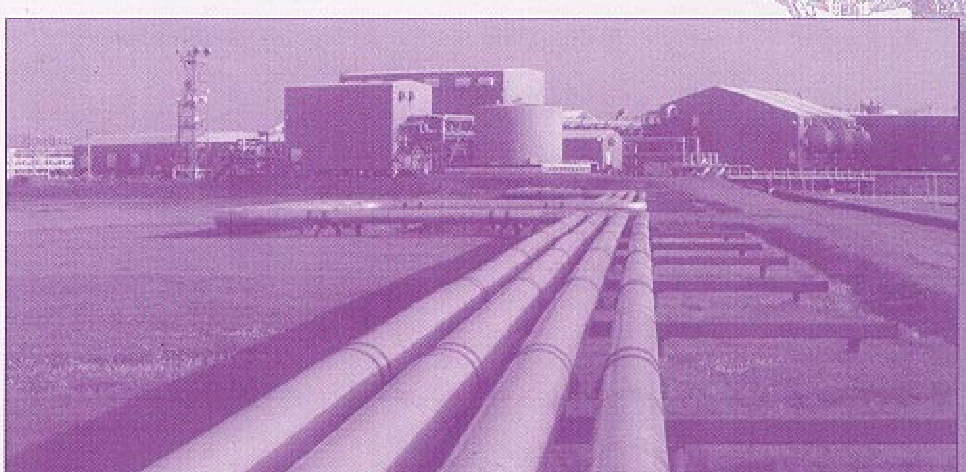
BP EXPLORATION

Multiphase Design Manual

A Reference Manual of Design Methods for Multiphase Oil and Gas Production Systems



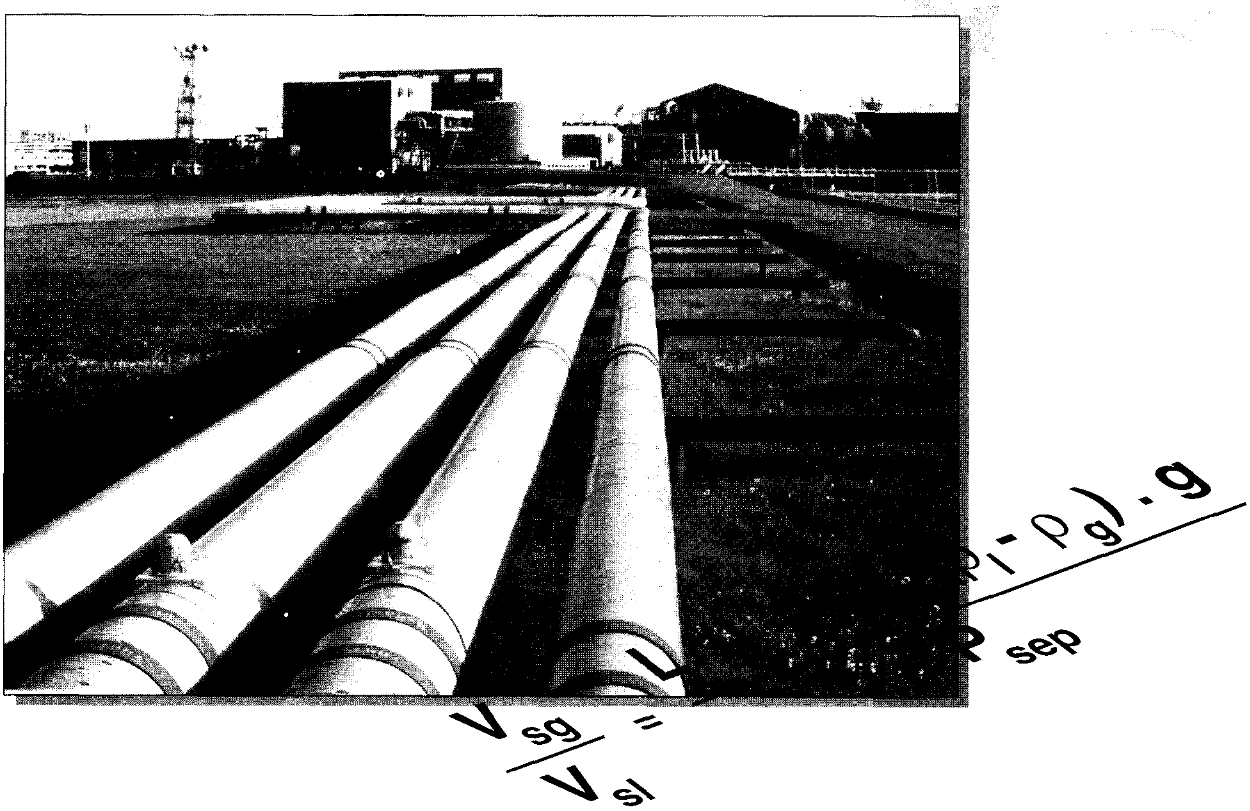
$$V_{sg} = \frac{L.(1-H_l).(\rho_l-\rho_g).}{\rho_{sep}}$$

Section 1. Multiphase Fundamentals

Introduction to Multiphase Flow

1.1 Multiphase Flow Nomenclature and Definitions

1.2 Flow Regimes



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1.1 Multiphase Flow Nomenchture and Definitions

Multiphase flow employs a wide range of concepts and nomenclature that require explanation for the engineer new to the area.

1.1.1 Volume formation factor and oil specific gravity

In a multiphase line gas dissolves in the oil according to the local pressure and temperature. The solubility of gