coefficient due to the gas mass flow rate ($\dot{m}_g$) being used to calculate the Reynolds number (only applicable to DP meters with discharge coefficients versus Reynolds number calibrations)

E = geometric constant called the Velocity of Approach [see eq. (A-20)]

$\dot{m}_g$ = actual gas mass flow rate due to the single-phase differential pressure (ΔP_g)

ρ_g = gas density

Y_g = the gas expansibility factor when it is predicted by the single-phase gas expansibility factor using the single-phase differential pressure (ΔP_g)

The derivation of eq. (G-1) can be found in most fluid mechanics and flowmetering textbooks.

When the flow is a wet gas the differential pressure read (ΔP_{tp}) is different to that which would have been read if that quantity of gas flowed alone. The result is an erroneous gas flow rate prediction [see eq. (G-2)].

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

$$\dot{m}_{g\,apparent} = EA_tY_{g_{tp}}C_{d_{tp}}\sqrt{2\rho_g\Delta P_{tp}} \qquad \text{(G-2)}$$

where

$C_{d_{tp}}$ = discharge coefficient due to the apparent gas mass flow rate ($\dot{m}_{g\,apparent}$) being used to calculate the Reynolds number. (If the discharge coefficient is considered constant this value does not change between Reynolds numbers or dry and wet gas flow operations.)

$\dot{m}_{g\,apparent}$ = gas mass flow rate prediction due to the two-phase (i.e., wet gas) differential pressure (ΔP_{tp})

$Y_{g_{tp}}$ = gas expansibility factor when it is predicted by the single-phase gas expansibility factor using the two-phase differential pressure (ΔP_{tp})

A positive error is usually called an "overreading." A negative error is usually called an underreading. It is the correction of this liquid induced error that is the basis for all DP meter wet gas correlations. Often the liquid induced error is expressed in terms of the ratio shown in eq. (G-3):

$$\frac{\dot{m}_{g\,Apparent}}{\dot{m}_g} = \frac{EA_tY_{g_{tp}}C_{d_{tp}}\sqrt{2\rho_g\Delta P_{tp}}}{EA_tY_gC_d\sqrt{2\rho_g\Delta P_g}} = \frac{Y_{g_{tp}}C_{d_{tp}}}{Y_gC_d}\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} \cong \sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} \qquad \text{(G-3)}$$

as typically $Y_{g_{tp}}C_{d_{tp}} \cong Y_gC_d$.

G-2 THE HOMOGENEOUS CORRECTION FACTOR FOR ALL DIFFERENTIAL PRESSURE FLOW METERS

The homogeneous flow model is almost certainly the oldest correction factor for two-phase flows. Up until the 1950s when there were no published wet gas corrections for any flow meter type, engineers often applied the homogeneous flow model assumption due to lack of any alternative. This model is still periodically used in industry for different meter types. Its use in DP meter wet gas flow applications has decreased since the release of several wet gas correlations in the last few decades, but it is still used in some high flow rate mist flow applications where the assumptions are closer to being correct and in cases where the engineers are not up to date with the state-of-the-art of wet gas metering.

The homogeneous model is a simple correction factor traditionally used to correct for any DP meter type's liquid-induced error when used with general two-phase flow. The model assumes a pseudo-single-phase flow by averaging the phase densities. That is, it is assumed the phases are perfectly mixed (i.e., homogenized) and therefore a single-phase flow eq. (G-1) can be used to predict the total mass flow [i.e., eq. (G-4)].

$$\dot{m}_{total} = EA_tY_{g_{tp}}C_{d_{total}}\sqrt{2\rho_{homogenous}\Delta P_{tp}} \qquad \text{(G-4)}$$

where

$C_{d_{total}}$ = discharge coefficient of the meter at the homogeneous fluids Reynolds number. This is only relevant for DP meters with discharge coefficients versus Reynolds number calibrations.

That is:

$$C_{d_{total}} = f\left(\text{Re}_{total}\right) = f\left(\frac{4\dot{m}_{total}}{\pi\mu_{\text{hom}\,ogenous}D}\right)$$

$Y_{g_{tp}}$ = expansibility prediction of the homogeneous fluid using the two-phase DP value

$\rho_{\text{hom}\,ogenous}$ = density of the perfectly mixed phases, i.e., the homogeneous fluid

where Re_{total} is the actual Reynolds number if the two-phase flow is homogeneous and $\mu_{\text{hom}\,ogenous}$ is the homogeneous fluids viscosity (see Nonmandatory Appendix H).

The derivation of the homogeneous density is given in Nonmandatory Appendix H. For a known quality or liquid flow rate, the gas mass flow rate is calculated using this homogeneous assumption by the iteration of eq. (G-5).

It can be seen that the liquid mass flow rate or the flow quality is required as an input. The desired gas mass flow rate is derived by iteration. The initial gas mass flow rate input of the uncorrected meter reading usually ensures a close enough initial estimation for a quick convergence of the iteration.

$$\dot{m}_g = \frac{EA_tYC_d\sqrt{2\rho_g\Delta P_{tp}}}{\sqrt{1+\left\{\sqrt{\frac{\rho_g}{\rho_l}}+\sqrt{\frac{\rho_l}{\rho_g}}\right\}X_{LM}+X_{LM}^2}} = \frac{m_{g,apparent}}{\sqrt{1+\left\{\sqrt{\frac{\rho_g}{\rho_l}}+\sqrt{\frac{\rho_l}{\rho_g}}\right\}X_{LM}+X_{LM}^2}}$$

(G-5)

In 1975 Smith (R.V.) and Leang [59] discussed the performance of the homogeneous model with orifice meters, but in 1977 Smith (L.T.) and Murdock [60] cast doubt on some of the data used and therefore the conclusions drawn. In 1986 Lin [40] discussed the homogeneous model's performance. Lin reports that the homogeneous model gave better predictions of orifice plate meter overreadings when the two-phase flow had a relatively low liquid loading. Lin also shows the homogeneous correction performs better with the higher-pressure orifice plate meter data. In 2001 Steven [11] discusses the performance of the homogeneous model with wet gas Venturi data and suggests that data fits better predicts the Venturi wet gas response than the homogeneous model across the range of conditions tested.

In many industrial applications the flow conditions are not such that a wet gas flow could be modeled as a homogeneous flow, so various DP meter types have correlations formed from various data sets to attempt to predict DP meter wet gas flow response.

G-3 THE ORIFICE PLATE METER WET GAS CORRELATIONS

The orifice plate meter is the most common single-phase DP meter and hence by default[2] has become the most common wet gas meter. There are several general two-phase flow metering papers discussing orifice plate meters in the literature, and while some include correlations covering the range of wet gas flow, others do not. Lin [40] gives a comprehensive list of general two-phase flow metering correlations. The first published two-phase / wet gas flow orifice plate meter work known to us was by Schuster [4] in 1958. This paper discusses 3-in. and 4-in. orifice meter tests of an unknown beta ratio and unknown gas flow rate. The liquid content was stated as up to 700 barrels per million standard cubic feet of gas,

[2] Many wet gas flows are metered as dry flows due to operator ignorance of the liquids' presence or lack of knowledge on what corrective action needs to be taken. Hence, the popular simple orifice plate meter is often installed in wet gas flow applications.

the fluids were natural gas with water and distillate, and tests were carried out at 1,000 psi, 600 psi, and 185 psi. The orifice plate meters had both pipe and flange taps. Schuster reported the effect of different liquids on the meter was the same and that there was no difference between the pipe and flange pressure tapping results. Schuster stated that the orifice plate gas meter gives an increasing overreading with increasing liquid loading. It also states, however, that the relationship of the overreading to pressure (or gas-to-liquid density ratio) was in reverse to that described in all later DP meter papers. With Schuster not presenting a wet gas correlation, field data being relatively high uncertainty data, and with the data now being untraceable, Schuster's paper is noted simply as the first paper to record the overreading characteristic of a DP meter metering wet gas flow.

The DP meter two-phase research paper by Murdock [5] discussed orifice plate meters and gave a two-phase correction factor (that encompassed wet gas flow). Murdock's correlation effectively[3] related the Lockhart–Martinelli parameter to the liquid-induced error. Chisholm [7, 8] later released a wet gas flow orifice plate meter correlation that includes a pressure effect as well as the Lockhart–Martinelli parameter effect for predicting the liquid-induced error. Other two-phase orifice plate meter correlations that were created with data sets that included wet gas flows are the James [62], Smith and Leang [59], and Lin [9] equations.

Since the publication of the Murdock and Chisholm correlations, orifice plate meters have become less popular for wet gas metering than other DP meters for several reasons. These include reported buckling of plates (see Fig. 9.1-7) in adverse flow conditions such as pressure pulses or slugging) and the trapping of liquid and dirt around the plate (see Fig. 9.1-7). Nevertheless, the orifice plate meter continues to be used with wet gas flows, and the Murdock correlation appears to be the best known and therefore most popular correction factor. Furthermore, Murdock was the first to plot the square root of the ratio of the actual read DP with two-phase flow to the DP that should have been read if that quantity of gas flowed alone in the pipe versus a parameter that is efectively the Lockhart–Martinelli parameter.[4] This has since been used by most DP meter wet gas flow metering researchers and is now generally described as a "Murdock plot." Note, with just a slight approximation to allow for second order discharge coefficient and expansibility shifts, from eq. (G-3) the ordinate in the Murdock plot can also be called the ratio of the apparent gas flow rate (for the uncorrected gas flow prediction) to the actual gas flow rate.

G-3.1 The Murdock Correlation

Murdock's analysis [5] has more scientific reasoning than the homogeneous model. The data points were from the U.S. Navy's steam research facilities (now decommissioned) and from various data sets from the oil and gas industry. The data is therefore for a variety of fluid combinations (wet steam, air/ water, gas/ salt water and gas/liquid hydrocarbon combinations) and a wide scope of wet gas conditions. The reported data ranges are shown in Table G-3.1.

The Murdock correlation is shown as eq. (G-6). The mathematical model it is based on has the assumption that the flow pattern is separated flow only (as is described in Nonmandatory Appendix H). However, the data set used to fit the constant gradient (M = 1.26) is not necessarily limited to separated flow. It should be noted that ASME orifice plate meter 1*D* and ½ *D* pressure tapping locations were used at all times.

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1 + M \dfrac{\dot{m}_l}{\dot{m}_g}\left(\dfrac{C_{d_g} Y_g}{C_{d_l}}\right)\sqrt{\dfrac{\rho_g}{\rho_l}}} \cong \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1 + MX_{Murdock}} \approx \frac{\dot{m}_{g_{Apparent}}}{1 + 1.26 X_{LM}} \qquad \text{(G-6)}$$

[3] See Nonmandatory Appendix A.

[4] Murdock [5] never used the term "Lockhart–Martinelli parameter." This term came into use at a later period. The parameter Murdock used is very similar but not identical to the Lockhart–Martinelli parameter. (See Nonmandatory Appendix A for details.)

It can be seen that the liquid mass flow rate or the liquid-to-gas flow rate ratio is required as an input. In fact Murdock states that one of the uses of the correlation is to find the quality (or "dryness fraction") for a known total mass flow rate (a common problem in boiler systems). The desired gas mass flow rate is derived by iteration of eq. (G-7).

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1+1.26X_{LM}} = \frac{\dot{m}_{g_{Apparent}}}{1+\left(1.26\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}\right)} \qquad \text{(G-7)}$$

Note that the Murdock correlation shown in eq. (G-7) is in fact the uncorrected orifice plate meter reading (i.e., the numerator term) corrected by dividing this uncorrected value by a correction factor.

The initial gas mass flow rate input of the uncorrected meter reading usually ensures quick convergence of the iteration. The limits of the data that was used in the correlation construction are shown in Table G-1 (where subscripts *sg* and *sl* denote values at superficial gas and liquid velocities, respectively).

Table G-1 The Murdock Data Range

1.01 bar ≤ P ≤ 63 bar	13,000 ≤ Re_{sg} ≤ 1,270,000
0.025 bar ≤ ΔP ≤ 1.25 bar	54 ≤ Re_{sl} ≤ 46,600
0.11 ≤ x ≤ 0.98	0.2602 ≤ β ≤ 0.5
63.35 mm ≤ D ≤ 101.6 mm	25.4 mm ≤ d ≤ 31.8 mm

Murdock claimed that the correlation fitted all the orifice plate meter data used in the correlation development to ±1.5% for a known quality. No independent check on the validity of the Murdock gradient being 1.26 is known to have been published, although Chisholm [8] and then Lin [9] have indicated that the gradient is actually a function of the gas-to-liquid density ratio.

Due to lack of alternatives, it is known to have been applied to other DP meters and to applications with conditions outside of the Murdock data set. It should also be noted that in recent years Murdock's equation has become the most widely known of the DP meter wet gas correlations. As a result of this and its simple final form, there have been cases in industry where the Murdock equation form has been applied to data fits of different DP meters to find a simple wet gas correction factor for that particular DP meter in a particular range of wet gas flow conditions. Examples are Conoco Phillips are known to have fitted Venturi meter wet gas data (obtained from a well in the North Sea while it had a dedicated separator) to the Murdock equation form, Steven [11] fitted the eq. (G-6) form to Venturi data and Emerson Process Management utilized Murdock's equation to describe the conditioning plate meter's response to wet gas [61].

G-3.2 The James Correlation

In 1965 James [62] looked at the response of orifice plate meters to wet steam flows. It was duly noted from Murdock's paper [5] that the DP read (ΔP_{tp}) was higher than when the vapor/steam phase flowed alone (ΔP_g), and James found this to be true from independent experimental results on wet steam flows through 7.9 in. orifice meters of very high betas (0.707 and 0.837). James indicated that by assuming a homogeneous density at the meter inlet, the homogeneous model did not give the correct results when tested on his data set. Therefore, it was concluded that the homogeneous model needed to be modified.

The resulting equation (derived in Nonmandatory Appendix H) is shown here as eq. (G-8).

$$\dot{m}_g = EA_t YC_d \sqrt{2\rho_g \Delta P_{tp}} \sqrt{\frac{x^2 \rho_l}{x^{1.5}\rho_l + (1 - x^{1.5})\rho_g}} \quad \text{(G-8)}$$

where

$$x = \frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g} \quad (16)$$

The mathematical model it is based on is the homogeneous flow with the quality/dryness fraction term (x) empirically modified to account for the fact that the flow is not perfectly homogenized. The modification of the quality/dryness fraction is presented in the form:

$$x_m = x^n \quad \text{(G-9)}$$

where *n* is found by experiment.

James worked with general two-phase flow data and not specifically wet gas flow data. However, engineers working with wet steam flows are known to have utilized the James method with wet gas data and produce in-house unpublished versions of the form shown in eq. (G-10).

$$\dot{m}_g = EA_t YC_d \sqrt{2\rho_g \Delta P_{tp}} \sqrt{\frac{x^2 \rho_l}{x^n \rho_l + (1 - x^n)\rho_g}} \quad \text{(G-10)}$$

It can be seen that the liquid mass flow rate is required as an input to eq. (G-9) (which will have a chosen value of *n*) before that equation is substituted into eq. (G-10). The desired gas mass flow rate is then derived by iterating eq. (G-10). It is normal procedure to use the DP meter's uncorrected gas flow rate prediction as the initial input to ensure a short iteration procedure.

James [62] gave some uncertainties for two pressures and a varying (conventional) quality. These are given in Table G-2. It will be immediately evident from Table G-2 that James was considering general two-phase flow and not wet gas flow. However, like Murdock's model the James model is known to have been chosen by engineers with real wet gas flow industrial problems. The same (or a similar) procedure as described in Nonmandatory Appendix H to create the James correlation is known to have been carried out on two-phase and wet gas flow orifice plate meter data to create a usable orifice plate wet gas correlation for saturated steam. Therefore, we have duly mentioned the correlation form here. (We do not know of any literature in the public domain describing details of these updated James style correlations.)

Table G-2 Quoted Uncertainties of the James Correlation

Pressure (psia)	Quality, *x*	% Variation in Total Mass Flow
100	0.5	± 4.5
100	0.22	± 12
100	0.05	± 8.6
400	0.5	±4.2
400	0.22	±10.5
400	0.05	±4.4

Note that the James correlation can only predict the gas mass flow rate if the information about the liquid mass flow rate or liquid-to-gas flow rate ratio is initially known. We do not know of any published work that discusses independent wet gas flow test analysis on the James correlation. Smith and Leang [59] do discuss the standard James correlation performance with general two-phase flow (although few points in the data sets considered by Smith and Leang cover wet gas flows). For the 87 orifice meter points, Smith and Leang had for flows with qualities (x) greater than 0.2, the root mean squared fractional deviation was reported at 0.063. For the 53 Venturi nozzle meter points Smith and Leang had for flows with qualities (x) greater than 0.2, the root mean squared fractional deviation was reported at 0.43.

Finally, it should be noted that the data James used was for larger orifice meters than most other available data. The two beta ratios were large, and the 0.837 beta meter was seen by many researchers as excessive and not within the common industrial range. There is currently little in the literature describing the effect diameter has on the wet gas response of otherwise identical meters.

G-3.3 The Chisholm Correlation

In 1967 Chisholm [7] published a general two-phase flow correlation for orifice plate meters. The Chisholm correlation is shown as eq. (G-11).

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} = \frac{\dot{m}_{g_{Apparent}}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} \qquad \text{(G-11)}$$

X_{LM} is the Lockhart–Martinelli parameter, and C is a function where the numerical value is found by experiment. In 1967 Chisholm [7] originally gave different values for C ($2.66 \le C \le 4.76$) when using different general two-phase data sets. In 1977, Chisholm [8] refined C for two-phase flows with Lockhart–Martinelli parameter values less than unity (i.e., including wet gas) to eq. (G-12).

$$C = \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} + \left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} \qquad \text{(G-12)}$$

Chisholm's model, assumes there is no liquid entrainment in the gas but it includes the effect of pressure [note that eq. (G-12) is a function of the gas density which is directly related to pressure]. The data sets Chisholm used is not necessarily limited to separated flow. The desired gas mass flow rate is derived by iteration of eq. (G-13).

$$\dot{m}_g = \frac{\dot{m}_{g_{Apparent}}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} = \frac{\dot{m}_{g_{Apparent}}}{\sqrt{1 + C\left(\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}\right) + \left(\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}\right)^2}} \qquad \text{(G-13)}$$

Note that the numerator of eq. (G-11) is the apparent gas flow rate if no correction for wet gas is applied. The denominator is the correction factor. Chisholm's equation is the first DP meter wet gas correction factor that indicates that the density ratio of the wet gas flow has an effect on the overreading that is independent to the Lockhart–Martinelli parameter. However, with the use of the Lockhart–Martinelli parameter in the correlation the gas mass flow rate can only be predicted if the information about the liquid mass flow rate or liquid-to-gas flow rate ratio is initially known. For a known liquid mass flow rate the gas mass flow rate is derived by iteration. The initial gas mass flow rate input of the uncorrected meter reading usually ensures quick convergence of the iteration.

The Chisholm correction factor indicates that the phase density ratio (i.e., the pressure for a given fluid combination) of a wet gas flow can directly affect the error induced by a given amount of liquid. However, although the Chisholm correlation is an advance in knowledge, it appears to be less well known and therefore used less than the Murdock correlation. The main interest of Chisholm's equation in modern day wet gas flow metering technology seems to be that de Leeuw's wet gas Venturi meter correlation is based on Chisholm's equation.

Chisholm [8] compares the C value found by experiment for $X_{LM} \leq 1$ to that found by eq. (G-12). For the various data sets with a wide range of flow conditions (many of which are not wet gas flows) and orifice plate geometries, the results range from very good to poor. We do not know of any published independent orifice plate meter wet gas flow check on the Chisholm equation. However, it is known that orifice meter research has recently occurred and results are due for release in 2007, which indicate that Chisholm's correlation works reasonably well within the limits of the data sets used to create the correlation.

Steven [11] discusses the results of using the Chisholm equation with a 6 in. 0.55 beta ratio Venturi meter. The Chisholm equation did not do particularly well, but it should be noted that the equation was for a different DP meter and Chisholm's data sets were for inlet diameters of 4 in. and less. This result indicates that it is unwise to use wet gas DP meter correlations for one meter type with another. (When through necessity this is required, it should be done with extreme caution and acceptance of higher uncertainties.)

No uncertainty prediction on the gas mass flow rate prediction was given by Chisholm. Most of the data set used by Chisholm was Murdock's data. The derivation of the Chisholm correlation is given in Nonmandatory Appendix H.

G-3.4 The Smith and Leang Correlation

The Smith and Leang correlation [59] is created from a collection of orifice plate meter wet saturated steam data sets. Most of the data is considered general two-phase data (as it is of the range $x \geq 0.1$) but some of the data extends into the wet gas flow region. Smith and Leang were the first researchers to state directly that the flow pattern would have an influence on a DP meter's wet gas flow performance. As such, when they added a correction factor [which they called a "blockage factor" (ξ)] for wet gas flow to the standard DP meter equation [eq. (G-14)] they choose a form that they considered appropriate to express this fact [eq. (G-15)].

$$\dot{m}_g = E\xi A_t Y C_d \sqrt{2\rho_g \Delta P_{tp}} \tag{G-14}$$

where

$$\xi = 0.637 + 0.4211x - \frac{0.00183}{x^2} \tag{G-15}$$

Smith and Leang do not state a performance uncertainty in terms of the maximum percentage difference from the reference meter. However, they do give the root mean squared fractional deviation of the results as 0.087 for flows with qualities, $x \geq 0.2$. Unfortunately, the technical paper discussions published at the end of the paper state that a significant amount of the data came from nonstandard orifice plate arrangements. Smith and Leang also had Venturi nozzle data and for the 53 points where the qualities were $x \geq 0.2$, the root mean squared fractional deviation was 0.198.

This research is discussed here for completion and to show the earliest known mention of the understanding of the importance of flow patterns. (However, it is noteworthy that Murdock used two types of orifice pressure tapping configurations and the wet gas flow performance results appear to have been the same for both configurations. This could suggest that the Smith and Leang nonstandard orifice plate arrangement data may still give useful information as some changes in the configuration of the orifice plate are seen to be irrelevant to the wet gas performance.)

It will be noticed that the liquid mass flow rate or the flow quality is required as an input to eq. (G-15) before that equation is substituted into eq. (G-14) for the iteration on the gas mass flow to be carried out. Use of the DP meter's uncorrected gas flow rate prediction as the initial input would ensure a short iteration procedure.

Equation (G-15) is formed with orifice plate data only but for the first time in wet gas/two-phase flow metering research literature a recognition is made by Smith and Leang that such correlations may be in practice applied to different types of DP meters. They therefore checked the correlation's performance with the limited nozzle and Venturi meter two-phase data available to them. These data set ranges are listed in the Smith and Leang paper [59]. The results showed not surprisingly, that the correlation did not work as well with the nozzle and Venturi meters as it did with the orifice plate meters.

The Smith and Leang correlation does not have the modeling detail of Murdock or Chisholm and does not account for any gas-to-liquid density ratio effect like Chisholm. However, for the first time it does register the possibility that the flow pattern may affect the liquid-induced meter error. It also registers for the first time the fact that in industry, engineers are sometimes, through lack of alternatives, forced to apply one DP meter type's wet gas correction factor to another DP meter type on the assumption that some correction is better than none.

Note that the Smith and Leang correlation can only predict the gas mass flow rate if the liquid mass flow rate or liquid-to-gas flow rate ratio is initially known.

G-2.5 The Lin Correlation

In 1982 Lin [9] published a two-phase flow correlation for orifice plate meters.

$$\dot{m}_g = \frac{E A_t C_d \sqrt{2\rho_g \Delta P_{tp}}}{1+\theta X_{LM}} \tag{G-16}$$

where

$$\theta = 1.48625 - 9.26541\left(\frac{\rho_g}{\rho_l}\right) + 44.6954\left(\frac{\rho_g}{\rho_l}\right)^2 - 60.165\left(\frac{\rho_g}{\rho_l}\right)^3 - 5.12966\left(\frac{\rho_g}{\rho_l}\right)^4 - 26.5743\left(\frac{\rho_g}{\rho_l}\right)^5 \tag{G-17}$$

The Lin correlation is created from a collection of orifice plate meter saturated steam data sets and a data set from tests with the two-phase flow of the refrigerant R-113. Most of the data is general two-phase data but some of the data extends into the wet gas flow region. Lin considered separated flow through orifice plate meters and considered the meter's wet gas error to be dependent on the shear between the separated phases and this shear to be solely dependent on pressure (i.e., the gas-to-liquid density ratio). This paper effectively updates the Murdock correlation to include the Chisholm statement that orifice plate meter wet gas errors are dependent on the gas-to-liquid density ratio. This correlation is lesser known than Murdock and Chisholm's correlations.

Note that the Lin correlation shown in eq. (G-16) is in fact the uncorrected orifice plate meter reading (i.e., the numerator term) corrected by dividing the uncorrected value by a correction factor. The correlation can only predict the gas mass flow rate if the information about the liquid mass flow rate or liquid-to-gas flow rate ratio is initially known. The gas mass flow rate will be found by substituting eq. (G-17) into eq. (G-16) and iterating. It is common practice to start the iteration with the uncorrected gas flow rate prediction to assure quick convergence of the iteration.

As with the Murdock and Chisholm models, it should be noted that with the Lin model there is an assumption of stratified two-phase flow through the orifice. However, with the various data sets Lin used it is likely different flow patterns existed. Lin claims that the correlation was found to predict the total mass flow rate to a root mean squared error (δ) of 0.012. No independent check on the correlation is known for orifice plate meters. Steven [11] applied Lin's equation to a 6 in., Schedule 80, 0.55 beta ratio standard Venturi meter with wet gas flows. A root mean squared error (δ) of the order of 0.0462 was reported.

G-2.6 Chevron Independent Wet Gas Orifice Plate Meter Tests

An independent investigation into orifice plate meter performance was presented by Ting [3, 22]. A single 200 mm orifice plate meter of 0.7 beta ratio was tested with air and water at 41.7 bar, 15.6^{o}C between gas Reynolds numbers of 4 to 9 million and with very low Lockhart–Martinelli parameters ($4.5e^{-5} \le X_{LM} \le 4.5e^{-3}$). This single test is of interest as it reports an orifice plate meter's response to a smaller Lockhart–Martinelli parameter value than tested in any of the existing correlation data sets and it does not perform as is predicted by these correlations. Ting reports that the orifice plate meter has a negative error (or "underreading") induced by the presence of the liquid rather than a positive error (or "overreading") predicted by all other researchers who tested higher Lockhart–Martinelli parameter values. Ting found that with an increasing Lockhart–Martinelli parameter from 4.5e^{-5} to 4.5e^{-3} the gas mass flow rate underreading increased from approximately 0 to -1.7%.

The conclusion drawn from this is that orifice plate meter users should be wary of applying published correction factors when the Lockhart–Martinelli parameter is less than 0.02. Below this value, the correlations tend to show an overreading of no more than 2% and the data stops at approximately a Lockhart–Martinelli parameter of 0.01. Therefore, at a Lockhart–Martinelli parameter of 0.02 or less, an uncertainty of ±2% could be assumed with no correlation applied.

G-4 THE NOZZLE METER WET GAS CORRELATION

Only one paper is known to still exist[5] discussing nozzle meters with general two-phase flow. Chisholm and Leishman [6] discuss nozzles' metering wet steam. The same equation as Chisholm used for orifice plate meters with wet gas [eq. (G-11)] was fitted with a nozzle meter value for "C" found from the available nozzle general two-phase and wet gas flow data. However, Chisholm indicates that there is not enough data for the analysis to have any real accuracy. The data used was air and water at atmospheric conditions, and Chisholm states it was assumed this data matched wet steam at 0.365 MN/m^2, the pressure where the wet steam density ratio matched that of air/water at atmospheric conditions. From analysis of this data set a value of C_{nozzle} = 14 was found that was considerably higher than the range of values found for orifice plate meters. Therefore, the only existing nozzle meter wet gas correlation is eq. (G-18).

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{\sqrt{1 + C_{nozzle} X_{LM} + X_{LM}^2}} = \frac{\dot{m}_{g_{Apparent}}}{\sqrt{1 + 14 X_{LM} + X_{LM}^2}} \qquad \text{(G-18)}$$

No uncertainty in the gas flow prediction is given, but if forced to meter wet gas flow with a nozzle meter, it is the only known correlation in the literature. It is therefore (as it is with all the stated correlations) the meter operator's discretion whether to use this correction factor. Again, the liquid mass flow rate is required as an input. The desired gas mass flow rate is derived by iteration. The initial gas mass flow rate input of the uncorrected meter reading usually ensures quick convergence of the iteration.

G-5 THE VENTURI METER WET GAS CORRELATIONS

The Venturi meter is a popular wet gas flowmeter in the oil and gas production industry. There are several well known papers in the literature regarding a classical Venturi meter's wet gas flow performance, notably Neverdeen and Washington [25, 26], de Leeuw [10], Stewart [16], and Britton [69]. (ASME knows of no published wet gas research with Venturi nozzle meter designs.)

The earliest three papers known on Venturi meter performance with wet gas flows were from Washington [25] and Neverdeen [26], who presented the joint work of Shell, and the Dutch gas company NAM and Jamieson [70], who presented the work of Shell Exploration & Production U.K. The papers of Nederveen and Washington et. al. are largely duplicate papers with Washington named on each and, like Jamieson, they do not offer any new correlations but investigate the possibility of extending the existing orifice plate meter correlations to use with Venturi meters.

In 1989 Nederveen and Washington et al. discussed the results of testing a wet gas flow with reported conditions of natural gas and water flow through a 4 in. Venturi meter (of unstated beta ratio) between 80 and 100 bar and a liquid-to-gas ratio (LGR) range with a maximum of 400 m^3/ 10e^6 Nm^3 (i.e., $X_{LM} \leq$ 0.14). The orifice plate meter correlations of Murdock and Chisholm were applied, and it was reported that they gave similar predictions to the experimentally found overreadings.

5 The literature references a NEL report that has since been lost by NEL: Graham E.J., "The Flow of Air/Water Mixtures Through Nozzles" National Engineering Laboratory Report No. 308, East Kilbride, Glasgow, Scotland, 1967. (unknown if any copies of this paper survive).

In 1993 Jamieson [70] reviews the earlier work of Nederveen and Washington et al. and comments: "Although there were differences between the field measurements and Murdock's and Chisholm's expressions, the errors involved were acceptable for NAM's applications. However, these differences are too large to be generally acceptable for fiscal and custody transfer purposes." Jamieson went on to state, "In the field measurements made for NAM the slope of the graph of overreading versus liquid content for Venturi meters was some 5% higher than that predicted by Murdock's equation." It was then suggested that as the Murdock gradient value of 1.26 was found empirically, this value could be replaced by a value found from a wet gas flow data set taken from any meter in question. That is, eq. (G-7) can have the constant gradient value of 1.26 changed to any empirically found value, *M*.

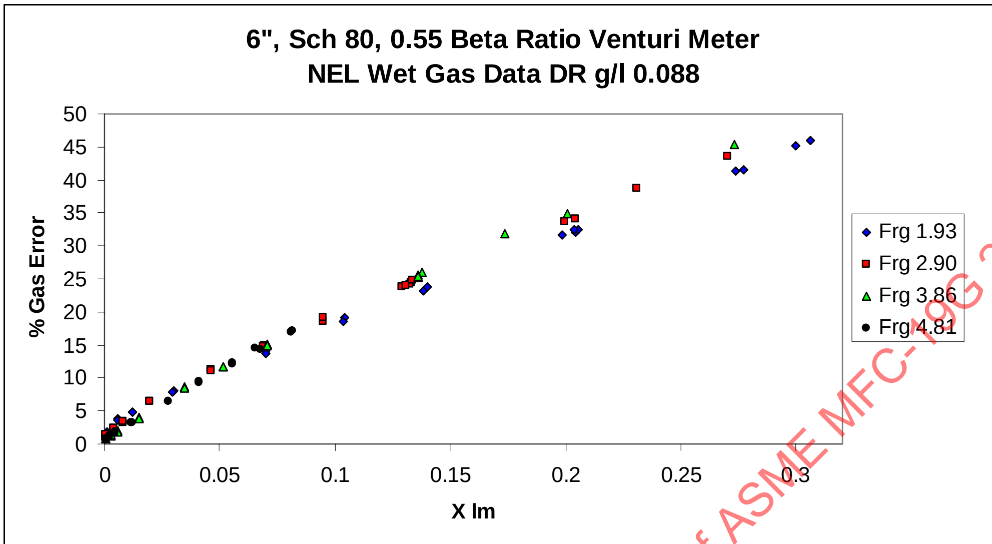
$$\dot{m}_g = \frac{\dot{m}_{g_{Apparent}}}{1 + MX_{LM}} \tag{G-19}$$

It is generally accepted that this procedure has since been done in-house by different oil and gas companies. No research regarding these projects is known to have been released. One oil and gas company operating in the North Sea has verbally told researchers that a 4 in. Venturi at 45 bara gave a Murdock gradient of 1.5. No further knowledge of the research is publicly known.

In 1997 de Leeuw [10] stated that the results of the analysis of the Coevorden field data discussed by Washington et al. [25] did not tell the complete story. It was reported that the fact that both the Murdock and Chisholm equations give good results at 90 bar, with up to 4% by volume liquid fraction, was a coincidence as extrapolation shows that for other line pressures, Murdock's and Chisholm's methods do not agree. However, de Leeuw does state that the Venturi meter has a higher overreading than the orifice plate meter. It is also pointed out that the experimental test range at Coevorden was relatively limited. That is, although natural gas at high pressure was used there was little variation in the pressure, and the flow conditions were all located in a small part of the Shell Expro flow pattern map (see Fig. 5-1), which indicated stratified wavy flow with no entrainment.

Initially de Leeuw discussed the performance of a 4 in. 0.4 beta ratio meter tested at the SINTEF wet gas flow facility in Trondheim, Norway and in a gas field application. Here de Leeuw shows that like the orifice plate meter, the Venturi meter has an overreading with wet gas flow that has a magnitude dependent on the Lockhart–Martinelli parameter and the gas-to-liquid density ratio. That is, the overreading increases with increasing Lockhart–Martinelli parameter, and for increasing gas-to-liquid density ratio with all other flow parameters constant, the overreading decreases. These are the same trends as noted for orifice plate meters. However, de Leeuw also went on to state the discovery that overreading is also dependent on the gas densiometric Froude number. For all other flow parameters constant, de Leeuw found that an increasing gas densiometric Froude number corresponded with an increasing overreading. Steven [11] independently verified these findings with 6 in. 0.55 beta ratio Venturi meter data. Fig. G-1 supplements the Figs. 6.1.1-3 through 6.1.1-5 in showing sample "Murdock plot" results from NEL wet gas tests with a 6 in., Schedule 80, 0.55 beta ratio Venturi meter.

Fig. G-1 NEL / DTI 6 in., 0.55 Beta Venturi Meter Wet Gas Data Set With Fixed Pressure / Gas-to-Liquid Density Ratios (DR g/l), and Separate Gas Densiometric Froude Numbers



Note that whereas a pressure effect is noticeable in Fig. 6.1.1-4 a gas densiometric Froude number effect is noticeable in the set pressure data plotted in Fig. 6.1.1-5 and here with Fig. G-1.

Stewart [16] discusses later Venturi meter wet gas tests carried out on behalf of the British government's Department of Trade and Industry (DTI) at NEL for several different Venturi meter geometries. The findings of de Leeuw are again independently verified, and Stewart also indicates that the beta ratio of a Venturi meter has an influence on the magnitude of a wet gas flow overreading. The trend found was as the beta ratio increases, the overreading decreases for a set wet gas flow condition. This then matched the earlier findings of Stewart et al. [12], where it was stated that the larger the beta ratio of a cone type DP meter the smaller the wet gas overreading for a set wet gas flow condition. Fig. 6.1.1-6 shows the beta ratio overreading influence Stewart reported.

It is evident then that wet gas correlations for particular DP meter types should be applied to the same meter types with the same beta ratios as the meters used to obtain the correlations data. Failure to do this could result in an increased uncertainty of the correlation results.

G-5.1 The de Leeuw Correlation

The de Leeuw correlation [10] is given as eqs. (G-11), (G-20), (G-21), and (G-22).

$$\dot{m}_g = \frac{EA_tC_{d_g}Y_{g_{tp}}\sqrt{2\rho_g\Delta P_{tp}}}{\sqrt{1+CX_{LM}+X_{LM}^2}} = \frac{\dot{m}_{g\,Apparent}}{\sqrt{1+CX_{LM}+X_{LM}^2}} \qquad \text{(G-11)}$$

$$C = \left(\frac{\rho_l}{\rho_g}\right)^n + \left(\frac{\rho_g}{\rho_l}\right)^n \qquad \text{(G-20)}$$

$$n = 0.606\left(1-e^{-0.746\,Fr_g}\right) \quad \text{For } Fr_g \geq 1.5 \qquad \text{(G-21)}$$

$$n = 0.41 \quad \text{For } 0.5 \leq Fr_g \leq 1.5 \qquad \text{(G-22)}$$

This correlation was created from data taken from two test matrices, one from a wet natural gas production field (Coevorden) and one from the SINTEF Multiphase Flow Laboratory test center at Trondhiem in Norway. The Coevorden test meter was a 4 in. Venturi, but no Schedule or beta ratio is recorded [25, 26]. The fluids were natural gas and water, and the pressure was between 78 and 98 bar. The Coevorden Lockhart–Martinelli parameter and gas densiometric Froude number ranges were not given, but an approximate Lockhart–Martinelli parameter range of 0.12 to 0.34 and a gas densiometric Froude number range of 0.5 to 1.5 is deducible from information given by de Leeuw [10].

The SINTEF meter was a Schedule 80, 0.401 beta ratio meter, and the fluids were nitrogen and diesel oil. The flow conditions were pressure range of 15 bar to 90 bar (i.e., an approximate gas-to-liquid density ratio range of 0.024 to 0.12), a Lockhart–Martinelli parameter range of 0 to 0.3 and a gas densiometric Froude number range of 0.5 to 4.8.

The de Leeuw 4 in. 0.401 beta ratio Venturi meter was tested at the SINTEF wet gas flow test centre with a combination of nitrogen and diesel oil, a pressure range of between 15 and 90 bar, a gas Densiometric Froude number range of between 1.5 and 4.8, and the maximum Lockhart–Martinelli parameter of 0.3. The basic model for the correlation was the orifice plate meter Chisholm eq. (G-11). However, like for Chisholm's data set, the data was not limited to a separated flow pattern. This led to de Leeuw's finding that the flow pattern map prediction of the Shell map used (see Fig. 5-1) predicted that the border of two flow patterns was where the data set happened to be showing a shift in Venturi meters' response.

The correlation is based on Chisholm's orifice plate eq. (G-11), but de Leeuw changed a constant in Chisholm's wet gas flow correlation to a function of the gas densiometric Froude number. Due to the change in the meter response across the flow pattern boundary de Leeuw offers two values for this parameter *n* depending on the gas densiometric Froude number [as shown by eqs. (G-21) and (G-22)].

Again the liquid mass flow rate is required as an input. The desired gas mass flow rate is derived by iteration. The initial gas mass flow rate input of the uncorrected meter reading usually ensures quick convergence of the iteration.

The uncertainty claimed by de Leeuw was 2% uncertainty with a few outliers within the scope of the data set. In 2003 de Leeuw [48] gave an update to the Venturi meter wet gas development and claimed that the de Leeuw correlation had been proven to extend up to 6 in. Venturi meters at the K-Lab wet gas facility in Norway. A more detailed discussion of the de Leeuw correlation is given in Appendix H.

G-5.2 Independent Wet Gas Venturi Meter Test

In 2001 Steven [11] applied de Leeuw's equation to a 6 in., Schedule 80, 0.55 beta ratio standard Venturi meter with wet gas flows. A root mean squared error (δ) of the order of 0.0211 was reported.

In 2003 Stewart et al. [16] compared the performance of the de Leeuw correlation with a Steven Venturi correlation[6] [11] by use of new NEL wet gas data. The de Leeuw correlation worked well for the 4 in. 0.4 beta Venturi within the range of test data used by de Leeuw. However, a distinct beta ratio effect was noticeable as would be expected as Stewart had shown Fig. 6.1.1-6 in this same paper and de Leeuw's correlation is for a set beta of 0.4 only.

Stewart showed that the de Leeuw correlation was reasonably accurate when applied to the flow conditions of the NEL 4-in. 0.4 beta ratio Venturi wet gas data set (the same geometry as de Leeuw's test meter). However, Stewart also reported that the de Leeuw correlation did not predict the 0.75 beta ratio Venturi data well, which is likely to be due to the beta effect. Fig. G-2 shows a sample of Stewart's results.

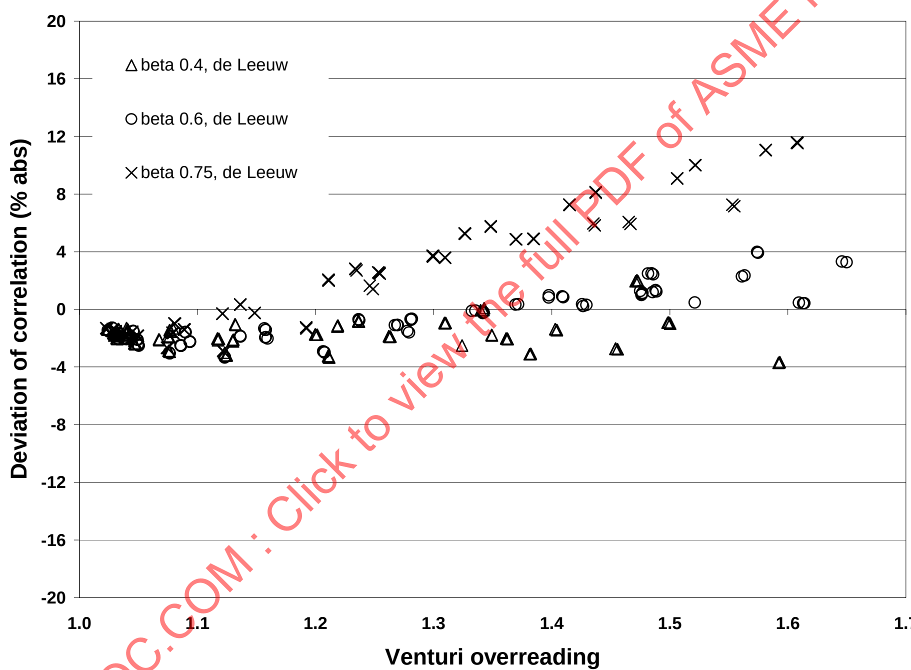
[6] By 2002 an update to the Steven Venturi correlation had been given by Stewart and Steven [12] where the data was re-fitted and the pressure term polynomials replaced with gas-to-liquid density ratio terms that did not diverge on extrapolation to higher values. Nevertheless, the later paper by Stewart [16] and the current NEL wet gas flow metering training course still reproduce the earlier Steven Venturi correlation. Due to the fact that no significant improvement was seen in the new data fit compared to de Leeuw's correlation the Steven Venturi correlation has been largely disregarded — including by Steven. Hence, no further details are given.

The x-axis shows the overreading (and note increasing overreading indicates increasing Lockhart–Martinelli parameter) and the y-axis shows the percentage deviation predicted from the actual (i.e., reference) gas flow rate. The 4 in. 0.40 beta ratio Venturi meter (i.e., the same geometry as de Leeuw's Venturi meter) had the overreading corrected to within 4%. The 4 in. 0.60 beta ratio Venturi meter had the overreading corrected to within 4%.

The 4 in. 0.75 beta ratio Venturi meter had differences of up to 12%. The trends were the same for all three pressures tested 15, 30, and 60 bar(g), but the higher the pressure the better de Leeuw's correlation performed for all beta ratio Venturi meters.

It is not currently known what, if any, diameter effect there is and hence the de Leeuw correlation should be applied to other diameters than 4 in. with caution. The latest work by NEL [17, 18] also suggests that if the liquid properties are significantly different between the diesel oil used by de Leeuw and the field application (e.g. water is present) this may affect the overreading magnitude, and even if all other parameters are constant and within the data range of de Leeuw's work, biases can occur.

Fig. G-2 NEL 4 in. Venturi Meter Wet Gas Data for 30 bar(g)
Corrected by the de Leeuw Correlation



G-5.3 The Cone Style DP Meter

Four major technical papers are known that discuss cone type DP meters' performance with wet gas flows [12, 13, 14, 15]. Steven [15] summarizes most of the known knowledge for cone type meter wet gas flow performance.

Figures G-3 and G-4 reproduce published results [15] in the form of Murdock plots from the NEL wet gas tests for a 6 in., Schedule 80, 0.75 beta ratio cone type DP meter. These figures show that the cone type DP meter has the same trends with wet gas flow as Venturi meters. That is, the liquid-induced gas flow rate overreading is dependent on the Lockhart–Martinelli parameter, the gas-to-liquid density ratio, and the gas densiometric Froude number. A pressure (i.e., gas-to-liquid density ratio) effect is noticeable in Fig. G-3 (Note that for set fluid types gas-to-liquid density and pressure are directly related). A gas densiometric Froude number effect is noticeable in the set gas pressure data plotted in Fig. G-4.

Fig. G-3 All NEL DTI 6 in. 0.75 Beta Ratio V-Cone Meter Data

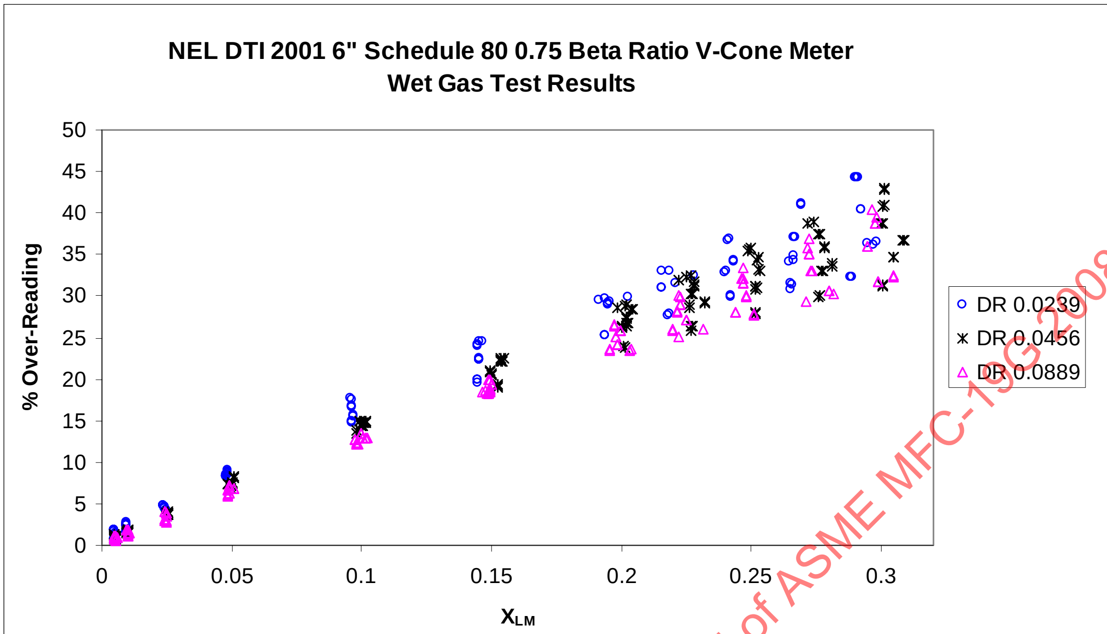
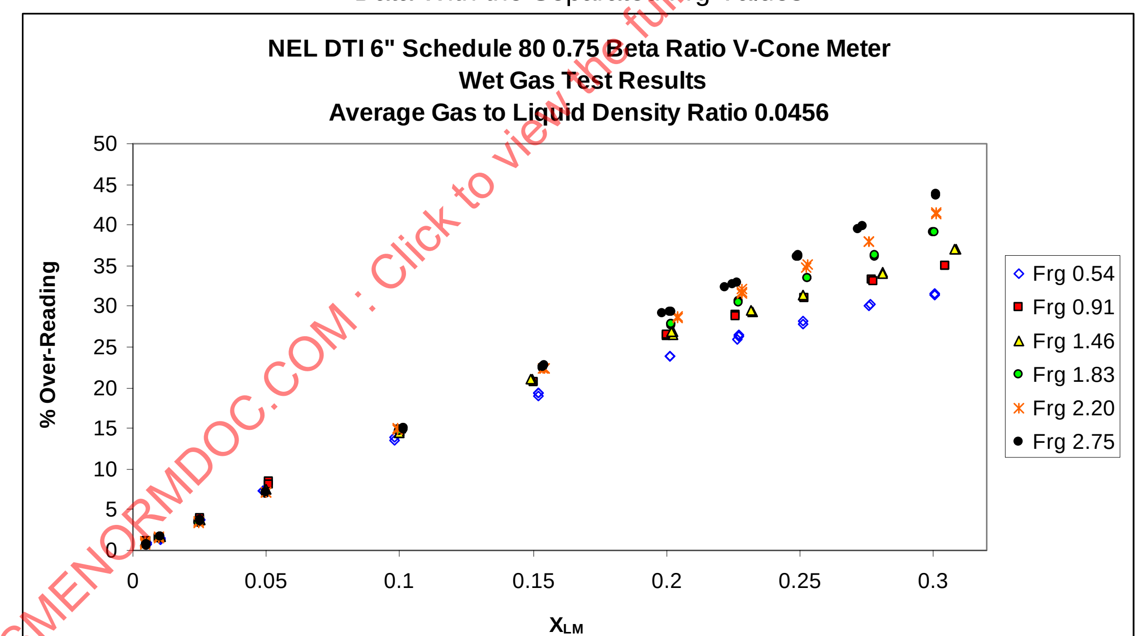


Fig. G-4 The 30 Bar NEL DTI 6 in. 0.75 Beta Ratio V-Cone Meter Data With the Separated Frg Values



Although the trends are the same, it has been reported by Stewart et al [12] that the cone type DP meter has a smaller overreading than the Venturi meter for the same flow conditions. Figure G-5 shows a reproduction of a graph from Ting [68] that compares 4 in. 0.75 beta ratio Venturi and V-Cone meters wet gas responses. (The third data set in Fig. G-5 shows the result of another wet gas meter design and is not relevant to this paragraph's discussion.)

Figure G-6 shows the original graph from Stewart et al. [12] that first indicated that the beta ratio of a generic DP meter may influence the magnitude of a liquid induced gas flow rate overreading. As was later reported for Venturi meters, the larger the beta ratio the lesser the overreading for a given wet gas flow condition.

In 2002 two beta ratio correlations were released. The majority of the later work published has been concentrated on the 0.75 beta ratio meter and in 2005, Steven [15] summarized the published work on the 0.75 beta ratio cone type DP meter. Here it was stated that at gas-to-liquid density ratios less than 0.027, the gas-to-liquid density ratio effect on the overreading disappeared (probably due to flow pattern considerations). Hence, a modified 0.75 beta ratio equation was produced. This correlation accounts for these gas-to-liquid density ratio considerations. The final 0.75 beta ratio cone-type DP meter corelation is therefore given by eqs. (G-23) and (G-24) through (G-26).

Fig. G-5 Published CEESI Results of the Comparison of Data Between a Venturi and Cone Type DP Meter by a Joint Industry Project (JIP)

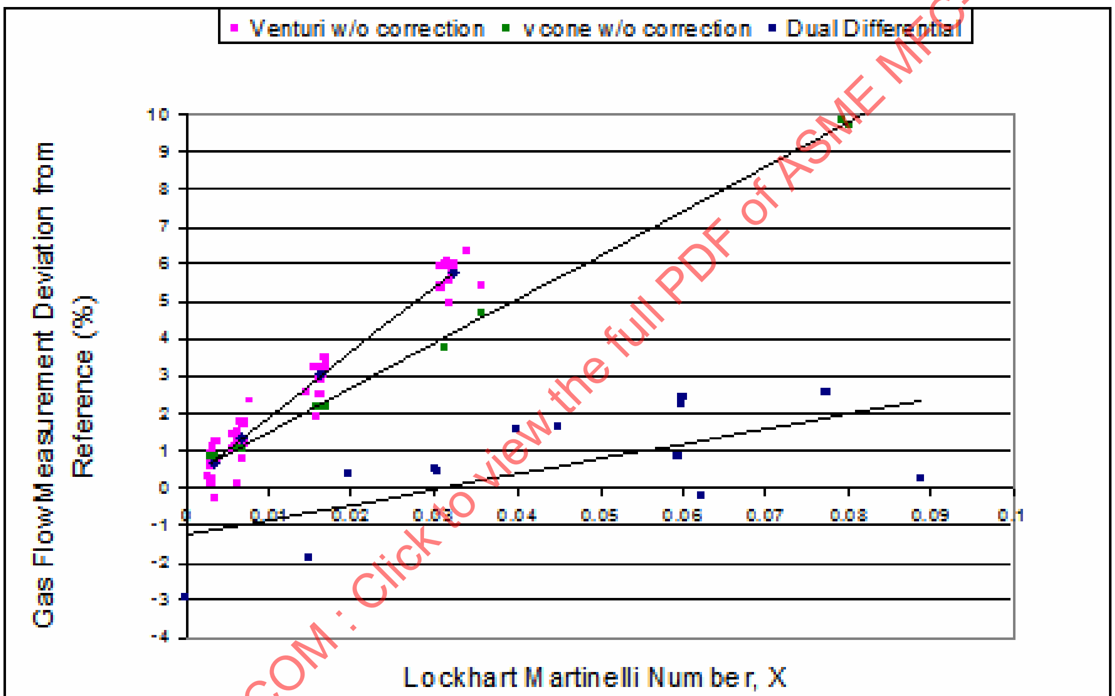
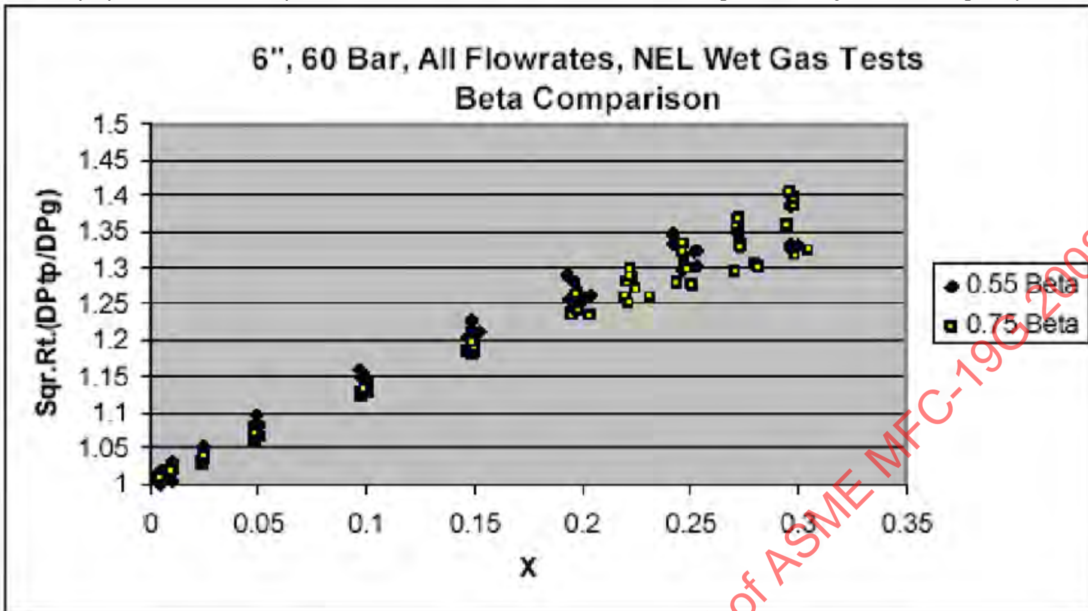


Fig. G-6 Published NEL Results of the Comparison of Cone-Type DP Meter Beta Ratios
(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom)



$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{\left(\frac{1 + AX + BFr_g}{1 + CX + BFr_g}\right)} \quad \text{(G-23)}$$

G-5.4 Steven 0.75 Beta Ratio Cone-Type DP Meter Correlation

For $\frac{\rho_g}{\rho_l} \geq 0.027$, then

$$A = -0.0013 + \frac{0.3997}{\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(G-24)}$$

$$B = 0.0420 - \frac{0.0317}{\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(G-25)}$$

$$C = -0.7157 + \frac{0.2819}{\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(G-26)}$$

For $\frac{\rho_g}{\rho_l} < 0.027$, then A=2.431 , B=-0.151 , C=1.

Note that at the gas-to-liquid density ratio value of 0.027, eqs. (G-24) through (G-26) gives the constants A = 2.431, B = -0.151, and C =1.

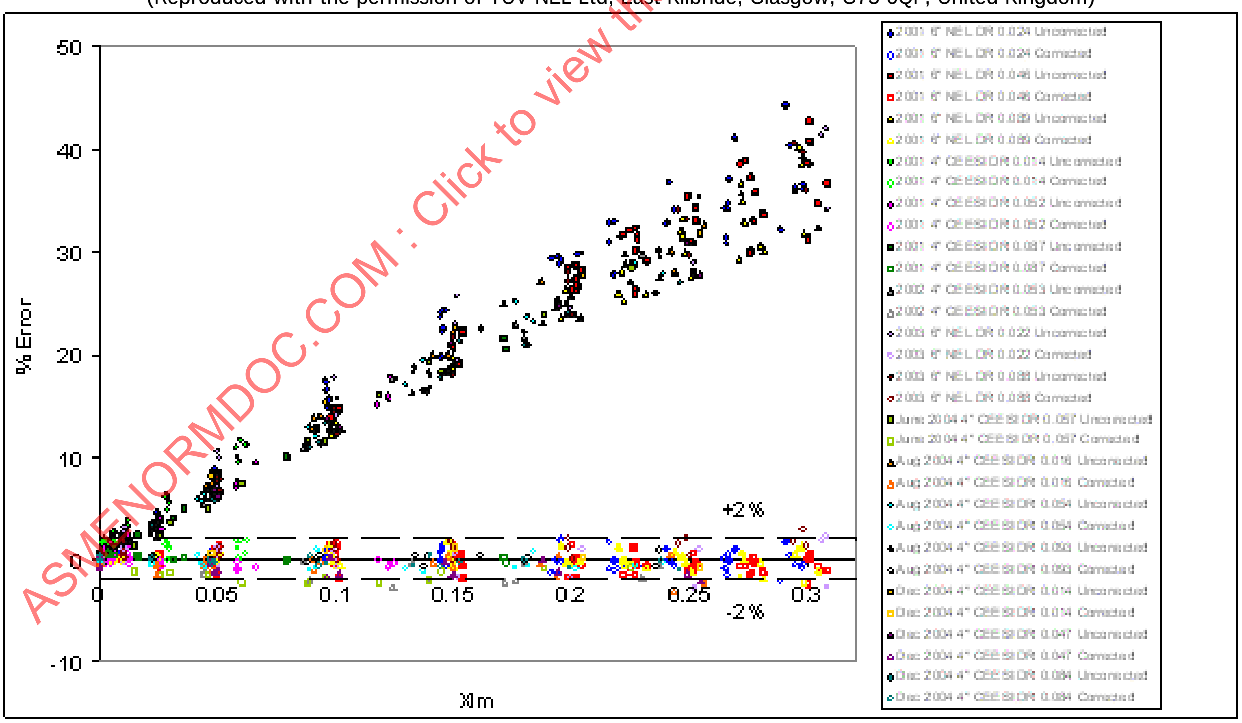
Figure G-7 (from Steven [15]) shows the 0.75 beta ratio correlation applied to two NEL 6 in., Schedule 80, 0.75 beta ratio data sets and four CEESI 4 in. Schedule 80, 0.75 beta ratio sets. This paper also showed this correlation correcting a 6 in., Schedule 160, 0.75 beta ratio data set from K-Lab and a 2 in. Schedule 80, 0.75 beta ratio data set from CEESI for a known liquid flow rate to a gas flow rate uncertainty of ±2% with a few outliers. However, it must be understood that all check data (i.e. data not used in the creation of the correlation) had data ranges within the range of the data that was used to create the correlation. Therefore, extreme caution must be used when extrapolating this correlation for use out with the data sets, e.g. for use with higher gas to liquid density ratios or gas densiometric Froude numbers or different liquids, for, as with any "blind fit" correlation gross errors can potentially occur in this situation."

The 0.55 beta ratio cone-type DP meter correlation has not been further discussed since the initial release in 2002, except that the 0.55 beta ratio cone-type DP meter correlation was tested for repeatability [14]. It was generally considered repeatable, but there were some outliers at lower pressure and higher Lockhart–Martinelli parameters.

The 0.55 beta ratio cone type DP meter wet gas flow correlation published [12] is eq. (G-23) with eqs. (G-27) through (G-29).

Fig. G-7 NEL and CEESI 0.75 Beta Ratio Cone-Type DP Meter Data Corrected by Eq. (G-23) with Eqs. (G-24) to (G-26)

(Reproduced with the permission of TUV NEL Ltd, East Kilbride, Glasgow, G75 0QF, United Kingdom)



$$A = 1.224 + \frac{0.141}{\left(\frac{\rho_g}{\rho_l}\right)} \qquad \text{(G-27)}$$

$$B = -0.0334 - \frac{0.00139}{\left(\frac{\rho_g}{\rho_l}\right)} \qquad \text{(G-28)}$$

$$C = \sqrt{0.0805 + \left(\frac{0.0109}{\left(\frac{\rho_g}{\rho_l}\right)^2}\right)} \qquad \text{(G-29)}$$

No low pressure 0.55 beta ratio cone-type DP meter data is known to be in the public domain, and no low-pressure modification to this correlation is known to have been issued. Like all correlations, this data fit is useable and valuable within the data range of the data that was used to create it. However, again, it is not advisable to extrapolate this blind fit correlation.

The liquid mass flow rate is required as an input. The desired gas mass flow rate is derived by iteration. The initial gas mass flow rate input of the uncorrected meter reading usually ensures quick convergence of the iteration.

Steven claimed 2% uncertainty within the scope of the data set. No independent testing on this meter is known to have been published except for Fig. G-7. This Joint Industry Project (JIP) is known to have carried out substantial research into several meters' performance with wet gas flows including the cone type DP meter but as yet the bulk of this information is not in the public domain.

G-6 FOUR-HOLE ORIFICE PLATE METER WET GAS RESEARCH

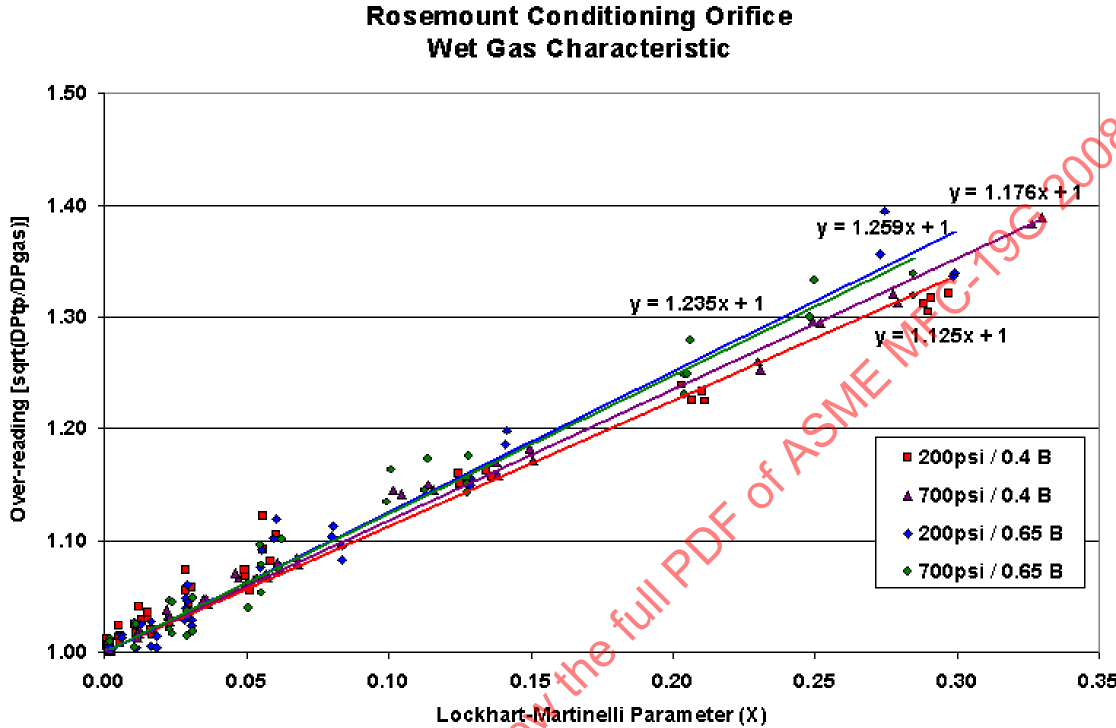
A recent entry into the flow metering market is a patented modified orifice plate type DP meter where the plate has four small holes, on an x-y axis through the center of the plate (i.e., flow centerline) all equidistant from the center, in place of the conventional one larger central hole. The meter operates in the same way as all DP meters (that is, the single-phase flow rate is found from eq. G-1).

In 2005 a paper discussing this four-hole orifice plate meter's response to wet gas flow tests at CEESI was released by Evans [61]. The orientation of the holes during the tests were 45 deg rotated from normal dry gas operation to allow one hole to be as low in the pipe work as possible in order to reduce the liquid dam effect orifice plates are commonly thought to have with wet gas flows.

Four 3 in., Schedule 40, 0.40 and 0.65 beta ratio meters were tested at CEESI with natural gas and decane at 200 psia and 700 psia across a Lockhart–Martinelli parameter range of $0 \le X_{LM} \le 0.3$. Each beta ratio had one meter with wall tappings and one meter with flange tappings. As found by Schuster [4] for standard orifice plate meters no difference in wet gas performance was observed with the tapping positions for the four-hole orifice plate. Therefore, the wall and flange pressure tap results for each beta are grouped together here as one meter result. The gas densiometric Froude number range was not discussed.

The general effect of the gas being wet was the same as for all DP meters in that the greater the Lockhart–Martinelli the greater the overreading. For a given set of wet gas flow conditions, the overreading to Lockhart–Martinelli parameter gradient was reported to be low compared to all other DP meters with recorded wet gas performance data. An unexpected reported finding with the four-hole orifice plate is that the higher beta ratio had the higher overreading. This is in reverse to all other publicly known DP meter wet gas flow results. Even though two separate pressures (or gas-to-liquid density

Fig. G-8 The Four-Hole Orifice Plate Meter Wet Gas Flow Results



ratios) were tested, no clear gas-to-liquid density ratio effect was reported. The differences in the pressure results are so small they could be accounted for by normal experimental data set scatter.

Figure G-8 shows the data reported. This figure shows original data presented in several graphs [61], but for this Report, the work has been updated to show all the relevant data on one graph, and the linear fits have been recalculated to adhere to the theoretical requirement of no liquid induced overreading with dry gas flow.

Therefore, the reported four-holed orifice plate meter wet gas correlations were as follows:

For a 0.40 beta ratio meter at 200 psi

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1+1.125X_{LM}} \tag{G-30}$$

For a 0.40 beta ratio meter at 700 psi

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1+1.176X_{LM}} \tag{G-31}$$

For a 0.65 beta ratio meter at 200 psi

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1+1.259X_{LM}} \tag{G-32}$$

For a 0.40 beta ratio meter at 700 psi

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{1+1.235 X_{LM}} \qquad \text{(G-33)}$$

Note that in effect this data has been developed into a wet gas flow correlation for the individual beta ratio four-hole orifice plate meters by analyzing and fitting the data to eq. (G-19) as Jamieson [70] suggested for different geometry Venturi meters in 1993.

Fig. G-9 The Performance of the Four-Hole Orifice Plate Murdock Type Correlations

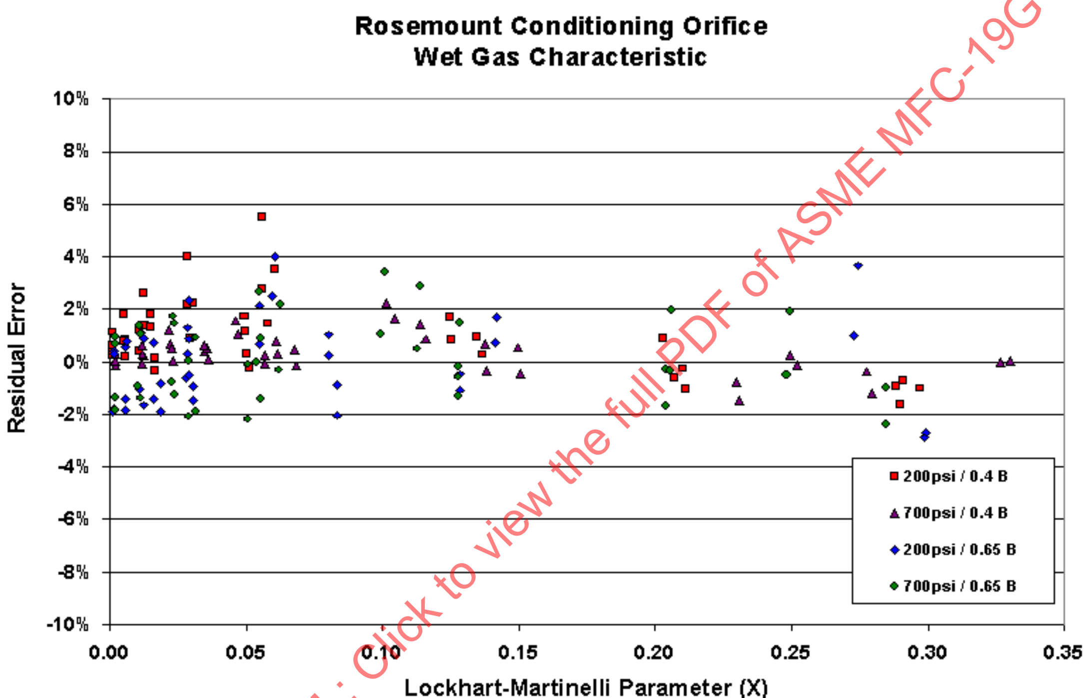


Figure G-9 shows the performance of applying eqs. (G-30) through (G-33) for the known (reference meter supplied) liquid flow rate. The linear fits have a 4% uncertainty.

Note that the liquid mass flow rate or Lockhart–Martinelli parameter is required as an input. The desired gas mass flow rate is derived by iteration. The initial gas mass flow rate input of the uncorrected meter reading usually ensures quick convergence of the iteration. Evans [61] showed graphs of "% Offset" vs. Lockhart–Martinelli parameter and these indicate that the correction uncertainties are 3% within the scope of the data set with a few outliers. No independent testing on this meter is known to have been published.

G-7 COMMENTS ON THE INDUSTRIAL USE OF DP METER WET GAS CORRELATIONS

Wet gas flow metering technology is a developing technology. Research on this topic is continuing, and knowledge on this topic is being gained at a steady rate, but much remains to be discovered and understood. Wet gas flow meters are developed and tested at wet gas flow test facilities or in field tests whenever an opportunity arises where the liquid and gas phases can be metered at some other location to a given uncertainty. This leaves the problem of whether these correlations can be extrapolated for use in any given situation. Very little information (e.g. Steven [20]) is given in the literature on whether individual DP meter wet gas correlations can be extrapolated. If so, to what range?

Currently, when an orifice plate meter is being used with a wet gas flow, the operator has a limited choice of wet gas orifice plate meter correlations. These are generally created from limited and often untraceable data sets, and little or no independent analysis on these wet gas correlations is available. The repeatability of these correlations is therefore largely unknown, as is the effect of extrapolating them to different flow conditions such as meter diameter, beta ratio, fluid types, gas and liquid flow rates, etc. Repeatability and suitability for extrapolation of various parameters when using orifice plate meter wet gas correlations is therefore an ongoing research topic.

For Venturi meters, the de Leeuw correlation for a 4 in. 0.40 beta ratio meter is reported by de Leeuw [48], Stewart [23], and Steven [20] to be reasonably repeatable within the parameters of the data set that created it. However, Stewart's work suggests extrapolations away from the beta ratio leads to significant errors if the 0.4 beta correlation is applied. Stewart showed that outside of the limited data set used to form the Steven 6 in. 0.55 beta ratio Venturi meter wet gas correlation, the correlation stopped being reliable. Steven [20] showed that when the same 4 in. 0.4 beta ratio Venturi meter was tested across a similar wet gas flow condition range in a different test facility, the wet gas results were the same as found by de Leeuw [10]. Steven went on to express concerns with the extrapolation of some wet gas correlations and gave theoretical reasons for the concern. Steven [19] discussed the possibility of a diameter effect on the wet gas response of a Venturi meter. That is, from a limited data set, it was shown that as the size of the Venturi meter being considered decreases for set values of Lockhart–Martinelli parameter, the gas-to-liquid density ratio, the gas densiometric Froude number, and liquid properties the magnitude of the overreading decreases. If this is confirmed, then DP meter wet gas correlations will be applicable to only set diameter meters just as they are currently known to be for set beta ratios. Also, it has been stated independently by Reader-Harris [17, 18] and Steven [19, 20] that Venturi meters show a dependence on the liquid properties for predicting the overreading for otherwise set flow conditions. Repeatability and suitability for extrapolation of various parameters when using Venturi wet gas correlations is therefore an ongoing research topic.

For cone-type DP meters, the Steven 0.75 beta ratio cone-type correlation is shown to be repeatable and some extrapolation of parameters can be permitted. However, the extrapolations are not extreme. The diameter range discussed was 2 in. to 6 in., and different liquids were used (which were mainly hydrocarbon liquids with some low pressure water test data points). The gas densiometric Froude number and maximum gas-to-liquid density ratio were not significantly extrapolated beyond the data set that created the correlation. The 0.55 beta ratio cone-type DP meter is said to be less repeatable than the repeatable 0.75 beta ratio cone-type DP meter. Repeatability and suitability for extrapolation of various parameters when using cone-type DP meter wet gas correlations is therefore an ongoing research topic.

A major problem in industry is the relatively common situation of real industrial wet gas flows having flow parameters outside of the available test facility capabilities. Higher gas-to-liquid density ratios, higher gas flow rates, and larger pipe diameters than any available test center can replicate are very common problems to the engineers trying to choose a wet gas metering system. It should be noted that there is no reason any user could not fit a unique wet gas correlation to any in situ meter if the situation arose where there was reliable reference data. This is occasionally done in the natural gas production industry when a DP meter in wet gas has a separator that allows for a given uncertainty of single-phase flows at the separator outlet. For DP meters that have flow conditions considerably outside the test facility's limits, this is a good way of reducing the uncertainty associated with an individual wet gas flow meter.

NONMANDATORY APPENDIX H[1]
ORIGINS OF THE EXISTING WET GAS FLOW DP METER CORRELATIONS

H-1 INTRODUCTION

There are several DP meter wet gas correlations used for wet gas metering and most users treat these correlations as black box technologies. However, the literature in the public domain does describe the development of many of these correlations including the assumptions made in any mathematical modeling and the limits of the data sets that were used. This appendix describes the development of these correlations to allow the user to have a better understanding of the abilities and limitation of each correlation so an informed choice can be made on which correlation to use with which DP meter for particular wet gas flow metering applications.

Research papers on the effect on gas meters of any entrained liquid started appearing in the late 1950s. Papers such as Schuster [4] discuss the scale of the liquid induced error on orifice plate meters and do state that a positive error is found that increases with the liquid loading. However, no correlations were given. At this time the only correction available was to assume the wet gas to be a homogeneous flow and carry out a homogeneous flow correction to predict the gas for a known liquid content. From the 1960s onwards correction factors were published to predict the scale of the overreading for a known liquid flow rate. There are also papers that discuss DP meter's performance with wet gas flows that do not offer correlations but do offer important findings (such as the beta ratio effect with wet gas flows). The correlations and then the other findings on DP meter wet gas performance are now discussed.

H-2 THE HOMOGENEOUS FLOW MODEL CORRELATION FOR GENERIC DP METERS

The homogeneous model correction factor for DP meters is unique in the following list of DP meter correction factors in as much as it is based purely on a set of assumptions and theory and no experimental data is involved.

The homogeneous flow model concept is to treat the two-phase flow as if it were a single-phase flow. In order to do this, the assumption must be made that the liquid and gas phases are perfectly mixed (i.e., that the liquid droplets are infinitely small and dispersed evenly throughout the gas phase). This assumption allows the two-phase flow to be treated as acting like a single-phase fluid with an averaged fluid density. This then allows the application of the single-phase DP meter equation with the averaged fluid density.

The algebraic steps required to implement the homogeneous model are as follows:

Let

$$v_{homogenous} = \frac{1}{\rho_{homogeneous}} \qquad \text{(H-1)}$$

where $v_{homogeneous}$ and $\rho_{homogeneous}$ are the specific volume and the density of the homogeneous mixture, respectively. Then:

$$v_{homogeneous} = \frac{V_{total}}{M_{total}} = \frac{V_l + V_g}{M_{total}} = \frac{V_l}{M_{total}} + \frac{V_g}{M_{total}} \qquad \text{(H-2)}$$

where V_{total} and M_{total} are the volume and mass of the homogeneous flow per unit time. V_l and V_g are the liquid and gas volumes per unit time, respectively. As we have the flow quality/dryness fraction definition [see eq. (16)] we have:

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

$$x = \frac{\dot{m}_g}{\dot{m}_g + \dot{m}_l} = \frac{\dot{m}_g}{\dot{m}_{total}} \qquad (9)$$

and

$$1 - x = \frac{\dot{m}_l}{\dot{m}_g + \dot{m}_l} = \frac{\dot{m}_l}{\dot{m}_{total}} \qquad \text{(H-3)}$$

For a unit time a set mass of liquid and gas flows (that is, $\dot{m}_g \equiv M_g$ and $\dot{m}_l \equiv M_l$). Therefore, the flow rate symbol can be dropped here. Substitution of eqs. (9) and (H-3) into eq. (H-2) for a unit of time therefore gives:

$$v_{homogeneous} = \frac{V_l}{M_{total}} + \frac{V_g}{M_{total}} = \frac{V_l}{\left(\frac{M_l}{1-x}\right)} + \frac{V_g}{\left(\frac{M_g}{x}\right)} \qquad \text{(H-4)}$$

that is,

$$v_{homogeneous} = x\frac{V_g}{M_g} + (1-x)\frac{V_l}{M_l} = \frac{1}{\rho_{homogeneous}} \qquad \text{(H-5)}$$

or,

$$\frac{1}{\rho_{homogeneous}} = \frac{x}{\rho_g} + \frac{1-x}{\rho_l} \qquad \text{(H-6)}$$

that is,

$$\rho_{homogeneous} = \frac{\rho_l \rho_g}{\rho_l x + \rho_g (1-x)} \qquad \text{(H-7)}$$

This density value should be applied to the single-phase DP meter equation along with the actual two-phase DP read. That is,

$$\dot{m}_{total} = \frac{\dot{m}_g}{x} = EA_t Y C_d \sqrt{2\rho_{homogeneous}\Delta P_{tp}} \qquad \text{(H-8)}$$

It should be noted in eq. (H-8) that the expansibility term Y and the discharge coefficient (C_d), if the discharge coefficient is a function of the homogeneous Reynolds number, are calculated for the homogeneous flows fluid properties. On rearranging eq. (H-8) and substituting in eq. (H-7):

$$\dot{m}_g = x\left(\frac{EA_t Y_g C_d \sqrt{2\rho_g \Delta P_{tp}}}{\sqrt{\frac{\rho_g}{\rho_l} + x\left(1 - \frac{\rho_g}{\rho_l}\right)}}\right) \qquad \text{(H-9)}$$

From eqs. (4) and (16) we can express the quality (x) in terms of the Lockhart–Martinelli parameter (X_{LM}). That is,

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} \qquad (4)$$

that is,

$$x = \frac{1}{1+\left\{X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}\right\}} \qquad (24)$$

Substituting eq. (24) into eq. (H-9) gives the homogeneous model in terms of the Lockhart–Martinelli parameter

$$\dot{m}_g = \frac{EA_tYC_d\sqrt{2\rho_g\Delta P_{tp}}}{\left(1+X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}\right)\sqrt{\frac{\rho_g}{\rho_l}+\frac{\left(1-\frac{\rho_g}{\rho_l}\right)}{\left(1+X_{LM}\sqrt{\frac{\rho_l}{\rho_g}}\right)}}} \qquad \text{(H-10)}$$

Rearranging gives

$$\dot{m}_g = \frac{EA_tYC_d\sqrt{2\rho_g\Delta P_{tp}}}{\sqrt{1+\left\{\sqrt{\frac{\rho_g}{\rho_l}}+\sqrt{\frac{\rho_l}{\rho_g}}\right\}X_{LM}+X_{LM}^2}} = \frac{\dot{m}_{g\,Apparent}}{\sqrt{1+CX_{LM}+X_{LM}^2}} \qquad \text{(G-5)}$$

where

$$C = \sqrt{\frac{\rho_g}{\rho_l}}+\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-11)}$$

that is,

$$C=\left(\frac{\rho_g}{\rho_l}\right)^n+\left(\frac{\rho_l}{\rho_g}\right)^n \qquad \text{(H-12)} \quad \text{and} \quad n=\frac{1}{2}.$$

That is,

$$\dot{m}_g = \frac{\dot{m}_{g,Apparent}}{\sqrt{1+\left[\left\{\sqrt{\frac{\rho_g}{\rho_l}}+\sqrt{\frac{\rho_l}{\rho_g}}\right\}\left(\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}\right)\right]+\left[\left(\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}\right)^2\right]}} \quad \text{(H-13)}$$

Equation (H-13) is the homogeneous model for all DP meters. Note that in order for the gas mass flow to be derived by iteration of this equation information on the liquid mass flow rate must be known from an external source. Note that

$$C_d = f\left(\mathrm{Re}_{homogenous}\right) = f\left(\frac{4\dot{m}_{total}}{\pi\mu_{homogenous}D}\right) \quad \text{(H-14)}$$

The viscosity term in the Reynolds number equation represents the viscosity of the homogeneous fluid. There is no generally agreed method for predicting this but common methods are

(a) McAdams

$$\frac{1}{\mu_{homogeneous}} = \frac{x}{\mu_{gas}} + \frac{1-x}{\mu_{liquid}}$$

(b) Cicchitti

$$\mu_{homogeneous} = x\mu_{gas} + (1-x)\mu_{liquid}$$

(c) Duckler et al.

$$\mu_{homogeneous} = (GVF)\mu_{gas} + (1-GVF)\mu_{liquid}$$

Note for the Duckler et al. homogeneous viscosity prediction method that by the definition of homogeneous two-phase flow there is no slip (s = 1) and the GVF is therefore here equivalent to the actual ratio of gas to pipe volume (i.e., the "Gas Void Fraction") at any instant in time. Also note that when the gas is dry [i.e., the quality (*x*) is unity] the homogeneous equations reduce to the single-phase equations.

H-3 THE MURDOCK CORRELATION FOR ORIFICE PLATE METERS

In 1962 Murdock [5] published a correlation for correcting the error in a gas flow measurement given by an orifice plate meter due to the presence of liquid in the gas flow.

The Murdock correlation is given as eq. (H-15).

$$\dot{m}_g = \frac{EA_tC_{d_g}Y\sqrt{2\rho_g\Delta P_{tp}}}{1+1.26X_{LM}} \quad \text{(G-6)}$$

The derivation of the Murdock equation is now given:

Murdock [5] presented the derivation in English units. Murdock considered the horizontal two-phase flow to be always in a separated flow pattern. Each phase was treated separately. The standard DP meter equation in English unit terms is:

$$\dot{\mathrm{w}} = \mathrm{EA_tYC_d}\sqrt{2\mathrm{g}\Delta\mathrm{P}\gamma} \quad \text{(H-15)}$$

where

$$\gamma = \rho g \tag{H-16}$$

and *Y* is unity for liquids (as liquids are considered in practical terms to be incompressible).

Murdock considered the equations required if each phase flowed alone through the meter. Murdock used the following standard gas weight flow equation if the gas flowed alone:

$$\dot{w}_g = EA_t YC_{dg}\sqrt{2g\Delta P_g \gamma_g} = A_t\left(K_g Y\right)\sqrt{2g\Delta P_g \gamma_g} \tag{H-17}$$

where

$$\gamma_g = \rho_g g \tag{H-18}$$

$$K_g = EC_{dg} \tag{H-19}$$

and C_{dg} is the discharge coefficient at the gas Reynolds number if the gas flowed alone. Murdock used the following standard liquid weight flow equation if the gas flowed alone:

$$\dot{w}_l = EA_t C_{dl}\sqrt{2g\Delta P_l \gamma_l} = A_t K_l \sqrt{2g\Delta P_l \gamma_l} \tag{H-20}$$

where

$$\gamma_l = \rho_l g \tag{H-21}$$

$$K_l = EC_{dl} \tag{H-22}$$

and C_{dl} is the discharge coefficient at the liquid Reynolds number if the liquid flowed alone.

Murdock assumed for the stratified flow model both phases have the same pressure drop (i.e., ΔP_{tp}) and that the two phases could have individual flow equations with unique discharge coefficients. Note that the orifice plate gas expansibility value would also be altered with two-phase flow. Therefore, with separated flow Murdock assumed there could be equations [i.e., (H-23) and (H-24)] for each phase:

$$\dot{w}_l = EA_{t_l} C_{dl_{tp}}\sqrt{2g\Delta P_{tp}\gamma_l} = A_{t_l}\left(K_l\right)_{tp}\sqrt{2g\Delta P_{tp}\gamma_l} \tag{H-23}$$

$$\dot{w}_g = EA_{t_g}\left(YC_d\right)_{tp}\sqrt{2g\Delta P_{tp}\gamma_g} = A_{t_g}\left(K_g Y\right)_{tp}\sqrt{2g\Delta P_{tp}\gamma_g} \tag{H-24}$$

where

$$\left(K_l\right)_{tp} = EC_{dl_{tp}} \tag{H-25}$$

and

$$\left(K_g Y\right)_{tp} = EYC_{dg_{tp}} \tag{H-26}$$

$C_{dl\,tp}$ and $C_{dg_{tp}}$ are the discharge coefficients for the liquid and gas phases in stratified two-phase flow.

Note that the total cross-sectional area of the orifice plate throat is the sum of the phase throat cross-sectional areas:

$$A_t = A_{tl} + A_{tg} \tag{H-27}$$

Therefore, substituting eqs. (H-23) and (H-24) into eq. (H-27) gives eq. (H-28).

$$A_t = \frac{\dot{w}_l}{(K_l)_{tp}\sqrt{2g\gamma_l \Delta P_{tp}}} + \frac{\dot{w}_g}{(K_g Y)_{tp}\sqrt{2g\gamma_g \Delta P_{tp}}} \qquad \text{(H-28)}$$

Substituting eqs. (H-17) and (H-20) into eq. (H-28) gives eq. (H-29):

$$A_t = \frac{A_t K_l \sqrt{2g\gamma_l \Delta P_l}}{(K_l)_{tp}\sqrt{2g\gamma_l \Delta P_{tp}}} + \frac{A_t K_g Y\sqrt{2g\gamma_g \Delta P_g}}{(K_g Y)_{tp}\sqrt{2g\gamma_g \Delta P_{tp}}} \qquad \text{(H-29)}$$

Equation (H-29) can be reduced to eq. (H-30):

$$1 = \left(\frac{K_l}{(K_l)_{tp}}\right)\sqrt{\frac{\Delta P_l}{\Delta P_{tp}}} + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right)\sqrt{\frac{\Delta P_g}{\Delta P_{tp}}} \qquad \text{(H-30)}$$

Equation (H-30) further reduces to eq. (H-31).

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = \left(\frac{K_l}{(K_l)_{tp}}\right)\sqrt{\frac{\Delta P_l}{\Delta P_g}} + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right) \qquad \text{(H-31)}$$

or by rearranging:

$$\sqrt{\Delta P_g} = \frac{\sqrt{\Delta P_{tp}}}{\left(\frac{K_l}{(K_l)_{tp}}\right)\sqrt{\frac{\Delta P_l}{\Delta P_g}} + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right)} \qquad \text{(H-32)}$$

which when considering eqs. (H-17) and (H-20) can be written as:

$$\sqrt{\Delta P_g} = \frac{\sqrt{\Delta P_{tp}}}{\left(\frac{K_l}{(K_l)_{tp}}\right)\left(\frac{\dot{w}_l}{A_t K_l \sqrt{2g\gamma_l}}\right)\left(\frac{A_t (K_g Y)\sqrt{2g\gamma_g}}{\dot{w}_g}\right) + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right)} \qquad \text{(H-33)}$$

i.e.,

$$\sqrt{\Delta P_g} = \frac{\sqrt{\Delta P_{tp}}}{\left(\frac{K_g Y}{(K_l)_{tp}}\right)\left(\frac{\dot{w}_l}{\dot{w}_g}\right)\sqrt{\frac{\gamma_g}{\gamma_l}} + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right)} \qquad \text{(H-34)}$$

Equation (H-30) can be expressed as eq. (H-35):

$$1 = \left(\frac{K_l}{(K_l)_{tp}}\right)\sqrt{\frac{\Delta P_l}{\Delta P_{tp}}} + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right)\sqrt{\frac{\Delta P_g}{\Delta P_{tp}}} \qquad \text{(H-30)}$$

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_l}} = \left(\frac{K_l}{(K_l)_{tp}}\right) + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right)\sqrt{\frac{\Delta P_g}{\Delta P_l}} \qquad \text{(H-35)}$$

or by rearranging to eq. (H-36):

$$\sqrt{\Delta P_l} = \frac{\sqrt{\Delta P_{tp}}}{\left(\frac{K_l}{(K_l)_{tp}}\right) + \frac{K_g Y}{(K_g Y)_{tp}}\sqrt{\frac{\Delta P_g}{\Delta P_l}}} \qquad \text{(H-36)}$$

Which can be expressed as eq. (H-37):

$$\sqrt{\Delta P_l} = \frac{\sqrt{\Delta P_{tp}}}{\left(\frac{K_l}{(K_l)_{tp}}\right) + \frac{K_g Y}{(K_g Y)_{tp}}\left(\frac{\dot{w}_g}{A_t K_g Y\sqrt{2g\gamma_g}}\right)\left(\frac{A_t K_l\sqrt{2g\gamma_l}}{\dot{w}_l}\right)} \qquad \text{(H-37)}$$

or in another form as eq. (H-38):

$$\sqrt{\Delta P_l} = \frac{\sqrt{\Delta P_{tp}}}{\left(\frac{K_l}{(K_l)_{tp}}\right) + \frac{K_l}{(K_g Y)_{tp}}\frac{\dot{w}_g}{\dot{w}_l}\sqrt{\frac{\gamma_l}{\gamma_g}}} \qquad \text{(H-38)}$$

Total weight flow is

$$\dot{w} = \dot{w}_l + \dot{w}_g \qquad \text{(H-29)}$$

Therefore, substituting eqs. (H-17) and (H-20) into eq. (H-29):

$$\dot{w} = A_t K_l\sqrt{2g\gamma_l \Delta P_l} + A_t K_g Y\sqrt{2g\gamma_g \Delta P_g} \qquad \text{(H-39)}$$

or by rearranging

$$\dot{w} = A_t\sqrt{2g}\left(K_l\sqrt{\gamma_l \Delta P_l} + K_g Y\sqrt{\gamma_g \Delta P_g}\right) \qquad \text{(H-40)}$$

Substituting eqs. (H-34) and (H-38) into eq. (H-40) gives eq. (H-41):

$$\dot{w} = A_t\sqrt{2g\Delta P_{tp}}\left(\frac{K_l\sqrt{\gamma_l}}{\left(\frac{K_l}{(K_l)_{tp}}\right)+\frac{K_l}{(K_gY)_{tp}}\frac{\dot{w}_g}{\dot{w}_l}\sqrt{\frac{\gamma_l}{\gamma_g}}}+\frac{K_gY\sqrt{\gamma_g}}{\left(\frac{K_gY}{(K_gY)_{tp}}\right)+\left(\frac{K_gY}{(K_l)_{tp}}\right)\frac{\dot{w}_l}{\dot{w}_g}\sqrt{\frac{\gamma_g}{\gamma_l}}}\right) \quad \text{(H-41)}$$

or by rearranging

$$\dot{w} = A_t\sqrt{2g\Delta P_{tp}}\left(\frac{K_l\sqrt{\gamma_l}}{\frac{K_l\,\dot{w}_g\sqrt{\gamma_l}(K_l)_{tp}+K_l(K_gY)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}{(K_gY)_{tp}(K_l)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}}+\frac{K_gY\sqrt{\gamma_g}}{\frac{(K_gY)\dot{w}_l\sqrt{\gamma_g}(K_gY)_{tp}+(K_gY)(K_l)_{tp}\,\dot{w}_g\sqrt{\gamma_l}}{(K_l)_{tp}\,\dot{w}_g\sqrt{\gamma_l}(K_gY)_{tp}}}\right) \quad \text{(H-42)}$$

And with further rearranging

$$\dot{w}=A_t\sqrt{2g\Delta P_{tp}}\left(\frac{\sqrt{\gamma_l}(K_gY)_{tp}(K_l)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}{\dot{w}_g\sqrt{\gamma_l}(K_l)_{tp}+(K_gY)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}+\frac{\sqrt{\gamma_g}(K_l)_{tp}\,\dot{w}_g\sqrt{\gamma_l}(K_gY)_{tp}}{\dot{w}_l\sqrt{\gamma_g}(K_gY)_{tp}+(K_l)_{tp}\,\dot{w}_g\sqrt{\gamma_l}}\right) \quad \text{(H-43)}$$

i.e.,

$$\dot{w}=A_t(K_gY)_{tp}\sqrt{2g\Delta P_{tp}\gamma_g}\left(\frac{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l}{\dot{w}_g\sqrt{\gamma_l}(K_l)_{tp}+(K_gY)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}+\frac{(K_l)_{tp}\,\dot{w}_g\sqrt{\gamma_l}}{\dot{w}_l\sqrt{\gamma_g}(K_gY)_{tp}+(K_l)_{tp}\,\dot{w}_g\sqrt{\gamma_l}}\right) \quad \text{(H-44)}$$

i.e.,

$$\dot{w} = A_t(K_gY)_{tp}\sqrt{2g\Delta P_{tp}\gamma_g}\left(\frac{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l+\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_g}{\dot{w}_g\sqrt{\gamma_l}(K_l)_{tp}+(K_gY)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}\right) \quad \text{(H-45)}$$

Dividing the numerator and denominator by $\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l$ gives eq. (H-46).

$$\dot{w} = A_t(K_gY)_{tp}\sqrt{2g\Delta P_{tp}\gamma_g}\left(\frac{\frac{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l}{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l}+\frac{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_g}{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l}}{\frac{\dot{w}_g\sqrt{\gamma_l}(K_l)_{tp}}{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l}+\frac{(K_gY)_{tp}\,\dot{w}_l\sqrt{\gamma_g}}{\sqrt{\gamma_l}(K_l)_{tp}\,\dot{w}_l}}\right) \quad \text{(H-46)}$$

This reduces to

$$\dot{w} = A_t (K_g Y)_{tp} \sqrt{2g\Delta P_{tp} \gamma_g} \left(\frac{1 + \frac{\dot{w}_g}{\dot{w}_l}}{\frac{\dot{w}_g}{\dot{w}_l} + \frac{(K_g Y)_{tp} \sqrt{\gamma_g}}{\sqrt{\gamma_l} (K_l)_{tp}}} \right) \qquad \text{(H-47)}$$

Multiplying the numerator and denominator by $\frac{\dot{w}_l}{\dot{w}_g}$ then gives

$$\dot{w} = A_t (K_g Y)_{tp} \sqrt{2g\Delta P_{tp} \gamma_g} \left(\frac{1 + \frac{\dot{w}_l}{\dot{w}_g}}{1 + \frac{(K_g Y)_{tp}}{(K_l)_{tp}} \frac{\dot{w}_l}{\dot{w}_g} \sqrt{\frac{\gamma_g}{\gamma_l}}} \right) \qquad \text{(H-48)}$$

Murdock defined the ratio of the liquid weight flow rate to the total weight flow rate as *y*. That is

$$y = \frac{\dot{w}_l}{\dot{w}_g + \dot{w}_l} \qquad \text{(H-49)}$$

Therefore

$$\frac{y}{1-y} = \frac{\dot{w}_l}{\dot{w}_g} \qquad \text{(H-50)}$$

Equation (H-48) can be rewritten as

$$\dot{w} = A_t (K_g Y)_{tp} \sqrt{2g\Delta P_{tp} \gamma_g} \left(\frac{1}{(1-y) + \frac{(K_g Y)_{tp}}{(K_l)_{tp}} y \sqrt{\frac{\gamma_g}{\gamma_l}}} \right) \qquad \text{(H-51)}$$

With eq. (H-51) the values of $(K_l)_{tp}$ and $(K_g Y)_{tp}$ are experimental unknowns. Murdock fitted the experimental data to eq. (H-31). That is:

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = \left(\frac{K_l}{(K_l)_{tp}} \right) \sqrt{\frac{\Delta P_l}{\Delta P_g}} + \left(\frac{K_g Y}{(K_g Y)_{tp}} \right) \qquad \text{(H-31)}$$

A reproduction of the linear line drawn by Murdock on a $\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}}$ vs. $\sqrt{\frac{\Delta P_l}{\Delta P_g}}$ plot is given in Fig. 6.1.1-2. Note from Nonmandatory Appendix A that $\sqrt{\frac{\Delta P_l}{\Delta P_g}} \approx X_{LM}$ for cases where the discharge coefficient and expansibility product for when the gas flows alone and the discharge coefficient for when the liquid flows alone are similar. The linear line is

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = M\sqrt{\frac{\Delta P_l}{\Delta P_g}} + C$$

where M is the gradient of the linear line fitted to the $\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}}$ vs. $\sqrt{\frac{\Delta P_l}{\Delta P_g}}$ plot and C is the line constant (i.e., where it cuts the ordinate). Murdock found from experiment that eq. (H-31) could be written as

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} \approx 1.26\sqrt{\frac{\Delta P_l}{\Delta P_g}} + 1 \qquad \text{(H-52)}$$

Therefore

$$\left(\frac{K_l}{(K_l)_{tp}}\right) = M \approx 1.26$$

and

$$\left(\frac{K_g Y}{(K_g Y)_{tp}}\right) = C \approx 1$$

In other words,

$$K_l \approx 1.26(K_l)_{tp} \quad \text{and} \quad K_g Y \approx (K_g Y)_{tp}$$

So eq. (H-51) becomes eq. (H-53):

$$\dot{w} = A_t(K_g Y)\sqrt{2g\Delta P_{tp}\gamma_g}\left(\frac{1}{(1-y) + 1.26\frac{K_g Y}{K_l} y\sqrt{\frac{\gamma_g}{\gamma_l}}}\right) \qquad \text{(H-53)}$$

In mass terms

$$\dot{m}_t = A_t(K_g Y)\sqrt{2\Delta P_{tp}\rho_g}\left(\frac{1}{(1-y) + 1.26\frac{K_g Y}{K_l} y\sqrt{\frac{\rho_g}{\rho_l}}}\right) \qquad \text{(H-54)}$$

Noting that

$$y = \frac{\dot{w}_l}{\dot{w}_g + \dot{w}_l} = \frac{g\dot{m}_l}{g\left(\dot{m}_g + \dot{m}_l\right)} = 1 - x$$

and

$$1 - y = \frac{\dot{w}_g}{\dot{w}_g + \dot{w}_l} = \frac{g\dot{m}_g}{g\left(\dot{m}_g + \dot{m}_l\right)} = x$$

where *x* is quality/dryness fraction [see eq.(16)] then eq. (H-54) can be rearranged to give

$$\dot{m}_t = A_t\left(K_g Y\right)\sqrt{2\rho_g \Delta P_{tp}}\left(\frac{1}{x + 1.26\frac{K_g Y}{K_l}(1-x)\sqrt{\frac{\rho_g}{\rho_l}}}\right) \quad \text{(H-55)}$$

Or rearranging to express the gas mass flow rate

$$\dot{m}_t = \dot{m}_g + \dot{m}_l = A_t\left(K_g Y\right)\sqrt{2\rho_g \Delta P_{tp}}\left(\frac{1}{\frac{\dot{m}_g}{\dot{m}_g + \dot{m}_l} + 1.26\frac{K_g Y}{K_l}\frac{\dot{m}_l}{\dot{m}_g + \dot{m}_l}\sqrt{\frac{\rho_g}{\rho_l}}}\right) \quad \text{(H-56)}$$

i.e.,

$$1 = A_t\left(K_g Y\right)\sqrt{2\rho_g \Delta P_{tp}}\left(\frac{1}{\dot{m}_g + 1.26\frac{K_g Y}{K_l}\dot{m}_l\sqrt{\frac{\rho_g}{\rho_l}}}\right) \quad \text{(H-57)}$$

$$\dot{m}_g = A_t\left(K_g Y\right)\sqrt{2\rho_g \Delta P_{tp}}\left(\frac{1}{\frac{\dot{m}_g + 1.26\frac{K_g Y}{K_l}\dot{m}_l\sqrt{\frac{\rho_g}{\rho_l}}}{\dot{m}_g}}\right) \quad \text{(H-58)}$$

$$\dot{m}_g = \frac{A_t(K_g Y)\sqrt{2\rho_g \Delta P_{tp}}}{1 + 1.26\frac{K_g Y}{K_l}\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(H-59)}$$

Equation (H-59) is the SI version of Murdock's final equation. However, note that a simplifying approximation can be made if it is assumed that $K_g Y \approx K_l$. Therefore (although Murdock did not state this), the equation could be (and usually is in recent times) approximated to eq. (H-60).

$$\dot{m}_g \approx \frac{A_t(K_g Y)\sqrt{2\rho_g \Delta P_{tp}}}{1 + 1.26\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(H-60)}$$

Noting eq. (4):

$$X_{LM} = \frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} \quad \text{(4)}$$

and that

$$K_g = EC_{dg} \quad \text{(H-19)}$$

the Murdock correlation can be written as considered to be eq. (H-61):

$$\dot{m}_g = \frac{EA_t C_d Y\sqrt{2\rho_g \Delta P_{tp}}}{1 + 1.26X_{LM}} = \frac{\dot{m}_{g,Apparent}}{1 + 1.26X_{Lm}} = \frac{\dot{m}_{g,Apparent}}{1 + 1.26\left(\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}}\right)} \quad \text{(H-61)}$$

H-4 THE JAMES CORRELATION FOR ORIFICE PLATE METERS

From experiment, the total mass flow of the saturated wet steam and the actual produced differential pressure were known. James then expressed these results in terms of a new density term denoted here as ρ_J where the subscript *J* (for James) indicates that this is a derived equivalent density that gives the correct total mass flow rate when applied to the standard DP meter equation for that flow condition. Therefore the standard equation is:

$$\dot{m}_{total} = EA_t Y C_d \sqrt{2\rho_J \Delta P_{tp}} \quad \text{(H-62)}$$

which lets the one unknown in the experimental data be found:

$$\rho_J = \left(\frac{\dot{m}_{total}}{EA_t Y C_d \sqrt{2\Delta P_{tp}}}\right)^2 \quad \text{(H-63)}$$

i.e.,

$$\nu_J = \left(\frac{EA_t YC_d \sqrt{2\Delta P_{tp}}}{\dot{m}_{total}} \right)^2 \qquad \text{(H-64)}$$

where ν_J is the specific volume associated with the equivalent density ρ_J. (That is, $\nu_J = 1/\rho_J$). From this value of density (ρ_J), James defined a new quality term by modifying eq. (H-6).

$$\frac{1}{\rho_{homogeneous}} = \frac{x}{\rho_g} + \frac{1-x}{\rho_l} = x\nu_g + (1-x)\nu_l \qquad \text{(H-6)}$$

where the specific gravity terms (ν) are the reciprocal of the density terms (ρ). That is $\nu_g = \frac{1}{\rho_g}$ and $\nu_l = \frac{1}{\rho_l}$. We now have

$$\nu_{homogeneous} = \frac{1}{\rho_{homogeneous}} = x\nu_g + (1-x)\nu_l \qquad \text{(H-65)}$$

where the specific gravity of the homogeneous flow ($\nu_{homogeneous}$) is the reciprocal of the homogeneous density ($\rho_{homogeneous}$). Taking the reciprocal of eq. (H-65):

$$\frac{1}{\nu_{homogeneous}} = \frac{1}{x\nu_g + (1-x)\nu_l} \qquad \text{(H-66)}$$

James modified the combination of eqs. (H-6) and (H-66) to:

$$\rho_J = \frac{1}{\nu_J} = \frac{1}{x_m\nu_g + (1-x_m)\nu_l} = \frac{\rho_g\rho_l}{x_m\rho_l + (1-x_m)\rho_g} \qquad \text{(H-67)}$$

where x_m is a modified quality parameter. In other words:

$$x_m = \frac{\nu_J - \nu_l}{\nu_g - \nu_l} \qquad \text{(H-68)}$$

Therefore, James could find the value of this modified x_m for each of the data points as all the parameters needed to produce the right hand side of the equation terms are known from the experiment data. James now plotted the conventional dryness fraction/quality value [eq. (16)] known from the wet steam 7.9 in. 0.707 and 0.837 beta ratio orifice meter experiments to this modified parameter, x_m.

James found by curve fitting

$$x_m = x^{1.5} \qquad \text{(H-69)}$$

Equation (H-62) leads to eq. (H-63).

$$\dot{m}_{total} = EA_t YC_d \sqrt{2\rho_J \Delta P_{tp}} \qquad \text{(H-62)}$$

$$\dot{m}_g = x\dot{m}_{total} = xEA_tYC_d\sqrt{2\rho_J \Delta P_{tp}} \qquad \text{(H-63)}$$

Substituting eq. (H-67) into eq. (H-63) gives:

$$\dot{m}_g = x\dot{m}_{total} = xEA_tYC_d\sqrt{2\left(\frac{\rho_g \rho_l}{x_m \rho_l + (1-x_m)\rho_g}\right)\Delta P_{tp}} \qquad \text{(H-70)}$$

and now substituting eq. (H-69) into eq. (H-70) gives:

$$\dot{m}_g = x\dot{m}_{total} = xEA_tYC_d\sqrt{2\left(\frac{\rho_g \rho_l}{x^{1.5} \rho_l + (1-x^{1.5})\rho_g}\right)\Delta P_{tp}} \qquad \text{(H-71)}$$

and rearranging gives:

$$\dot{m}_g = EA_tYC_d\sqrt{2\rho_g \Delta P_{tp}}\sqrt{\frac{x^2 \rho_l}{x^{1.5}\rho_l + (1-x^{1.5})\rho_g}} \qquad \text{(G-8)}$$

That is:

$$\dot{m}_g = \dot{m}_{g,Apparent}\sqrt{\frac{\left(\frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g}\right)^2 \rho_l}{\left(\frac{\dot{m}_g}{\dot{m}_l + \dot{m}_l}\right)^{1.5}\rho_l + \left(1 - \left(\frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g}\right)^{1.5}\right)\rho_g}} \qquad \text{(H-72)}$$

H-5 THE CHISHOLM CORRELATION FOR ORIFICE PLATE METERS

In 1967, Chisholm [7] published a general two-phase flow correlation for orifice plate meters. Chisholm assumed separated flow only and included the shear between the phases with such a flow pattern. Note that he did not state that the separated flow had to be stratified flow. Separated and stratified flow are often thought of as the same flow pattern but technically an annular flow with no mist (i.e., no entrained droplets) is also a separated flow. This point is of importance during the derivation of the Chisholm correlation.

In 1977 Chisholm published a research note [8] in which the general two-phase flow orifice plate meter correlation from 1967 was split into two sections. One correlation was for use at Lockhart–Martinelli parameters of less than unity and this therefore includes wet gas flow according to the ASME definition.

The Chisholm correlation development is considerably more detailed than Murdock development and the original paper [7] takes large algebraic steps that are not simple to follow. Furthermore, an assumption is made by Chisholm during the physical modeling that is not appropriate to many general wet gas metering situations. However, as will be shown now these assumptions prove to be unnecessary, as the end equation can be derived without the assumptions restrictions.

Chisholm starts by discussing the work of Jobson [63]. This work discusses compressible flow through orifices for fluid discharging from large vessels. Hence the upstream kinetic energy/velocity can be assumed negligible. Chisholm assumes that this is true for orifice plate meters in pipelines. This not appropriate in all industrial wet gas metering cases. As a consequence of this assumption, Chisholm defines the contraction coefficient, denoted as C_c, as

$$C_c = \left(\frac{A_{VC}}{A_N}\right)_{\text{Chisholm Notation}} \left(= \frac{A_2}{A_t}\right)_{\text{ASME Notation}} \quad \text{(H-73)}$$

where

A_N = cross-sectional area at orifice
A_{VC} = cross-sectional area at the vena contracta

A force balance then gave the equation:

$$(P_1 - P_2)A_N + F = \frac{W}{g}U \quad \text{(H-74)}$$

where

A_N = orifice area (denoted as A_t)
F = "force applied due to variation of pressure over upstream face of orifice"
g = gravitational constant
$P1$ = pressure at the orifice
$P2$ = pressure in the vena contracta
U = velocity at the vena contracta
W = weight flow

Note that Chisholm worked in U.S. Customary units. With the assumptions of negligible upstream kinetic energy Jobson had derived from dimensional considerations an expression for F. This was

$$F = \frac{fW^2}{g\rho A_N} \quad \text{(H-75)}$$

where

$$f = \frac{1}{C_c} - \frac{1}{2C_c^2} \quad \text{(H-76)}$$

Here it was decided to remove the assumption of negligible upstream kinetic energy, as it is not appropriate for all industrial applications and reevaluate the model accordingly[2].

Figure H-1 shows the single-phase model where the upstream kinetic energy is assumed relevant. For simplicity, the upstream cross-sectional area is denoted by subscript 1 and the vena contracta cross sectional area is from now denoted by subscript 2. A one-dimensional force balance along the x-axis gives eq. (C-5).

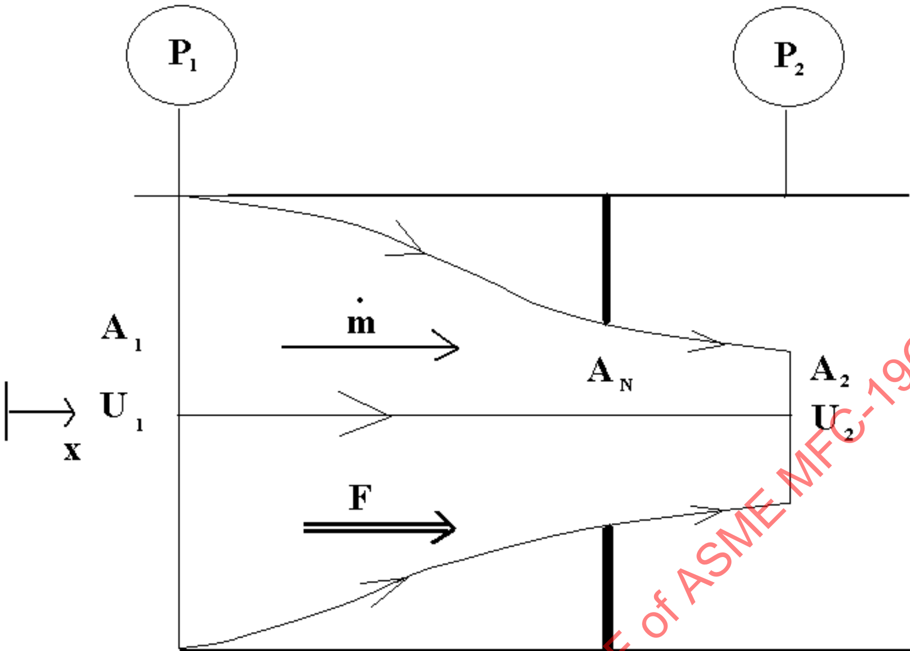
$$\sum F_x = \dot{m}(U_2 - U_1) \quad \text{(H-77)}$$

that is

$$P_1\vec{A}_1 + \vec{F}_w - P_2\overleftarrow{A}_2 = \dot{m}(U_2 - U_1) \quad \text{(H-78)}$$

[2] Dr. A. Gilchrist (retired) of Strathclyde University, Glasgow, UK is acknowledged as the original author of this updated derivation.

Fig H-1 Schematic Diagram of Flow Through Orifice Plate



where $\vec{F}_w$ is the force applied on the flowing fluid by the surfaces it has contact with (with a randomly assigned sign). Rearranging we get

$$(P_1 - P_2)A_2 + F_w + P_1(A_1 - A_2) = \dot{m}(U_2 - U_1) \qquad \text{(H-79)}$$

or if we let

$$F = F_w + P_1(A_1 - A_2) \qquad \text{(H-80)}$$

we get

$$(P_1 - P_2)A_2 + F = \dot{m}(U_2 - U_1) \qquad \text{(H-81)}$$

F in eq. (H-80) is the *F* defined by Chisholm in reference [7].

Now as

$$F_w = \int_{A_1}^{A_2} P dA \qquad \text{(H-82)}$$

an expression for pressure, *P*, in terms of the area, *A* is required. Chisholm assumed the flow to be effectively incompressible, as it was assumed that $P_1 - P_2 << P_1$. Therefore applying the Bernoulli/conservation of energy equation (for horizontal flow and energy losses due to shear forces assumed negligible) gives

$$\frac{P}{\rho} + \frac{U^2}{2} = \frac{P_2}{\rho} + \frac{U_2^2}{2} \qquad \text{(H-83)}$$

where the parameters with no subscript are conditions at arbitrary areas. Developing eq. (H-83), still for incompressible flow, gives

$$P = P_2 + \frac{\rho}{2}\left(U_2^2 - U^2\right) = P_2 + \frac{\rho U_2^2}{2}\left(1 - \left(\frac{U}{U_2}\right)^2\right) \quad \text{(H-84)}$$

From mass continuity

$$\dot{m} = \rho A U = \rho_1 A_1 U_1 = \rho A_2 U_2 \quad \text{(H-85)}$$

Therefore, we have

$$\frac{U}{U_2} = \frac{A_2}{A} \quad \text{(H-86)}$$

and

$$U_2 = \frac{\dot{m}}{\rho A_2} \quad \text{(H-87)}$$

Therefore

$$U_2^2 = \frac{1}{\rho^2}\left(\frac{\dot{m}}{A_2}\right)^2 \quad \text{(H-88)}$$

Therefore, substituting eqs. (H-86) and (H-88) into eq. (H-84) gives

$$P = P_2 + \frac{\rho U_2^2}{2}\left(1 - \left(\frac{U}{U_2}\right)^2\right) = P_2 + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left(1 - \left(\frac{A_2}{A}\right)^2\right) \quad \text{(H-89)}$$

Now, substituting eqs. (H-89) into eq. (H-82) gives

$$F_w = \int_{A_1}^{A_2}\left\{P_2 + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left(1 - \left(\frac{A_2}{A}\right)^2\right)\right\}dA \quad \text{(H-90)}$$

Developing this equation gives

$$F_w = P_2\int_{A_1}^{A_2} dA + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\int_{A_1}^{A_2} dA - \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2 A_2^2\int_{A_1}^{A_2}\frac{dA}{A^2} \quad \text{(H-91)}$$

i.e.,

$$F_w = P_2 A\Big|_{A_1}^{A_2} + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2 A\Big|_{A_1}^{A_2} - \frac{1}{2\rho}\dot{m}^2\left\{\frac{-1}{A}\right\}\Big|_{A_1}^{A_2} \quad \text{(H-92)}$$

i.e.,

$$F_w = P_2\left(A_2 - A_1\right) + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left(A_2 - A_1\right) + \frac{1}{2\rho}\dot{m}^2\left\{\frac{1}{A_2} - \frac{1}{A_1}\right\} \quad \text{(H-93)}$$

Therefore, substituting eq. (H-93) into eq. (H-80) we get

$$F = P_1\left(A_1 - A_2\right) + P_2\left(A_2 - A_1\right) + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left(A_2 - A_1\right) + \frac{1}{2\rho}\dot{m}^2\left\{\frac{1}{A_2} - \frac{1}{A_1}\right\} \qquad \text{(H-94)}$$

and now substituting eq. (H-89) into eq. (H-94) we get

$$F = \left(P_2 + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left(1 - \left(\frac{A_2}{A_1}\right)^2\right)\right)\left(A_1 - A_2\right) + P_2\left(A_2 - A_1\right) + \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left(A_2 - A_1\right) + \frac{1}{2\rho}\dot{m}^2\left\{\frac{1}{A_2} - \frac{1}{A_1}\right\} \qquad \text{(H-95)}$$

which reduces to

$$F = \frac{1}{2\rho}\dot{m}^2\left(\frac{1}{A_2} - \frac{1}{A_1}\right) - \frac{1}{2\rho}\left(\frac{\dot{m}}{A_2}\right)^2\left[\left[\left(\left(\frac{A_2}{A_1}\right)^2\left(A_1 - A_2\right)\right)\right]\right] \qquad \text{(H-96)}$$

which reduces further to

$$F = \frac{1}{2\rho}\frac{\dot{m}^2}{A_2}\left[1 - 2\frac{A_2}{A_1} + \left(\frac{A_2}{A_1}\right)^2\right] = \frac{\dot{m}^2}{\rho A_2}\left[\frac{1}{2} - \frac{A_2}{A_1} + \frac{1}{2}\left(\frac{A_2}{A_1}\right)^2\right] \qquad \text{(H-97)}$$

If we let $C_i = \frac{A_2}{A_1}$, then we have

$$F = \frac{\dot{m}^2}{\rho A_2}\left[\frac{1}{2} - C_i + \frac{1}{2}C_i^2\right] = f\,\frac{\dot{m}^2}{\rho A_2} \qquad \text{(H-98)}$$

where

$$f = \left[\frac{1}{2} - C_i + \frac{1}{2}C_i^2\right] \qquad \text{(H-99)}$$

Note that

$$C_i = \frac{A_2}{A_1} = \frac{d_2^2}{D_1^2} \qquad \text{(H-100)}$$

where

d_2 = diameter of the cross section at the vena contracta

D_1 = diameter at the inlet to the meter

In single-phase DP meter theory, the square root of the ratio of the upstream to throat area is defined as the beta ratio, β. With orifice plate meters the minimum flows cross sectional area is known as the vena contracta and its location downstream of the orifice is not precisely known due to it being dependent on the flow conditions. Therefore, as Chisholm was primarily discussing orifice plate meters but A_2 is the cross-sectional area at the vena contract and hence different to the orifice plate area it should be noted that $\beta \neq \sqrt{C_i}$. In fact, one of the factors that dictate the orifice plate meter discharge coefficient found from tables (derived from experimental data) is the correction for the known orifice cross sectional area being used in orifice plate meter flow equations instead of the actual unknown downstream cross sectional area. The reason for why this apparently irrelevant issue is discussed here will be evident later in the discussion of the Chisholm correlation derivation.

Note that this force balance derivation from first principles, which accounts for the upstream kinetic energy of the flow, has resulted in similar but different expressions than those obtained by Jobson [63], i.e., eq. (H-99) compared to eq. (H-76).

Now eq. (H-81) becomes eq. (H-101) with the substitution of eq. (H-98):

$$(P_1 - P_2)A_2 + f\frac{\dot{m}^2}{\rho A_2} = \dot{m}(U_2 - U_1) \qquad \text{(H-101)}$$

Here Chisholm assumed the flow pattern was separated flow. Each phase can therefore now be dealt with separately. The only indication of the existence of the other phase is the inclusion of the shear force, *S*, at the interface in the individual phase force balance equations. These are

For gas
$$(P_1 - P_2)A_{g2} + f\frac{\dot{m}_g^2}{\rho_g A_{g2}} - S = \dot{m}_g(U_{g2} - U_{g1}) \qquad \text{(H-102)}$$

For liquid
$$(P_1 - P_2)A_{l2} + f\frac{\dot{m}_l^2}{\rho_l A_{l2}} + S = \dot{m}_l(U_{l2} - U_{l1}) \qquad \text{(H-103)}$$

The shear force *S* sign is arbitrarily designated. In Chisholm's derivation it is assumed the upstream velocities are negligible (i.e., U_{g1} and U_{l1} are assumed to be zero). This cannot be considered appropriate for all industrial wet gas flow metering applications. The following derivation is Chisholm's derivation with the exception that the upstream kinetic energy has not been ignored. The end result is found to be the same.

Developing eq. (H-102)

$$(P_1 - P_2)A_{g2} - S = \dot{m}_g U_{g2} - \dot{m}_g U_{g1} - f\frac{\dot{m}_g^2}{\rho_g A_{g2}} \qquad \text{(H-104)}$$

and if the flow is considered effectively incompressible (i.e., the gas density change is small through the meter), mass continuity gives

$$\dot{m}_g = \rho_g A_{g1} U_{g1} = \rho_g A_{g2} U_{g2} \qquad \text{(H-105)}$$

and therefore we get

$$(P_1 - P_2)A_{g2} - S = \rho_g A_{g2} U_{g2}^2 - \rho_g A_{g1} U_{g1}^2 - f\frac{\rho_g^2 A_{g2}^2 U_{g2}^2}{\rho_g A_{g2}} \qquad \text{(H-106)}$$

Canceling terms and diving both sides of the equation by $\rho_g A_{g2}$ gives

$$\frac{(P_1 - P_2)}{\rho_g} - \frac{S}{\rho_g A_{g2}} = U_{g2}^2 - \frac{A_{g1}}{A_{g2}} U_{g1}^2 - fU_{g2}^2 \qquad \text{(H-107)}$$

Again from continuity, for an incompressible flow we know that

$$U_{g1}^2 = U_{g2}^2 \left(\frac{A_{g2}}{A_{g1}} \right)^2 \quad \text{(H-108)}$$

Therefore we have

$$\frac{(P_1 - P_2)}{\rho_g} - \frac{S}{\rho_g A_{g2}} = U_{g2}^2 - \frac{A_{g1}}{A_{g2}} U_{g2}^2 \left(\frac{A_{g2}}{A_{g1}} \right)^2 - fU_{g2}^2 = U_{g2}^2 \left[1 - \frac{A_{g2}}{A_{g1}} - f \right] \quad \text{(H-109)}$$

$$\frac{(P_1 - P_2)}{\rho_g} - \frac{S}{\rho_g A_{g2}} = \frac{U_{g2}^2}{2} \left[2 - 2\frac{A_{g2}}{A_{g1}} - 2f \right] \quad \text{(H-110)}$$

Chisholm makes the assumption that the ratio of upstream to vena contracta area is the same for single-phase and separated phase flows. Hence

$$C_i = \frac{A_2}{A_1} = \frac{A_{g2}}{A_{g1}} = \frac{A_{l2}}{A_{l1}} \quad \text{(H-111)}$$

Therefore

$$\frac{(P_1 - P_2)}{\rho_g} - \frac{S}{\rho_g A_{g2}} = \frac{U_{g2}^2}{2} [2 - 2C_i - 2f] \quad \text{(H-112)}$$

and with substitution of eq. (H-99) we get:

$$\frac{(P_1 - P_2)}{\rho_g} - \frac{S}{\rho_g A_{g2}} = \frac{U_{g2}^2}{2} [1 - C_i^2] \quad \text{(H-113)}$$

Letting

$$\Omega = 1 - C_i^2 = 1 - \left(\frac{A_2}{A_1} \right)^2 \quad \text{(H-114)}$$

we get

$$\frac{(P_1 - P_2)}{\rho_g} - \frac{S}{\rho_g A_{g2}} = \frac{U_{g2}^2}{2} \Omega \quad \text{(H-115)}$$

or

$$\frac{(P_1 - P_2)}{\rho_g \Omega} \left[1 - \frac{S}{A_{g2}(P_1 - P_2)} \right] = \frac{U_{g2}^2}{2} \quad \text{(H-116)}$$

Chisholm now defines a parameter called the shear force ratio. He used S_R to denote this which has the potential to be very confusing as ASME defines S_R as the slip ratio [see eq. (21)]. Therefore, here Chisholm's shear force ratio will be denoted by S_{FR}.

$$S_{FR} = \frac{S}{A_{g2}(P_1 - P_2)} \quad \text{(H-117)}$$

Therefore, we now have:

$$\frac{(P_1 - P_2)}{\rho_g \Omega}[1 - S_{FR}] = \frac{U_{g2}^2}{2} \qquad \text{(H-118)}$$

Taking the same approach with the liquid phase, we can develop a similar expression.

$$(P_1 - P_2)A_{l2} + f\frac{\dot{m}_l^2}{\rho_l A_{l2}} + S = \dot{m}_l(U_{l2} - U_{l1}) \qquad \text{(H-103)}$$

Developing eq. (H-103)

$$(P_1 - P_2)A_{l2} + S = \dot{m}_l U_{l2} - \dot{m}_l U_{l1} - f\frac{\dot{m}_l^2}{\rho_l A_{l2}} \qquad \text{(H-119)}$$

and for incompressible flow, mass continuity is given by

$$\dot{m}_l = \rho_l A_{l1} U_{l1} = \rho_l A_{l2} U_{l2} \qquad \text{(H-120)}$$

we get

$$(P_1 - P_2)A_{l2} + S = \rho_l A_{l2} U_{l2}^2 - \rho_l A_{l1} U_{l1}^2 - f\frac{\rho_l^2 A_{l2}^2 U_{l2}^2}{\rho_l A_{l2}} \qquad \text{(H-121)}$$

Canceling terms and dividing both sides of the equation by $\rho_l A_{l2}$ gives

$$\frac{(P_1 - P_2)}{\rho_l} + \frac{S}{\rho_l A_{l2}} = U_{l2}^2 - \frac{A_{l1}}{A_{l2}} U_{l1}^2 - fU_{l2}^2 \qquad \text{(H-122)}$$

Again from continuity for an incompressible flow we know that

$$U_{l1}^2 = U_{l2}^2 \left(\frac{A_{l2}}{A_{l1}}\right)^2 \qquad \text{(H-123)}$$

Therefore we have

$$\frac{(P_1 - P_2)}{\rho_l} + \frac{S}{\rho_l A_{l2}} = U_{l2}^2 - \frac{A_{l1}}{A_{l2}} U_{l2}^2 \left(\frac{A_{l2}}{A_{l1}}\right)^2 - fU_{l2}^2 = U_{l2}^2\left[1 - \frac{A_{l2}}{A_{l1}} - f\right] \qquad \text{(H-124)}$$

or in other words

$$\frac{(P_1 - P_2)}{\rho_l} + \frac{S}{\rho_l A_{l2}} = \frac{U_{l2}^2}{2}\left[2 - 2\frac{A_{l2}}{A_{l1}} - 2f\right] \qquad \text{(H-125)}$$

From eq. (H-111) we can say then that

$$\frac{(P_1 - P_2)}{\rho_l} + \frac{S}{\rho_l A_{l2}} = \frac{U_{l2}^2}{2}[2 - 2C_i - 2f] \qquad \text{(H-126)}$$

and again, as with the gas phase case, eq. (H-99) can be substituted in to give

$$\frac{(P_1 - P_2)}{\rho_l} + \frac{S}{\rho_l A_{l2}} = \frac{U_{l2}^2}{2}\left[1 - C_i^2\right] \qquad \text{(H-127)}$$

or with eq. (H-114) we get

$$\frac{(P_1 - P_2)}{\rho_l} + \frac{S}{\rho_l A_{l2}} = \frac{U_{l2}^2}{2}\Omega \qquad \text{(H-128)}$$

Or

$$\frac{(P_1 - P_2)}{\rho_l \Omega}\left[1 + \frac{S}{A_{l2}(P_1 - P_2)}\right] = \frac{U_{l2}^2}{2}$$

From eq. (H-117) it is found that

$$\frac{A_{g2}}{A_{l2}} S_{FR} = \frac{A_{g2}}{A_{l2}} \frac{S}{A_{g2}(P_1 - P_2)} = \frac{S}{A_{l2}(P_1 - P_2)} \qquad \text{(H-129)}$$

So we now have

$$\frac{(P_1 - P_2)}{\rho_l \Omega}\left[1 + \frac{A_{g2}}{A_{l2}} S_{FR}\right] = \frac{U_{l2}^2}{2} \qquad \text{(H-130)}$$

Chisholm denoted a slip ratio as *K*. This does not match the denotation in this Report for a slip ratio, which is S_R [see eq. (21)]. The definition in this Report is therefore used here. (Note this can be potentially confusing if reading Chisholm's paper [7], as there Chisholm denotes S_R as the shear force ratio that is denoted here as S_{FR}.) Let

$$S_R = \frac{U_g}{U_l} \qquad \text{(H-131)}$$

Therefore, by substituting eqs. (H-118) and (H-130) into eq. (H-131) for the case of the slip ratio at the vena contracta, we get

$$S_R = \frac{U_{g2}}{U_{l2}} = \sqrt{\frac{\frac{U_{g2}^2}{2}}{\frac{U_{l2}^2}{2}}} = \sqrt{\frac{\frac{(P_1 - P_2)}{\rho_g \Omega}(1 - S_{FR})}{\frac{(P_1 - P_2)}{\rho_l \Omega}\left(1 + \frac{A_{g2}}{A_{l2}} S_{FR}\right)}} = \sqrt{\frac{\rho_l}{\rho_g} \frac{1 - S_{FR}}{\left(1 + \frac{A_{g2}}{A_{l2}} S_{FR}\right)}} \qquad \text{(H-132)}$$

Chisholm defined a parameter called the Shear Force Function denoted by *Z*. It is given here as eq. (H-133).

$$Z = \sqrt{\frac{\left(1 + \frac{A_{g2}}{A_{l2}} S_{FR}\right)}{1 - S_{FR}}} \qquad \text{(H-133)}$$

Therefore, substituting eq. (H-133) into eq. (H-132) gives

$$S_R = \frac{U_{g2}}{U_{l2}} = \frac{1}{Z}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-134)}$$

Now again taking mass continuity eqs. (H-105) and (H-120) and rearranging we get

$$A_{g2} = \frac{\dot{m}_g}{\rho_g U_{g2}} \quad \text{and} \quad A_{l2} = \frac{\dot{m}_l}{\rho_l U_{l2}} \qquad \text{(H-135)}$$

Therefore,

$$\frac{A_{g2}}{A_{l2}} = \frac{\dot{m}_g}{\rho_g U_{g2}} \frac{\rho_l U_{l2}}{\dot{m}_l} = \frac{\rho_l}{\rho_g}\frac{\dot{m}_g}{\dot{m}_l}\frac{U_{l2}}{U_{g2}} \qquad \text{(H-136)}$$

Noting that $\dot{m}_g = x\dot{m}$ and $\dot{m}_l = (1-x)\dot{m}$ and substituting in eq. (H-134) to eq. (H-136) gives:

$$\frac{A_{g2}}{A_{l2}} = \frac{\rho_l}{\rho_g}\frac{x\dot{m}}{(1-x)\dot{m}} Z\sqrt{\frac{\rho_g}{\rho_l}} = Z\frac{x}{1-x}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-137)}$$

Chisholm defined the parameter *Y*[3] as

$$Y = \frac{x}{1-x}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-138)}$$

we have:

$$\frac{A_{g2}}{A_{l2}} = ZY \qquad \text{(H-139)}$$

Now Chisholm introduced “the basic relationship for flow of liquid alone through a sharp-edged orifice, if the upstream velocity is neglected.” The equation was given in U.S. Customary units as

$$W_{ls} = C_c A_N \sqrt{2g\rho_l (P_1 - P_2)} \qquad \text{(H-140)}$$

where W_{LS} is said by Chisholm to be the “Liquid flow rate by weight during single-phase flow with two-phase pressure drop.” Note here that eq. (H-140) assumes no energy losses and A_N is Chisholm’s notation for the orifice or “throat” area (which is denoted here as A_t). The parameter C_c is defined by Chisholm as

$$C_c = \left(\frac{A_{vc}}{A_N}\right)_{\text{Chisholm Notation}} \left(= \left(\frac{A_2}{A_t}\right)_{\text{ASME Notation}} \right) \qquad \text{(H-73)}$$

Note that the term $C_C A_N$ (H-140) is the vena contracta cross-sectional area (denoted as A_2), which is in practice unknown. Switching to using the known orifice area (denoted as A_t) requires a correction for the fact that this is larger than the unknown vena contracta cross sectional area. This is usually done by including this correction with the energy loss corrections (i.e., the discharge coefficient, which is found in tables derived from experiment). Therefore, the SI version of eq. (H-140) using notation in this Report

[3] Care must be taken to note that this parameter is not the expansion factor term in standard single-phase DP meter calculations that is also denoted in the US as “Y” (but ε in Europe).

includes the discharge coefficient and the orifice area as well as the velocity of approach term *E* as upstream kinetic energy is not always negligible.

$$\dot{m}_{ls} = EA_tC_d\sqrt{2\rho(P_1 - P_2)} = E^*A_2\sqrt{2\rho_l(P_1 - P_2)} \qquad \text{(H-141)}$$

Note that the traditionally used velocity of approach, *E*, is

$$E = \frac{1}{\sqrt{1-\left(\frac{A_t}{A_1}\right)^2}} \qquad \text{(H-142)}$$

and the velocity of approach where the upstream to vena contracta is

$$E^* = \frac{1}{\sqrt{1-\left(\frac{A_2}{A_1}\right)^2}} \qquad \text{(H-143)}$$

Note that eq. (H-141) is a standard single-phase equation for liquid flows through orifice plate meters.

Chisholm (who was dealing with general two-phase flow) developed the equations in terms of liquid flow and then simply gave the final result if the same procedure was done for gas. Here the gas equation is developed with the assumption of negligible upstream kinetic energy removed.

Assuming like Chisholm that the gas is effectively incompressible and there are no losses, but assuming unlike Chisholm that upstream gas velocity is not negligible, the single-phase gas flow rate ($\dot{m}_{gs}$) that would produce the differential pressure ($P_1 - P_2$) of an arbitrary wet gas flow is

$$\dot{m}_{gs} = EA_tC_d\sqrt{2\rho_g(P_1 - P_2)} = E^*A_2\sqrt{2\rho_g(P_1 - P_2)} \qquad \text{(H-144)}$$

Note that eq. (H-144) is actually the single-phase eq. (G-1) with the gas mass flow subscript having an added *s* to indicate single-phase flow in this two-phase derivation.

Taking eq. (H-118): $$\frac{(P_1 - P_2)}{\rho_g\Omega}[1 - S_{FR}] = \frac{U_{g2}^2}{2} \qquad \text{(H-118)}$$

and rearranging gives: $$U_{g2} = \sqrt{\frac{2(P_1 - P_2)}{\rho_g\Omega}[1 - S_{FR}]} \qquad \text{(H-145)}$$

Taking the mass continuity eq. (H-105): $$\dot{m}_g = \rho_g A_{g1} U_{g1} = \rho_g A_{g2} U_{g2} \qquad \text{(H-105)}$$

and substituting eq. (H-145) into eq. (H-105):

$$\dot{m}_g = \rho_g A_{g2}\sqrt{\frac{2(P_1 - P_2)}{\rho_g\Omega}(1 - S_{FR})} \qquad \text{(H-146)}$$

or

$$\dot{m}_g = A_{g2}\sqrt{\frac{2\rho_g(P_1 - P_2)}{\Omega}(1 - S_{FR})} \qquad \text{(H-147)}$$

Dividing the gas mass flow prediction if the wet gas differential pressure was read with a single-phase gas flow] i.e., eq. (H-144)] by the separated gas mass flow equation [i.e., eq. (H-147)] gives

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{E^* A_2\sqrt{2\rho_g(P_1 - P_2)}}{A_{g2}\sqrt{\frac{2\rho_g(P_1 - P_2)}{\Omega}(1 - S_{FR})}} = \frac{E^*\sqrt{\Omega}A_2}{A_{g2}\sqrt{1 - S_{FR}}} \qquad \text{(H.-148)}$$

Now, at the downstream flow's cross sectional area (A_2), the area is made up of a gas area (A_{g2}) and a liquid area (A_{l2}). That is

$$A_2 = A_{g2} + A_{l2} \qquad \text{(H-149)}$$

Substituting eq. (H-149) into eq. (H-148) gives

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{E^*\sqrt{\Omega}}{\sqrt{1 - S_{FR}}}\left(1 + \frac{A_{l2}}{A_{g2}}\right) \qquad \text{(H-150)}$$

Note that

$$\Omega = 1 - C_i^2 = 1 - \left(\frac{A_2}{A_1}\right)^2 \qquad \text{(H-114)}$$

and

$$E^* = \frac{1}{\sqrt{1 - \left(\frac{A_2}{A_1}\right)^2}} \qquad \text{(H-143)}$$

Therefore

$$E^*\sqrt{\Omega} = 1 \qquad \text{(H-151)}$$

and

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{1}{\sqrt{1 - S_{FR}}}\left(1 + \frac{A_{l2}}{A_{g2}}\right) \qquad \text{(H-152)}$$

The reduction of the term $E^*\sqrt{\Omega}$ to unity has in fact made Chisholm's hugely limiting assumption of negligible upstream velocity/kinetic energy superfluous. That is, it has been shown that there is no need for such a limiting assumption and, in fact, the Chisholm equation in reality has no such restriction.

Now, carrying on with the Chisholm derivation, we bring back eq. (H-133)

$$Z = \sqrt{\frac{\left(1 + \frac{A_{g2}}{A_{l2}}S_{FR}\right)}{1 - S_{FR}}} \qquad \text{(H-133)}$$

and rearranging gives

$$S_{FR} = \frac{Z^2 - 1}{Z^2 + \frac{A_{g2}}{A_{l2}}} \qquad \text{(H-153)}$$

which can be substituted into eq. (H-152):

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{\left(1 + \frac{A_{l2}}{A_{g2}}\right)}{\sqrt{1 - \left(\frac{Z^2 - 1}{Z^2 + \frac{A_{g2}}{A_{l2}}}\right)}} \qquad \text{(H-154)}$$

Noting that

$$\frac{A_{g2}}{A_{l2}} = ZY \qquad \text{(H-139)}$$

We have from eq. (H-154):

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{\left(1 + \frac{1}{ZY}\right)}{\sqrt{1 - \left(\frac{Z^2 - 1}{Z^2 + ZY}\right)}} \qquad \text{(H-155)}$$

i.e.,:

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{\left(\frac{ZY + 1}{ZY}\right)}{\sqrt{\frac{ZY + 1}{Z^2 + ZY}}} = \frac{(ZY + 1)\sqrt{(Z^2 + ZY)}}{ZY\sqrt{ZY + 1}} \qquad \text{(H-156)}$$

i.e.,

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \frac{\sqrt{(ZY + 1)}\sqrt{(Z^2 + ZY)}}{ZY} = \sqrt{\frac{(ZY + 1)(Z^2 + ZY)}{(ZY)^2}} \qquad \text{(H-157)}$$

i.e.,

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \sqrt{\frac{Z^3Y + Z^2Y^2 + Z^2 + ZY}{(ZY)^2}} = \sqrt{\frac{Z}{Y} + 1 + \frac{1}{Y^2} + \frac{1}{ZY}} \qquad \text{(H-158)}$$

i.e.,

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \sqrt{\frac{1}{Y^2} + \left(Z + \frac{1}{Z}\right)\frac{1}{Y} + 1} \qquad \text{(H-159)}$$

At this point Chisholm denoted *C* as

$$C = Z + \frac{1}{Z} \qquad \text{(H-160)}$$

Therefore, we have

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \sqrt{\frac{1}{Y^2} + C\frac{1}{Y} + 1} \qquad \text{(H-161)}$$

Now, from

$$Y = \frac{x}{1-x}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-138)}$$

and

$$X_{LM} = \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} \qquad \text{(4)}$$

we have

$$\frac{\dot{m}_{gs}}{\dot{m}_g} = \sqrt{1 + CX_{LM} + X_{LM}^2} \qquad \text{(H-162)}$$

or

$$\dot{m}_g = \frac{\dot{m}_{gs}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} \qquad \text{(H-163)}$$

where

$$\dot{m}_{gs} = EA_t C_d \sqrt{2\rho_g \Delta P_{tp}} \qquad \text{(H.144)}$$

as $(P_1 - P_2) = \Delta P_{tp}$ by definition. That is, $\dot{m}_{gs} = \dot{m}_{g\,Apparent}$. So finally we have the Chisholm correlation form

$$\dot{m}_g = \frac{\dot{m}_{g\,Apparent}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} \qquad \text{(G-11)}$$

It is common in actual use to include the gas expansion factor in the uncorrected or apparent gas mass flow calculation (even though Chisholm assumed incompressible flow).

It should be noted that the Chisholm model, like the Murdock model is based on the assumption of separated two-phase flow only. However again like Murdock, Chisholm used various data sets that more than likely had flow patterns that included in practice some liquid entrainment. In 1967 Chisholm had new data [7] as well as Murdock's data and the data sets of James [62], Bizon [64] and Thom [65]. All the data sets except some of Murdock's were for water/steam flows. All data except Murdock's spanned general two-phase flow conditions and many orifice plate geometries; of these, only Bizon and James' data sets included wet gas flow conditions as defined in this document, and these two data sets also included many data points far outside the wet gas limit. The experimental values of *C* for the wet gas relevant data sets where reported to be Bizon 2.34, James 2.41, and Murdock 2.66. (In general, for two-phase flow with higher Lockhart–Martinelli parameters the C values were much higher.) No independent check on these values is known to have been published.

In 1977, Chisholm published a research note [8] in which he derived an expression for his parameter *C* [see eq. (H-160)], which was solely a function of the gas-to-liquid density ratio. This function is now derived.

The homogeneous model gives eq. (H-8):

$$\dot{m}_{total} = \frac{\dot{m}_g}{x} = EA_t\varepsilon C_d\sqrt{2\rho_{homogeneous}\Delta P_{tp}} \quad \text{(H-8)}$$

If it is imagined that the total wet gas mass flowing flows as liquid, it can be said that:

$$\dot{m}_{total} = \dot{m}_g + \dot{m}_l = EA_tC_d\sqrt{2\rho_l\Delta P_{lo}} \quad \text{(H-164)}$$

where ΔP_{lo} is the differential pressure that would be measured under the assumption of all the wet gas mass flow being liquid. Note that the expansibility term, ε, has dropped out of eq. (H-164) as liquids are effectively incompressible. Also note that Chisholm assumed that the orifice meter's discharge coefficient would be the same for when the gas phase flowed alone and when the wet gas phase flowed as a homogeneous mixture: $C_{d_{homogeneous}} \cong C_d$. Therefore, by dividing eq. (H-8) by eq. (H-164) we get

$$\frac{EA_t\varepsilon C_d\sqrt{2\rho_{homogeneous}\Delta P_{tp}}}{EA_tC_d\sqrt{2\rho_l\Delta P_{lo}}} = 1 \quad \text{(H-165)}$$

and as Chisholm assumed the expansibility factor to be approximately unity for the homogeneous flow the following equation was derived:

$$\frac{\Delta P_{tp}}{\Delta P_{lo}} = \frac{\rho_l}{\rho_{homogeneous}} \quad \text{(H-166)}$$

Considering eq. (19)

$$\alpha_l = 1 - \alpha = \frac{A_l}{A} \quad \text{(19)}$$

and continuity eqs. (H-167) and (H-168)

$$\dot{m}_g = x\dot{m} = \rho_g A_g U_g \quad \text{(H-167)}$$

$$\dot{m}_l = (1-x)\dot{m} = \rho_l A_l U_l \quad \text{(H-168)}$$

then from eq. (19) and continuity eqs. (H-167) and (H-168)

$$\frac{1}{\alpha_l} = \frac{A}{A_l} = \frac{A_l + A_g}{A_l} = \frac{\frac{(1-x)\dot{m}}{\rho_l U_l} + \frac{x\dot{m}}{\rho_g U_g}}{\frac{(1-x)\dot{m}}{\rho_l U_l}} = \frac{\frac{U_g}{U_l}\frac{\rho_g}{\rho_l}(1-x) + x}{\frac{U_g}{U_l}\frac{\rho_g}{\rho_l}(1-x)} \quad \text{(H-169)}$$

or by rearranging [and using eq. (H-131)]:

$$\frac{1}{\alpha_l} = \frac{S_R\frac{(1-x)}{\rho_l} + \frac{x}{\rho_g}}{S_R\frac{(1-x)}{\rho_l}} \quad \text{(H-170)}$$

Now the gas volume fraction is defined as

$$\mathrm{GVF} = \frac{\dot{Q}_g}{\dot{Q}_l + \dot{Q}_g} \qquad (12)$$

which, with substitution of the continuity eqs. (H-167) and (H-168) in the form of eqs. (H-171) and (H-172),

$$\dot{Q}_g = \frac{x\dot{m}}{\rho_g} = A_g U_g \qquad \text{(H-171)}$$

$$\dot{Q}_l = \frac{(1-x)\dot{m}}{\rho_l} = A_l U_l \qquad \text{(H-172)}$$

can be written as

$$\mathrm{GVF} = \frac{\dfrac{x}{\rho_g}}{\dfrac{(1-x)}{\rho_l} + \dfrac{x}{\rho_g}} \qquad \text{(H-173)}$$

Chisholm now stated that as the square root of the differential pressure that would be produced if the liquid phase flowed alone (ΔP_l) would be

$$\sqrt{\Delta P_l} = \frac{(1-x)\dot{m}_{total}}{EA_t C_d \sqrt{2\rho_l}} \qquad \text{(H-174)}$$

then as from eq. (H-164)

$$\sqrt{\Delta P_{lo}} = \frac{\dot{m}_{total}}{EA_t C_d \sqrt{2\rho_l}} \qquad \text{(H-175)}$$

it can be said that

$$\frac{\Delta P_l}{\Delta P_{lo}} = (1-x)^2 \qquad \text{(H-176)}$$

Developing eq. (H-170) further gives

$$\frac{1}{\alpha_l} = \frac{\left(S_R \dfrac{(1-x)}{\rho_l} + \dfrac{x}{\rho_g}\right)\left(\dfrac{x}{\rho_g} + \dfrac{1-x}{\rho_l}\right)}{S_R \dfrac{(1-x)}{\rho_l}\left(\dfrac{x}{\rho_g} + \dfrac{1-x}{\rho_l}\right)} \qquad \text{(H-177)}$$

therefore

$$\frac{1}{\alpha_l} - 1 = \frac{1}{\alpha_l}(1-\alpha_l) = \frac{\left(\dfrac{x}{\rho_g} + \dfrac{1-x}{\rho_l}\right)\left(\dfrac{x}{\rho_g}\right)}{S_R \dfrac{(1-x)}{\rho_l}\left(\dfrac{x}{\rho_g} + \dfrac{1-x}{\rho_l}\right)} \qquad \text{(H-178)}$$

or

$$\frac{1}{\alpha_l}=\frac{\left(\frac{x}{\rho_g}+\frac{1-x}{\rho_l}\right)\left(\frac{x}{\rho_g}\right)}{S_R(1-\alpha_l)\frac{(1-x)}{\rho_l}\left(\frac{x}{\rho_g}+\frac{1-x}{\rho_l}\right)} \qquad \text{(H-179)}$$

Now Chisholm introduced a parameter called the Armand constant, C_A, which was a parameter named after a Russian two-phase flow research engineer in the late 1940s. The Armand constant is the ratio of void fraction to the gas volume flow ratio. That is

$$C_A=\frac{1-\alpha_l}{GVF}=\frac{\alpha_g}{GVF} \qquad \text{(H-180)}$$

So now substituting eq. (H-173) into eq. (H-180) we get

$$C_A=\frac{1-\alpha_l}{\left(\frac{x}{\rho_g}\right)}\left(\frac{x}{\rho_g}+\frac{1-x}{\rho_l}\right) \qquad \text{(H-181)}$$

and therefore eq. (H-179) becomes

$$\frac{1}{\alpha_l}\frac{\left(\frac{x}{\rho_g}+\frac{1-x}{\rho_l}\right)}{S_R\frac{(1-x)}{\rho_l}C_A} \qquad \text{(H-182)}$$

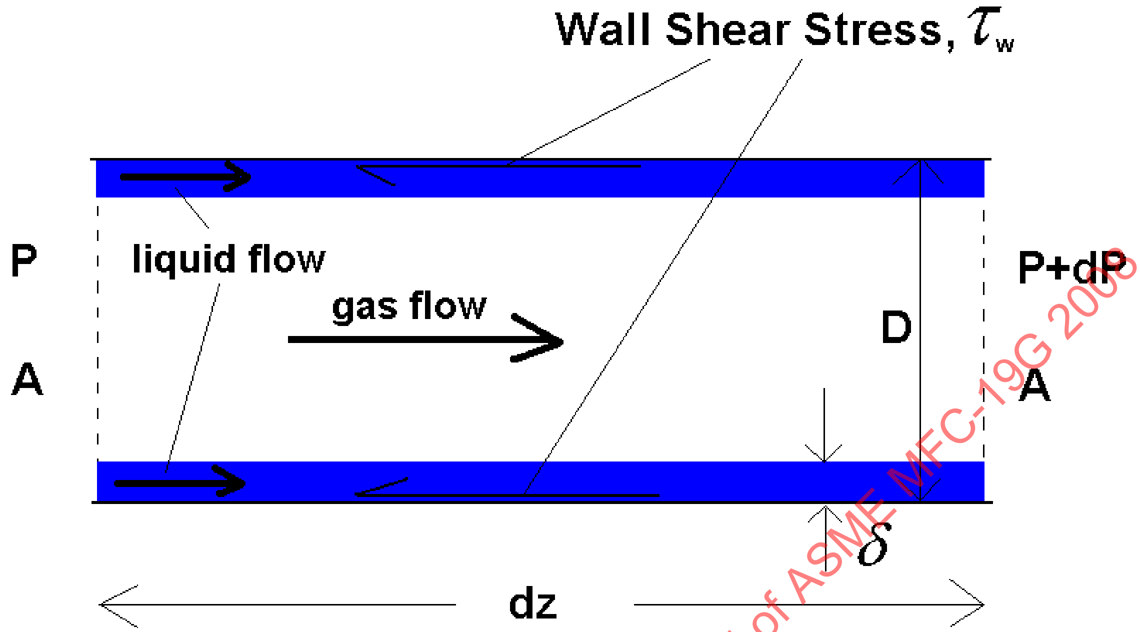
From analysis of the available two-phase flow data Chisholm stated that the Armand constant could be approximated to unity. This is indeed a reasonable approximation for wet gas flows. The reasons are described in Nonmandatory Appendix B. Chisholm then introduces "a well known approximate relationship for friction":

$$\frac{\Delta P_{tp}}{\Delta P_l}\approx\frac{1}{\alpha_l^2} \qquad \text{(H-183)}$$

It should be noted here that eq. (H-183) is well known among two-phase flow researchers for annular flow. That is, Chisholm is using an annular flow pattern equation. Note that Chisholm claimed the model to be for separated flow, but did not say it was for stratified flow as could be falsely presumed by readers of his paper that do not follow each step of the equation's development carefully. Equation (H-183) is derived[4] as follows:

[4] This derivation was supplied by Dr. Dempster of Strathclyde University, Glasgow, UK.

Fig. H-2 Schematic Diagram of a Horizontal Annular Flow Pattern



Force balance on whole pipe section for two-phase steady horizontal flow:

$$-dP_{tp}A - \tau_w(\pi D)dz = 0 \quad \text{(H-184)}$$

Force balance on whole pipe section if the liquid phase flowed alone through the pipe section:

$$-(dP)_l A - \tau_l(\pi D)dz = 0 \quad \text{(H-185)}$$

where

τ_l = shear stress on the control volume surface (i.e., the pipe wall) for if the liquid phase flowed alone

τ_w = shear stress on the control volume surface (i.e., the pipe wall) for the actual two-phase flow

Therefore, we know that $-(dP)_l A = \tau_l(\pi D)dz$. Dividing eq. (H-184) by eq. (H-185) gives

$$\frac{-(dP_{tp})A}{-(dP_l)A} = \frac{\tau_w(\pi D)dz}{\tau_l(\pi D)dz} \quad \text{(H-186)}$$

That is

$$\frac{\Delta P_{tp}}{\Delta P_l} = \frac{\tau_w}{\tau_l} \quad \text{(H-187)}$$

Now, by definition

$$\tau_w = f_{tp}\left(\frac{\rho_l U_l^2}{2}\right) \quad \text{(H-188)}$$

and

$$\tau_l = f_l\left(\frac{\rho_l U_{sl}^2}{2}\right) \quad \text{(H-189)}$$

where

f_l = friction factor if the liquid flowed alone in the pipe

f_{tp} = two-phase friction factor

U_l = actual average liquid velocity in the annular ring

U_{sl} = average liquid velocity if the liquid flowed alone in the pipe

Substituting eqs. (H-188) and (H-189) into eq. (H-187) gives

$$\frac{\Delta P_{tp}}{\Delta P_l} = \frac{\tau_w}{\tau_l} = \frac{f_{tp}\left(\dfrac{\rho_l U_l^2}{2}\right)}{f_l\left(\dfrac{\rho_l U_{sl}^2}{2}\right)} = \frac{f_{tp}}{f_l}\left(\frac{U_l^2}{U_{sl}^2}\right) \quad \text{(H-190)}$$

From continuity

$$A_l U_l = A U_{sl} \quad \text{(H-191)}$$

Therefore, substituting eq. (H-191) into eq. (H-190) gives:

$$\frac{\Delta P_{tp}}{\Delta P_l} = \frac{f_{tp}}{f_l}\left(\frac{A^2}{A_l^2}\right) = \frac{f_{tp}}{f_l}\frac{1}{\alpha_l^2} \quad \text{(H-192)}$$

For annular flow, often researchers assume that there is not a great difference in the friction factors of the actual two-phase flow and that which exists if the liquid flows alone. Hence, the ratio of friction factors in eq. (H-192) is often approximated to unity $f_{tp}/f_l \approx 1$. Hence, eq. (H-192) reduces to eq. (H-183) as required

$$\frac{\Delta P_{tp}}{\Delta P_l} \approx \frac{1}{\alpha_l^2} \quad \text{(H-183)}$$

Chisholm applied this approximation to the model. Now, letting C_A = 1, squaring eq. (H-182), and combining the result with eq. (H-183) gives

$$\frac{1}{\alpha_l^2} = \left[\frac{\left(\dfrac{x}{\rho_g} + \dfrac{1-x}{\rho_l}\right)}{S_R\dfrac{(1-x)}{\rho_l}}\right]^2 = \frac{\Delta P_{tp}}{\Delta P_l} \quad \text{(H-193)}$$

and as from eqs. (H-166) and (H-176) we have

$$\frac{\Delta P_{tp}}{\Delta P_l} = \frac{\Delta P_{tp}}{\Delta P_{lo}}\frac{\Delta P_{lo}}{\Delta P_l} = \frac{\rho_l}{\rho_{\text{homogeneous}}}\frac{1}{(1-x)^2} \quad \text{(H-194)}$$

We also have from eqs. (H-193) and (H-194):

$$\left[\frac{\left(\frac{x}{\rho_g}+\frac{1-x}{\rho_l}\right)}{S_R\frac{(1-x)}{\rho_l}}\right]^2=\frac{\rho_l}{\rho_{\text{homogeneous}}}\frac{1}{(1-x)^2} \qquad \text{(H-195)}$$

which rearranges to

$$S_R^2=\rho_l\rho_{\text{homogeneous}}\left(\frac{x}{\rho_g}+\frac{(1-x)}{\rho_l}\right)^2 \qquad \text{(H-196)}$$

and as by definition

$$\frac{1}{\rho_{\text{homogeneous}}}=\frac{x}{\rho_g}+\frac{1-x}{\rho_l} \qquad \text{(H-6)}$$

we have

$$S_R=\sqrt{\frac{\rho_l}{\rho_{\text{homogeneous}}}}=\sqrt{\rho_l}\sqrt{\frac{x\rho_l+(1-x)\rho_g}{\rho_g\rho_l}}=\sqrt{1+x\left\{\frac{\rho_l}{\rho_g}-1\right\}} \qquad \text{(H-197)}$$

Chisholm now stated that analysis of experimental results show that eq. (H-197) is valid at Lockhart–Martinelli parameters above unity (i.e., $X_{LM}>1$) and at Lockhart–Martinelli parameters of unity or less ($X_{LM}\leq 1$), the slip ratio, S_R, becomes independent of the quality/dryness fraction, x. Chisholm states that no discontinuity of slip ratio at the Lockhart–Martinelli parameter of unity should be assumed. Therefore, noting that, from eq. (4) we now have

$$X_{LM}=\frac{(1-x)}{x}\sqrt{\frac{\rho_g}{\rho_l}}=1 \qquad \text{(H-198)}$$

It is worth noting here that there is an assumption by some in industry that the Chisholm equation is formed purely from theoretical considerations and is therefore a theoretical equation and not a correlation. The use of eq. (H-198) is based solely on experimental results and therefore, although there is considerable theoretical reasoning in the development of the correction factor, the Chisholm wet gas orifice meter correction factor is not a theoretical model but a data fitted correlation.

Equation (H-198) can be rearranged to give eq. (H-199):

$$x=\frac{\sqrt{\rho_g}}{\sqrt{\rho_g}+\sqrt{\rho_l}} \qquad \text{(H-199)}$$

or

$$\frac{\rho_l}{\rho_g}=\left(\frac{1-x}{x}\right)^2 \qquad \text{(H-200)}$$

Substituting eq. (H-200) into eq. (H-197) gives

$$S_R = \sqrt{\frac{\rho_l}{\rho_{\text{hom}\,ogeneous}}} == \sqrt{1 + x\left\{\left(\frac{1-x}{x}\right)^2 - 1\right\}} = \sqrt{\frac{1}{x} - 1} \qquad \text{(H-201)}$$

Substituting eq. (H-199) into eq. (H-201) gives

$$S_R = \sqrt{\frac{1}{x} - 1} = \sqrt{\frac{1}{\frac{\sqrt{\rho_g}}{\sqrt{\rho_g} + \sqrt{\rho_l}}} - 1} = \sqrt{\frac{\sqrt{\rho_g} + \sqrt{\rho_l}}{\sqrt{\rho_g}} - \frac{\sqrt{\rho_g}}{\sqrt{\rho_g}}} = \sqrt{\frac{\sqrt{\rho_l}}{\sqrt{\rho_g}}} = \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} \qquad \text{(H-202)}$$

Now noting eq. (H-134) and eq. (H-160)

$$S_R = \frac{U_{g2}}{U_{l2}} = \frac{1}{Z}\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-134)}$$

$$C = Z + \frac{1}{Z} \qquad \text{(H-160)}$$

we can say that

$$C = \frac{1}{S_R}\sqrt{\frac{\rho_l}{\rho_g}} + S_R\sqrt{\frac{\rho_g}{\rho_l}} \qquad \text{(H-203)}$$

and substituting eq. (H-202) into eq. (H-203) gives

$$C = \left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}}\sqrt{\frac{\rho_l}{\rho_g}} + \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}}\sqrt{\frac{\rho_g}{\rho_l}} = \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} + \left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} \qquad \text{(G-12)}$$

Therefore, instead of an empirically found value for *C* in eq. (G-11) Chisholm had now updated the two-phase orifice plate meter correlation for use in high-quality/dryness fraction two-phase flows (i.e., $X_{LM} \leq 1$, which includes the ASME wet gas flow region) to an equation that had the parameter *C* calculated by a function of the gas-to-liquid density ratio. The final Chisholm wet gas flow orifice plate meter correlation is therefore

$$\dot{m}_g = \frac{EA_t\varepsilon C_d\sqrt{2\rho_g \Delta P_{tp}}}{\sqrt{1 + \left\{\left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} + \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}}\right\}X_{LM} + X_{LM}^2}} \qquad \text{(H-204)}$$

or

$$\dot{m}_g = \frac{\dot{m}_{g_{Apparent}}}{\sqrt{1+\left[\left\{\left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}}+\left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}}\right\}\left(\frac{\dot{m}_g}{\dot{m}_l+\dot{m}_g}\right)\right]+\left[\frac{\dot{m}_g}{\dot{m}_l+\dot{m}_g}\right]^2}} \qquad \text{(H-205)}$$

H-6 THE SMITH AND LEANG CORRELATION FOR ORIFICE PLATE METERS

The Smith and Leang equation [59] uses the concept of the blockage factor. The single-phase orifice plate flow equation for gas is

$$\dot{m}_g = EC_dY_gA_t\sqrt{2\rho_g\Delta P_g} \qquad \text{(G-1)}$$

This was modified by Smith and Leang to take account of the blockage by the liquid in a wet gas flow by introducing a blockage factor ξ:

$$\dot{m}_g = E\xi A_tY_{tp}C_d\sqrt{2\rho_g\Delta P_{tp}} \qquad \text{(G-14)}$$

The blockage factor effectively corrects the meters throat area to the effective throat area as seen by the gas phase due to the liquid's presence. Smith and Leang state that the assumed primary influences on the blockage factor are

(a) the volume of liquid in the gas flow and the associated phenomena of wakes behind drops (i.e., the correlation does take some account of entrainment) and in between waves on liquid films.

(b) additional blockage caused by liquid breakup for lower quality/higher liquid-to-gas ratio flows.

Smith and Leang postulate that the first influence between the blockage factor, ξ, and quality/dryness fraction, *x*, may be linear. They also acknowledge that the second influencing factor is significant for flows with changing flow patterns, and the flows quality or dryness fraction is a major influence on the flow pattern. For these reasons Smith and Leang postulated a relationship between the quality, *x*, and the blockage factor, ξ, for high quality flow that is close to linear, with the higher the quality the smaller the blockage effect. However, they postulated the situation would become more complex as the quality or dryness fraction reduces to below a critical value where a separate correlation may be required. Smith and Leang chose to form a correlation for qualities above 10% (i.e., $x = 0.1$), evidence that they did not consider the second mentioned influence of the blockage factor as significant for wet gas flow.

The correlation chosen by Smith and Leang was of the form:

$$\xi = C_1 + C_2x + \frac{C_2}{x^2} \qquad \text{(H-206)}$$

This form was chosen by consideration of the flow phenomena. The first two terms represent the expected linear relationship between the blockage factor and quality, *x* at higher quality flows. The third term represents the increasing influence of an additional blockage caused by a changing flow pattern at lower qualities. The third term's exponent was simply an estimate. Using saturated steam orifice plate meter data (at $x \geq 0.1$) from various sources, Smith and Leang fitted the following equation:

$$\xi = 0.637 + 0.4211x - \frac{0.00183}{x^2} \qquad \text{(G-15)}$$

The second term shows the linearity of the relationship between the blockage factor, ξ, and quality/dryness fraction, x, at high-quality flow conditions. It can be seen that for high-quality flows the third term has little influence and only becomes significant at lower qualities. A research discussion at the end of the Smith and Leang paper [59] comments that not all the orifice plate data is from standard ASME specified orifice plate meters. This then reduces the validity of universally applying eq. (G-15).

The Smith and Leang correlation is a general two-phase flow correlation formed from steam / water data only. Some of the experimental data used does overlap the wet gas region, but other data used does not. For this reason, the data fit [eq. (G.15)] at extremely high qualities gives blockage factors in excess of unity and at a quality of unity the correlation does not reduce to the gas equation (as required by theory) but indicates an over reading of 5.627%, which, of course, is not correct. The quality range of the assorted data sets is seen from a data set table given by Smith and Leang [59] to be $0 \le x \le 0.9672$. Therefore, the Smith and Leang Blockage correlation should (like all correlations) be used only within the limits stated for the correlation data.

The final equation formed by Smith and Leang for total mass flow of a two-phase flow is

$$\dot{m}_g = E\left(0.637 + 0.4211x - \frac{0.00183}{x^2}\right) A_t Y_{tp} C_d \sqrt{2\rho_g \Delta P_{tp}} \qquad \text{(H-207)}$$

i.e.,

$$\dot{m}_g = E\left(0.637 + 0.4211\left(\frac{\dot{m}_g}{\dot{m}_g + \dot{m}_l}\right) - 0.00183\left(\frac{\dot{m}_g + \dot{m}_l}{\dot{m}_g}\right)^2\right) A_t Y_{tp} C_d \sqrt{2\rho_g \Delta P_{tp}} \qquad \text{(H-208)}$$

H-7 THE LIN CORRELATION FOR ORIFICE PLATE METERS

Lin published a two-phase flow correlation for orifice plate meters in 1982 [9]. Like Murdock and Chisholm, Lin assumed a separated flow through the meter, no thermodynamic effects, incompressible flow and that the pressure drop across the meter is the same for both phases. The correlation development is similar to that of Murdock's. Lin derives the same eq. (H-31) as Murdock

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = \left(\frac{K_l}{(K_l)_{tp}}\right)\sqrt{\frac{\Delta P_l}{\Delta P_g}} + \left(\frac{K_g Y}{(K_g Y)_{tp}}\right) \qquad \text{(H-31)}$$

However, Lin then goes on to say that the meter error is dependent on the shear between the separated phases and that the shear is solely dependent on pressure. Although it is not stated by Lin as such, the Lin correlation effectively upgrades Murdock's correlation to a similar modeling detail to that of the Chisholm correlation. The derivation is now given.

From Murdock's eq. (H-31), Lin assumes that

$$K_l = \theta (K_l)_{tp} \qquad \text{(H-209)}$$

and

$$K_g Y_g \cong (K_g Y_g)_{tp} \qquad \text{(H-210)}$$

Although it is not stated as so by Lin at this stage, these assumptions expressed as eqs. (H-209) and (H-210) in fact match the findings of Murdock where

$$M = \frac{K_l}{K_{l_{tp}}} \equiv \theta \qquad \text{(H-211)}$$

Lin now expresses eq. (H-213) as eq. (H-212)

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = \theta\sqrt{\frac{\Delta P_l}{\Delta P_g}} + 1 \qquad \text{(H-212)}$$

From here Lin produces an expression for the parameter θ that is solely a function of the gas-to-liquid density ratio and hence Lin can be said to have upgraded the Murdock correlation to include the effect of pressure (similar to the Chisholm correlation). The expression is derived as follows:

If the gas phase in the two-phase flow is considered alone:

$$\dot{m}_g = EA_{t_g} Y_{g_{tp}} C_{d_{g,tp}} \sqrt{2\rho_g \Delta P_{tp}} \qquad \text{(H-213)}$$

or

$$\sqrt{\Delta P_{tp}} = \frac{\dot{m}_g}{EA_{t_g} Y_{g_{tp}} C_{d_{,g,tp}} \sqrt{2\rho_g}} \qquad \text{(H-214)}$$

If the gas phase is imagined to flow alone through the meter:

$$\dot{m}_g = EA_t Y_g C_{d_g} \sqrt{2\rho_g \Delta P_g}$$

or

$$\sqrt{\Delta P_g} = \frac{\dot{m}_g}{EA_t Y_g C_{d_g} \sqrt{2\rho_g}} \qquad \text{(H-215)}$$

If the liquid phase is imagined to flow alone through the meter:

$$\dot{m}_l = EA_t C_{d_l} \sqrt{2\rho_l \Delta P_l}$$

or

$$\sqrt{\Delta P_l} = \frac{\dot{m}_l}{EA_t C_{d_l} \sqrt{2\rho_l}} \qquad \text{(H-216)}$$

Then substituting eqs. (H-214), (H-215), and (H-216) into eq. (H-212) gives:

$$\left(\frac{\dfrac{\dot{m}_g}{EA_{t_g} Y_{g_{tp}} C_{d_g} \sqrt{2\rho_g}}}{\dfrac{\dot{m}_g}{EA_t Y_g C_{d\,g} \sqrt{2\rho_g}}} \right) = \theta \left(\frac{\dfrac{\dot{m}_l}{EA_t C_{d_l} \sqrt{2\rho_l}}}{\dfrac{\dot{m}_g}{EA_t Y_g C_{d_g} \sqrt{2\rho_g}}} \right) + 1 \qquad \text{(H-217)}$$

Lin makes the following unstated assumptions:

(a) The products of the discharge coefficient and expansibility of the separated gas phase in the two-phase flow ($Y_{g_{tp}} C_{d_{g_{tp}}}$) and of the gas flowing alone ($Y_g C_{d_g}$) are approximately the same.

(b) The product of the expansibility and discharge coefficient of the gas flow if it flowed alone ($C_{d_g} Y_g$) is approximately the same as the discharge coefficient of the liquid flow if it flowed alone (C_{d_l}).

(c) The throat void fraction ($\alpha_t = \frac{A_{t_g}}{A_t}$) is equal to the meter inlet void fraction ($\alpha = \frac{A_g}{A}$).

Therefore, eq. (H-217) can be expressed as follows [with use of eq. (16)]:

$$\left(\frac{\frac{1}{A_{t_g}}}{\frac{1}{A_t}}\right) = \frac{1}{\alpha} = \theta\left(\frac{\frac{\dot{m}_l}{\sqrt{\rho_l}}}{\frac{\dot{m}_g}{\sqrt{\rho_g}}}\right) + 1 = \theta\frac{\dot{m}_l}{\dot{m}_g}\sqrt{\frac{\rho_g}{\rho_l}} + 1 = \theta\frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} + 1 \qquad \text{(H-218)}$$

or alternatively,

$$\alpha = \frac{1}{1+\theta\sqrt{\frac{\rho_l}{\rho_g}\left\{\frac{\rho_g}{\rho_l}\left(\frac{1-x}{x}\right)\right\}}} \qquad \text{(H-219)}$$

Note that from the definition of the void fraction and mass continuity,

$$\alpha = \frac{A_g}{A} = \frac{A_g}{A_l + A_g} = \frac{\frac{x\dot{m}}{\rho_g \bar{U}_g}}{\frac{x\dot{m}}{\rho_g \bar{U}_g} + \frac{(1-x)\dot{m}}{\rho_l \bar{U}_l}} = \frac{1}{1+\left(\frac{1-x}{x}\right)\left(\frac{\bar{U}_g}{\bar{U}_l}\right)\left(\frac{\rho_g}{\rho_l}\right)} = \frac{1}{1+S_R\frac{\rho_g}{\rho_l}\frac{1-x}{x}} \qquad \text{(H-220)}$$

Therefore from eqs. (H-219) and (H-220):

$$\theta = S_R\sqrt{\frac{\rho_l}{\rho_g}} \qquad \text{(H-221)}$$

Lin now quotes Thom [65] and says the slip ratio, S_R, is solely a function of the line pressure. [This, it should be noted is a similar statement to with regard to two-phase slip as made later by Chisholm [8] as seen in eq. (H-202)]. Equation (H-221) can be written as

$$\theta = S_R\sqrt{\frac{\rho_l}{\rho_g}} = f\left(\frac{\rho_g}{\rho_l}\right) \qquad \text{(H-221)}$$

where *f* is a function found from experiment. Using various data sets of steam / water (from James [62], Murdock [5], Collins [66], Bizon [64], Ragolin [67]) as well as data from the Harbin Boiler Institute and Lin's own data set of R-113 mixtures, Lin produced the following equation:

$$\theta = 1.48625 - 9.26541\left(\frac{\rho_g}{\rho_l}\right) + 44.6954\left(\frac{\rho_g}{\rho_l}\right)^2 - 60.165\left(\frac{\rho_g}{\rho_l}\right)^3 - 5.12966\left(\frac{\rho_g}{\rho_l}\right)^4 - 26.5743\left(\frac{\rho_g}{\rho_l}\right)^5 \quad \text{(G-17)}$$

Some of the data used to create this correlation was obtained from nonstandard pressure tap positions. Due to this problem, Lin carried out some nonstandard pressure tapping tests and concluded that the method still gave good results. However, it should be noted that these tests were restricted to steam / water flows in small bore pipe at low pressures and with qualities low for wet gas metering considerations. (Lin considered general two-phase flow.)

The final expression Lin offered for calculating the mass flow for a known quality is derived as follows:

$$\sqrt{\frac{\Delta P_{tp}}{\Delta P_g}} = \theta\sqrt{\frac{\Delta P_l}{\Delta P_g}} + 1 \quad \text{(H-212)}$$

i.e.,

$$\sqrt{\Delta P_{tp}} = \theta\sqrt{\Delta P_l} + \sqrt{\Delta P_g} \quad \text{(H-222)}$$

Substituting in eqs. (H-215) and (H-216) and assuming the gas and liquid phases flowing alone to have the same discharge coefficient

$$\sqrt{\Delta P_{tp}} = \theta\frac{(1-x)\dot{m}_{total}}{EA_tC_d\sqrt{2\rho_l}} + \frac{x\dot{m}_{total}}{EA_tY_gC_d\sqrt{2\rho_g}} = \left[\theta\frac{(1-x)}{\sqrt{\rho_l}} + \frac{x}{Y_g\sqrt{\rho_g}}\right]\frac{\dot{m}_{total}}{EA_tC_d\sqrt{2}} \quad \text{(H-223)}$$

Lin assumed the gas expansibility term to be negligible ($Y_g \cong 1$) and then developed a final expression for total mass flow based on a liquid flow calculation. This is due to the fact that Lin was concerned with general two-phase flow. However, Lin's data has many points within the ASME definition of wet gas flow and hence an expression for total mass flow based on a gas flow calculation is equally valid. Such an expression is

$$\dot{m}_{total} = \frac{EA_tC_d\sqrt{2\rho_g\Delta P_{tp}}}{\left[\theta(1-x)\sqrt{\frac{\rho_g}{\rho_l}} + x\right]} \quad \text{(H-224)}$$

which with eq. (16) leads to

$$\dot{m}_g = x\dot{m}_{total} = \frac{EA_tC_d\sqrt{2\rho_g\Delta P_{tp}}}{\left[\theta\frac{(1-x)}{x}\sqrt{\frac{\rho_g}{\rho_l}} + 1\right]} \quad \text{(H-225)}$$

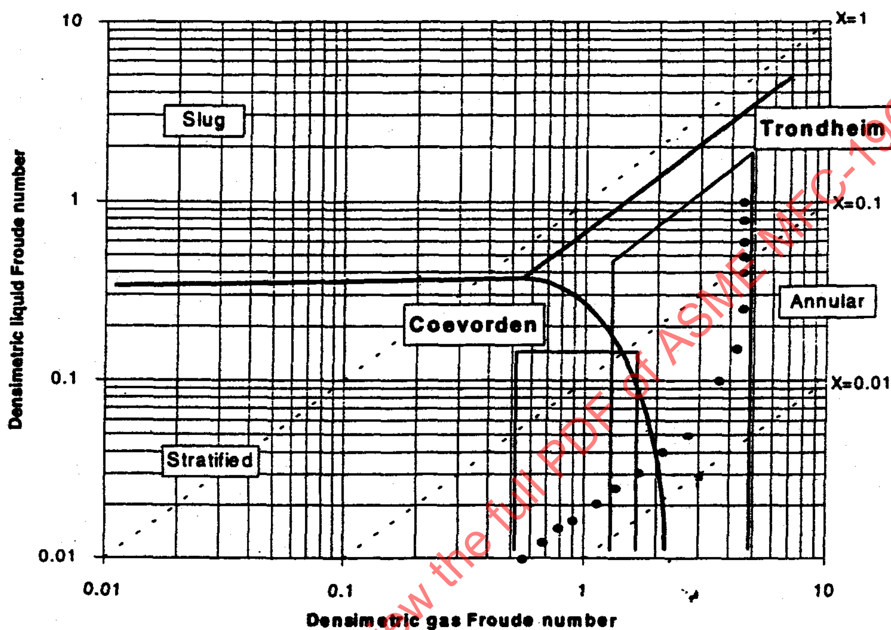
i.e.,

$$\boxed{\dot{m}_g = \frac{EA_tC_d\sqrt{2\rho_g\Delta P_{tp}}}{1+\theta X_{LM}} = \frac{\dot{m}_{g,Apparent}}{1+\theta X_{LM}}} \quad \text{(G-16)}$$

H-8 THE DE LEEUW CORRELATION FOR VENTURI METERS

In 1997 de Leeuw [10] presented a wet gas flow Venturi meter correlation. This is the first paper to specifically discuss Venturi meter performance with wet gas flows. De Leeuw had two data sets, one from a field location (Coevorden) and one from a test facility (SINTEF at Trondheim). The two test matrices are plotted on a Shell flow pattern map in Fig. H-3.

Fig. H-3 The Shell Flow Pattern Map With the Coevorden and Trondheim Test Matrices Superimposed

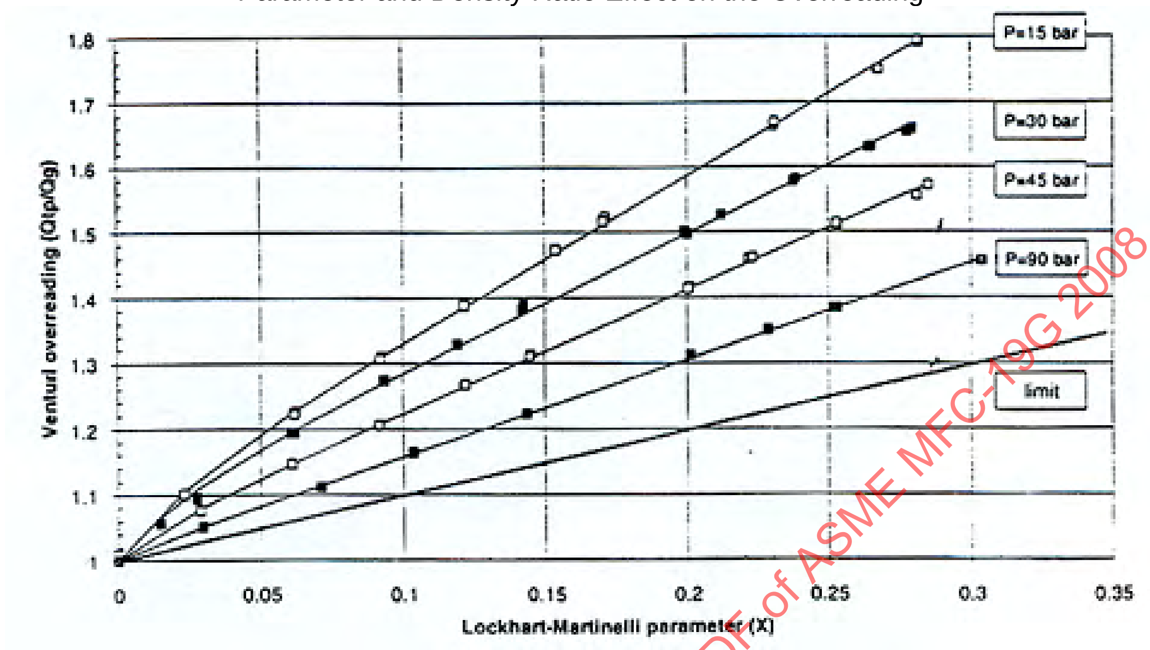


The de Leeuw research was groundbreaking. Along with confirming that the Lockhart–Martinelli parameter and gas-to-liquid density ratio were as important for a Venturi meter in determining the liquid induced gas overreading as with orifice plate meters (see Fig. H-4) important new discoveries were that the gas flow rate itself (or in dimensionless terms the gas densiometric Froude number) and the flow pattern were also important. De Leeuw stated that for a set Lockhart–Martinelli parameter and gas-to-liquid density ratio as the gas densiometric Froude number (i.e., the gas flow rate for a set pressure and liquid-to-gas mass flow rate ratio) increased, so did the liquid-induced error (see Fig. H-5).

In creating the wet gas Venturi meter correlation, de Leeuw stated that "… the experimental data for a single gas Froude number and line pressure can be described very well by the Chisholm equation by tuning the C factor." Chisholm had stated

$$C = \left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} \sqrt{\frac{\rho_l}{\rho_g}} + \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} \sqrt{\frac{\rho_g}{\rho_l}} = \left(\frac{\rho_l}{\rho_g}\right)^{\frac{1}{4}} + \left(\frac{\rho_g}{\rho_l}\right)^{\frac{1}{4}} \qquad \text{(G-12)}$$

Fig. H-4 The de Leeuw Wet Gas Venturi Meter Data Showing Lockhart–Martinelli Parameter and Density Ratio Effect on the Overreading



Here de Leeuw tuned the *C factor* by replacing Chisholm's constant $1/4$ with an experimentally found exponent he denoted as *n*. That is

$$C = \left(\frac{\rho_l}{\rho_g}\right)^n + \left(\frac{\rho_g}{\rho_l}\right)^n \quad \text{(G-20)}$$

That is, with each experimental combination of density ratio and gas densiometric Froude number a particular value for Chisholm's parameter *C* was found and then the corresponding value for the parameter *n* was found. Now, de Leeuw plotted constant gas densiometric Froude number lines on a parameter C vs. gas-to-liquid density graph (see Fig. H-6). It was stated by de Leeuw that constant gas densiometric Froude number lines plotted on Fig. H-6 could be represented with eq. (G-20) having a unique value of de Leeuw's parameter, *n*.

Fig. H-5 The de Leeuw Wet Gas Venturi Meter Correlation Showing the Effect of Gas Densiometric Froude Number and the Comparison of the de Leeuw Prediction to the Predictions of Murdock and Chisholm's Correlations

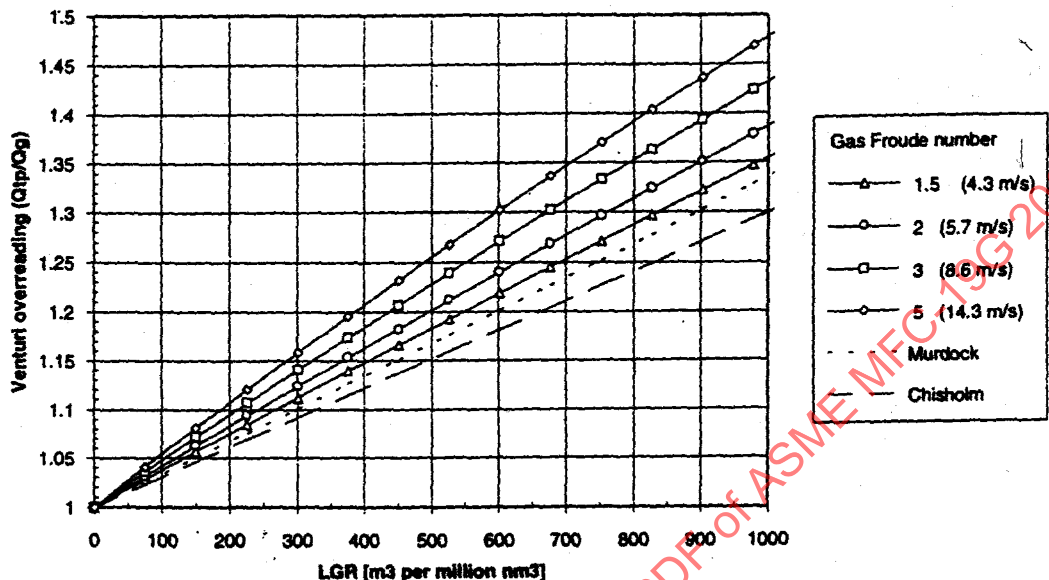
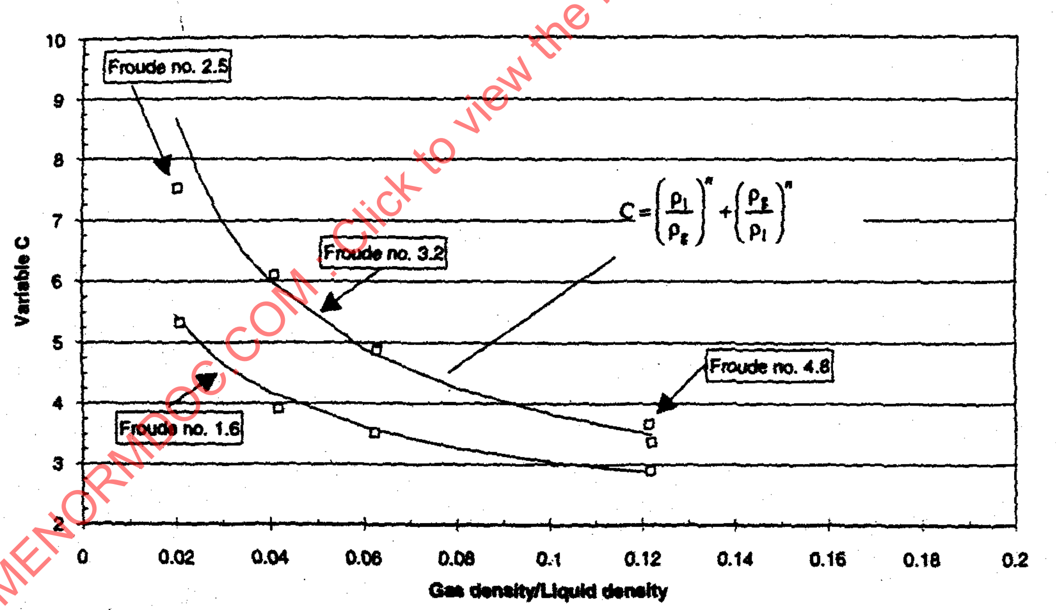


Fig. H-6 The de Leeuw C vs. Gas-to-Liquid Density Ratio Plot



NOTE: For any independent researcher planning to attempt to validate this, it is not immediately mathematically obvious how to separate *n* out of eq. (G-20). The following is an explanation:[5]

Let $$p = \frac{\rho_g}{\rho_l} \qquad \text{(H-226)}$$

[5] As supplied by Dr. van Maanen of Shell Global Solutions

and

$$q = p^n \qquad \text{(H-227)}$$

Therefore eq. (G-20) becomes:

$$C = \left(\frac{\rho_g}{\rho_l}\right)^n + \left(\frac{\rho_g}{\rho_l}\right)^{-n} = p^n + p^{-n} \qquad \text{(H-228)}$$

i.e.,

$$p^n - C + p^{-n} = 0 \qquad \text{(H-229)}$$

Therefore, multiplying by p^n gives

$$p^{2n} - Cp^n + 1 = 0 \qquad \text{(H-230)}$$

or

$$q^2 - Cq + 1 = 0 \qquad \text{(H-231)}$$

Therefore the positive root of q is

$$q = \frac{C \pm \sqrt{C^2 - 4}}{2} \qquad \text{(H-232)}$$

Hence

$$\left(\frac{\rho_g}{\rho_l}\right)^n = \frac{C \pm \sqrt{C^2 - 4}}{2} \qquad \text{(H-233)}$$

Therefore

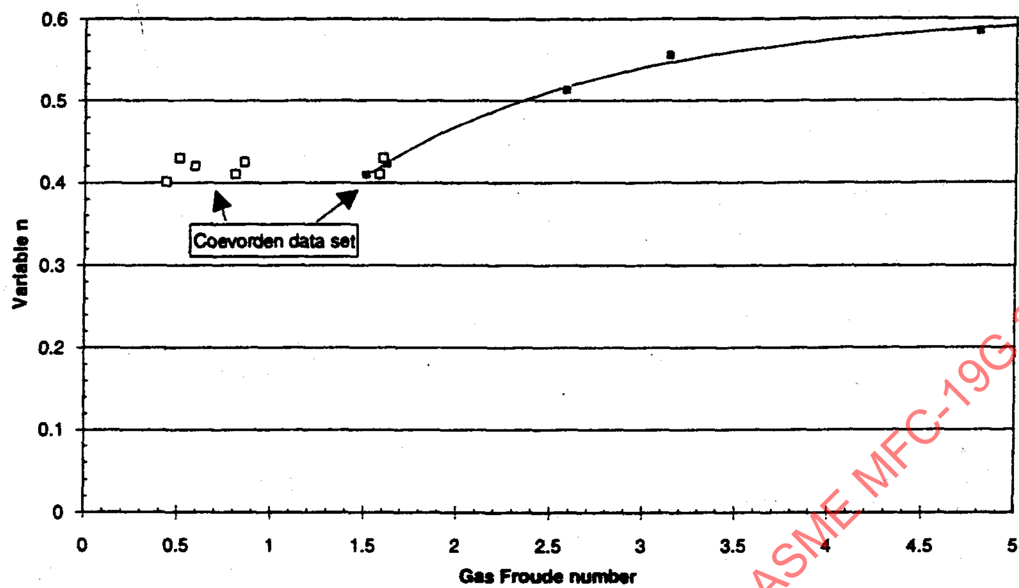
$$\ln\left\{\left(\frac{\rho_g}{\rho_l}\right)^n\right\} = \ln\left(\frac{C \pm \sqrt{C^2 - 4}}{2}\right) \qquad \text{(H-234)}$$

So finally

$$n = \frac{\ln\left(C \pm \sqrt{C^2 - 4}\right) - \ln 2}{\ln\left\{\left(\frac{\rho_g}{\rho_l}\right)\right\}} \qquad \text{(H-235)}$$

Having found by experiment that for the available data the parameter n was solely dependent on the gas densiometric Froude number de Leeuw then plotted n versus Fr_g. This plot is reproduced here as Fig. H-7.

Fig. H-7 The de Leeuw "n" vs. Gas Densiometric Froude Number Ratio Plot



Here relating Figs. H-3 and H-7 shows the first experimental confirmation of what Smith and Leang had speculated for DP meters. That is, the flow pattern can directly affect the liquid-induced overreading. This is indicated here by noting that the Coevorden data which has a gas densiometric Froude number range of 0.5 to 1.5 and is shown in Fig. H-3 to be almost entirely in the stratified flow pattern region appears to have a constant *n* value of approximately 0.41.

The Trondheim data that has a gas densiometric Froude number range of 1.5 to 4.8 and is shown in Fig. H-3 to be almost entirely in the annular mist flow pattern region has a varying *n* value which at the overlap of the Coevorden data matched the 0.41 value. It was pointed out by de Leeuw that the overlap gas densiometric Froude number of 1.5 is shown as being at the transition line between stratified and annular mist flow patterns (again see Figs. H-3 and H-7). Hence, it is postulated by de Leeuw that the change of flow pattern causes the change in the overreading's relationship with the gas densiometric Froude number and hence the value of the parameter *n*.

De Leeuw therefore stated a constant "n" value for stratified flow patterns within the experimental range (i.e., $0.5 < Fr_g < 1.5$) and an empirical fit to the data for the annular mist flow data (i.e., $Fr_g \geq 1.5$). The de Leeuw correlation is shown as eqs. (G-20), (G-21), and (G-22), which is used in conjunction with Chisholm's eq. (G-11).

$$\dot{m}_g = \frac{\dot{m}_{g_{Apparent}}}{\sqrt{1 + CX_{LM} + X_{LM}^2}} \qquad \text{(G-11)}$$

where

$$C = \left(\frac{\rho_g}{\rho_l}\right)^n + \left(\frac{\rho_l}{\rho_g}\right)^n \qquad \text{(G-20)}$$

and

$$n = 0.606 * \left(1 - e^{-0.746 Fr_g}\right) \quad \text{for } Fr_g \geq 1.5 \qquad \text{(G-21)}$$

$$n = 0.41 \quad \text{for } 0.5 \leq Fr_g \leq 1.5 \qquad \text{(G-22)}$$

This correlation is designed for an extrapolation of the data set in the direction of increasing gas-to-liquid density ratio. The correlation is not designed for extrapolation to lower gas-to-liquid density ratios than the data limit.

H-9 THE CONE-TYPE DP METER WET GAS FLOW CORRELATION

The Steven correlation was developed in 2000 at the time where industry was beginning to become more aware of the importance of the flow pattern on influencing a liquid-phase-induced DP meter error. With the physical modeling of changing flow patterns prohibitively complex and only possible with the inclusion of many assumptions, Steven followed de Leeuw's decision of fitting a "blind" data fitted equation to the data. (Note de Leeuw in fact took Chisholm's stratified flow orifice plate meter model equation and upgraded it by including in this a blind equation for the parameter *n*.) This task was relatively easy with researchers now having the curve fitting software tools available in the market that were unavailable to early researchers.

The cone-type DP meter 0.55 and 0.75 beta ratio wet gas flow correlations are both based on eq. (G-23) where A, B, and C are expressed as unique eqs. (G-24) to (G-26) for the 0.75 beta ratio and as equation sequence (G-27) to (G-29) for the 0.55 beta ratio cone-type DP meter.

$$\dot{m}_g = \frac{EA_t C_{d_g} Y_{g_{tp}} \sqrt{2\rho_g \Delta P_{tp}}}{\left(\frac{1 + AX + BFr_g}{1 + CX + BFr_g}\right)} \quad \text{(G-23)}$$

(a) Steven 0.75 Beta Ratio Cone Type DP Meter Correlation for

$\frac{\rho_g}{\rho_l} \geq 0.027$, then:

$$A = -0.0013 + \frac{0.3997}{\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(G-24)}$$

$$B = 0.0420 - \frac{0.0317}{\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(G-25)}$$

$$C = -0.7157 + \frac{0.2819}{\sqrt{\frac{\rho_g}{\rho_l}}} \quad \text{(G-26)}$$

For $\frac{\rho_g}{\rho_l} < 0.027$, then $A = 2.431$, $B = -0.151$, $C = 1$.

Note that at the gas-to-liquid density ratio value of 0.027 eqs. (6), (7), and (F-6) gives the constants $A = 2.431$, $B = -0.151$ and $C = 1$.

(b) Steven 0.55 Beta Ratio Cone-Type DP Meter Correlation

$$A = 1.224 + \frac{0.141}{\left(\frac{\rho_g}{\rho_l}\right)} \quad \text{(G-27)}$$

$$B = -0.0334 - \frac{0.00139}{\left(\frac{\rho_g}{\rho_l}\right)} \quad \text{(G-28)}$$

$$C = \sqrt{0.0805 + \left(\frac{0.0109}{\left(\frac{\rho_g}{\rho_l}\right)^2}\right)} \quad \text{(G-29)}$$

NONMANDATORY APPENDIX I[1]
THROTTLING CALORIMETER WORKED EXAMPLE

Let flow conditions be subscript 1 and chamber conditions be subscript 2.
Initial Wet Steam Conditions:
P_1 = 20 bara
$T_1 = T_{Saturated} = 212.4^0C$

Therefore, we can find from Steam Tables the enthalpy values associated with this saturated steam:

H_{v1} = 2797.22 kJ/kg
H_{l1} = 908.58 kJ/kg
$H_{lv1} = h_{v\,1} - h_{l\,1}$ = 1 890 kJ/kg

When a sample is isenthalpically expanded across throttling device such that it becomes superheated (as shown in Fig. 6.1.4.2):

P_2 = 5 bar
T_2 = 160^0C (with $T_{Saturated}$ = 151.8 ^{0}C)

Note $T_2 > T_{Saturated}$ steam is therefore superheated (as required).

Steam tables give the enthalpy value at the chamber conditions:
$h_{v\,2}$ = 2767.27 kJ/kg.

Therefore, as the expansion is isenthalpic we have $h_1 = h_{v\,2}$ = 2767.27 kJ/kg

Applying eq. (31):

$$x = \frac{h - h_l}{h_v - h_l} = \frac{2767.27 - 908.58}{2797.22 - 908.58} = 0.984$$

The wet steam flow has a quality of 0.984. That is:

$$x = \frac{\dot{m}_g}{\dot{m}_g + \dot{m}_l} = 0.983$$

The steam tables also state at the flow conditions the steam vapor and subcooled water densities. They are

$$\rho_g = 10 \ \ kg/m^3 \quad \text{and} \quad \rho_l = 846.8 \ \ kg/m^3$$

Therefore, from eq. (22) we can calculate the Lockhart–Martinelli parameter:

$$X_{LM} \frac{1-x}{x}\sqrt{\frac{\rho_g}{\rho_l}} = \frac{1-0.983}{0.983}\sqrt{\frac{10}{846.8}} \approx 0.002 \qquad (22)$$

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

It is noticeable that the Lockhart–Martinelli parameter of this example is extremely low. The main problem with throttling calorimeters is their limitation on the minimum quality (and therefore maximum Lockhart–Martinelli parameter) that they can read. This limitation is directly due to the absolute requirement to expand the saturated steam sample to a superheated sample in the chamber. This process requires a drop in pressure (see Fig. 6.1.4.2). As the chamber has to have a finite pressure the lower the flow lines pressure the smaller the available difference in the flow and chamber pressures and the smaller the range of qualities that can be measured.

Table I-1 shows typical minimum qualities for given saturated steam pressures for which throttling calorimeters can generally throttle the saturated steam to superheated steam. This is a guide only as posted on the internet by a throttling calorimeter supplier, and ASME does not take responsibility for its accuracy.

Table I-1 Guideline for Minimum Steam Qualities for Throttling Calorimeter Devices

Steam Pressure, psia	Minimum Steam Quality, *x*, Measurable	Steam Pressure, psia	Minimum Steam Quality, *x*, Measurable
20	99.385	120	95.45
25	98.93	140	94.09
30	98.55	160	94.8
35	98.23	180	94.65
40	97.93	200	94.31
45	97.67	300	93.53
50	97.43	400	93.18
60	97.03	500	92.85
70	96.68	600	92.78
80	96.37	700	92.84
90	96.1	800	93.01
100	95.86	900	93.27

NONMANDATORY APPENDIX J[1]
DETAILS OF GENERIC WET GAS FLOW METERING CONCEPTS

J-1 SINGLE-PHASE METERS IN SERIES

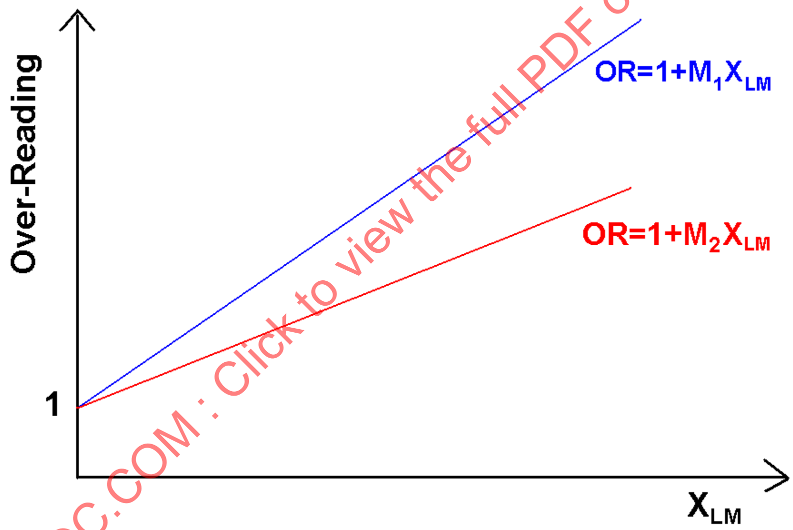
J-1.1 Differential Pressure Meters in Series

Paragraph 6-2.1 discusses the wet gas flow design of fitting multiple meters in series. The following is a simple theoretical example for the case of two different geometry DP meters with different wet gas meter responses in series. That is, each DP meter's wet gas data set has a unique gradient "M" on a Murdock plot (see Fig. 10). Fitting a linear line fit to two such Murdock plots gives two different equations in the Murdock correlation form as shown in eq. (J-1). Note that this equation is the same form as the Murdock correlation [eq. (G-6) — where Murdock stated for orifice meters that M = 1.26].

$$\dot{m}_g = \frac{\dot{m}_{g\,Apparent}}{1 + MX_{LM}} \qquad \text{(J-1)}$$

The Murdock graph with the two different DP meter data sets fitted would look like Fig. J-1.

Fig. J-1 Two DP Meters With Different Murdock Plots



Therefore, for each meter the actual gas mass flow rate can be found if the liquid flow rate is known by applying eqs. (J-2) and (J-3), respectively.

DP Meter 1:
$$\dot{m}_g = \frac{\dot{m}_{g,Apparent1}}{1 + M_1 X_{LM}} \qquad \text{(J-2)}$$

DP Meter 2:
$$\dot{m}_g = \frac{\dot{m}_{g,Apparent2}}{1 + M_2 X_{LM}} \qquad \text{(J-3)}$$

[1] In this Appendix, equations from other parts of the book are sometimes repeated for reference. These equations retain their original numbering when cited in this Appendix.

However, if no significant phase change is assumed, as the meters are in series the actual gas mass flow rate is the same for both meters. Therefore, the equations can be combined to give eq. (J-4):

$$\dot{m}_g = \frac{\dot{m}_{g,Apparent1}}{1 + M_1 X_{LM}} = \frac{\dot{m}_{g,Apparent2}}{1 + M_2 X_{LM}} \quad \text{(J-4)}$$

where the subscripts indicate unique values for each meter. Note that the numerator on each of these equations is the uncorrected erroneous gas mass flow rate prediction of that individual meter in question. Rearranging eq. (J-4) to give an expression for the Lockhart–Martinelli parameter, X_{LM}, gives eq. (J-5).

$$X_{LM} = \frac{m_{g,Apparent1} - m_{g,Apparent2}}{m_{g,Apparent2} M_1 - m_{g,Apparent1} M_2} \quad \text{(J-5)}$$

Notice that this expression for the Lockhart–Martinelli parameter contains no unknowns on the right hand side of the equation (assuming M_1 and M_2 are known from experiment) and hence the value of the Lockhart–Martinelli parameter is estimated. Therefore, eq. (J-5) can be substituted into either eq. (J-2) or eq. (J-3) and the gas mass flow rate found.

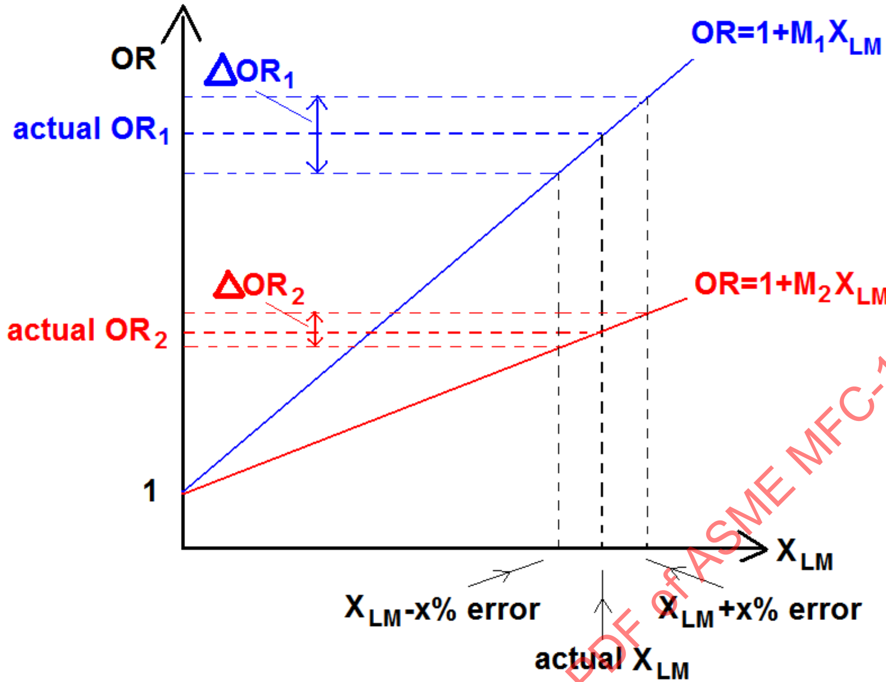
Aside

In practice it is best to substitute eq. (J-5) into the equation with the smallest Murdock gradient as this reduces the knock on effect of the uncertainty associated with the Lockhart–Martinelli parameter on the gas mass flow rate prediction. That is, the lower Murdock gradient equation has a lower sensitivity to errors in the Lockhart–Martinelli parameter. This is shown graphically in Fig. J-2.

In Fig. J-2 two Murdock-type correction factors are represented for meters with different Murdock gradients. If we assume that both corrections for the two different meters give the same uncertainties regarding the correct answer if the correct Lockhart–Martinelli parameter value is used, then, we see that when the Lockhart–Martinelli parameter has an error (marked in the Figure as ±*x*%), as in reality it will have due to the correlation uncertainties, the lower Murdock gradient correlation is less sensitive to this error than the higher Murdock gradient correlation and the actual overreading to be corrected (and hence the gas mass flow rate prediction itself) lies in a smaller range $OR \pm \Delta OR$. (Steven [13] gives a similar discussion on the relative sensitivities of DP meter correlations to uncertainties in liquid mass flow rates for the case where a stand alone DP meter is being used with a liquid mass flow rate found from sources external to the DP meter system.)

End of Aside

Fig. J-2 Relative Sensitivities of Murdock Fit Equations to Lockhart–Martinelli Parameter Errors



Here then we have:

$$\dot{m}_g = \frac{\dot{m}_{g,Apparent2}}{1+M_2\left(\dfrac{\dot{m}_{g,Apparent1}-\dot{m}_{g,Apparent2}}{\dot{m}_{g,Apparent2}M_1-\dot{m}_{g,Apparent1}M_2}\right)} \quad \text{(J-6)}$$

The Lockhart–Martinelli parameter and gas mass flow rate predictions will give a liquid flow rate prediction from eq. (4) rearranged:

$$\dot{m}_l = X_{LM}\,\dot{m}_g\sqrt{\frac{\rho_l}{\rho_g}} \quad \text{(4)}$$

This method of wet gas metering is a relatively simple concept. However, most meters do not have their wet gas flow response characterized well by a simple linear fit such as the Murdock fit.

The complexity of this method increases with more complex correlations for each of the meters, especially if it is not algebraically possible to separate the parameters required in the correlation mathematical forms. It should be noted here that there are different mathematical methods for achieving the measurement of the gas and liquid flow rates with DP meters in series. The method described above is a simple procedure. Most complex DP meter correlation pairs require iteration techniques.

The method obviously does not work well at low Lockhart–Martinelli parameters. As the method is a measurement by difference technique and at dry gas there is no difference between the meter readings (except that due to the relatively small dry gas metering uncertainties) the problem is at low liquid loadings there is little difference between the meter readings, and in such cases, measurement by difference techniques can have very significant uncertainties. It should also be noted that for the case of dry gas flow, if the individual meters have dry gas readings that are slightly different due to the existence of their respective meter uncertainties the system could see the difference in reading between the two meters and falsely claim a small quantity of liquid was present. For these reasons it is typical for these measurement by difference techniques to be limited to a minimum liquid loading. Nevertheless, this relatively simple wet gas metering method can give good results in some wet gas flow metering applications. This method generally has lower uncertainty gas flow rate measurement than the liquid flow rate measurement.

J-1.2 A Positive Displacement Meter and Differential Pressure Meter in Series

The earliest two-phase metering research works regarding a positive displacement (PD) meter and a secondary single-phase gas meter in two-phase flow were the works of Medvejev et al. [46] in 1972 and Chen et al. [47] in 1982. These papers are unavailable at the time of writing, but both works are summarized by Lin [40]. Lin's summary of Medvejev's research is now reproduced with comments:

Medvejev et al. used a volumetric (i.e., an oval gear PD) flow meter and an orifice plate (i.e., DP) meter in series to meter the gas and liquid flow rates of air/water mixtures. From experimental data, they found the volumetric flow rate of the mixture (denoted Q_{TP}) could be determined from the orifice plate equation as follows:

$$\dot{Q}_{tt} = C_{d_{tp}} A_t \sqrt{\frac{2\Delta P_{tp}}{\rho_{\mathrm{hom}\,ogenous}}} \qquad \text{(J-7)}$$

where

$\dot{Q}_t$ = the total volume flow of the two-phase flow

$C_{d_{tp}}$ = an orifice plate discharge coefficient for the two-phase mixture

$\rho_{\mathrm{hom}\,ogenous}$ = the homogeneous two-phase density

Medvejev expresses the orifice plate discharge coefficient for two-phase mixture $C_{d_{tp}}$ as

$$C_{d_{tp}} = kC_d \qquad \text{(J-8)}$$

where

k = a "corrective coefficient and is a function of quality (x) and the orifice plate meters beta ratio" found by experiment

C_d = the single phase discharge coefficient

[Notice that eq. (J-8) substituted into eq. (J-7) gives an expression for volumetric flow that is in effect the same method as expressed by James for orifice plate meters. This is, because as the corrective coefficient k is effectively a function of quality for a set beta ratio that corrects the error inherent in assuming perfectly homogeneous flow in a nonperfect homogenized flow as required by James [62].] Medvejev describes the two-phase flow homogeneous density by the following expression:

$$\rho_{\mathrm{hom}\,ogenous} = (GVF)\rho_g + (1 - GVF)\rho_l \qquad \text{(J-9)}$$

This expression it should be noted is the algebraic result of substituting eq. (24) (with quality x in terms of the GVF), i.e.:

$$x = \frac{1}{1 + \left(\left(\frac{1-(GVF)}{(GVF)}\right) * \frac{\rho_l}{\rho_g}\right)} \qquad \text{(24)}$$

into eq. (H-7), i.e.:

$$\rho_{homogenous} = \frac{\rho_l \rho_g}{\rho_l x + \rho_g (1 - x)} \tag{H-7}$$

Medvejev claimed that experiments indicated that for orifice plate meter beta ratios greater than 0.5 and GVFs less than 50%, the corrective coefficient k is approximately unity.

The total volume flow rate is obtained from the PD meter. Due to PD meter design, the total volume of the two-phase flow is simply the volume that has passed through the meter cavities. Therefore, Medvejev claims the total volume flow $\dot{Q}_t$ was known to 3% uncertainty. As we know that for no significant pressure drop or phase change:

$$\dot{Q}_t = \dot{Q}_g + \dot{Q}_l \tag{J-10}$$

and that eq. (J-7) can be rearranged to be expressed as

$$\rho_{homogenous} = \frac{2\left(C_{d_{tp}} A_t\right)^2 \Delta P_{tp}}{Q_t^2} \tag{J-11}$$

Finally, Medvejev states that as eq. (J-9) can be arranged to express the GVF:

$$GVF = \frac{\rho_{homogenous} - \rho_l}{\rho_g - \rho_l} \tag{J-12}$$

then the gas and liquid volume flows can be found by eqs. (J-13) and (J-14) where the total volume flow is found by the PD meter and the GVF is obtained from eq. (J-12), which in turn has the homogenized density found from eq. (J-11) where it was assumed from experimental data that the two-phase and single-phase discharge coefficients are the same:

$$\dot{Q}_g = (GVF)\dot{Q}_t \tag{J-13}$$

$$\dot{Q}_l = (1 - GVF)\dot{Q}_t \tag{J-14}$$

The mass flow rates are predicted by equations:

$$\dot{m}_g = \rho_g \dot{Q}_g = \rho_g (GVF)\dot{Q}_t \tag{J-15}$$

$$\dot{m}_l = \rho_l \dot{Q}_l = \rho_l (1 - GVF)\dot{Q}_t \tag{J-16}$$

In Medvejev's experiments, the experimental loop diameter was 50 mm. The orifice plate meter beta ratio range was 0.1 to 0.5. The pressure range and the gas flow rate range was not listed by Lin [40]. The GVF ranged from 0.2 to 0.98. So Medvejev tested across a wide range of GVFs and some data points were certainly wet gas flow points. However, Lin does not mention what process Medvejev used for cases of GVFs greater than 50% when the corrective coefficient k was no longer approximated to unity.

It is, however, intuitive that if the corrective coefficient k was found by experiment to be a particular function (say " f ") of quality, *x*, at higher GVF's [say $k = f(x)$], then there is still enough information to predict the flow rates. That is, we know that the two-phase discharge coefficient $C_{d_{tp}}$ was a function of quality. That is,

$$C_{d_{tp}} = kC_d = f(x)C_d \tag{J-17}$$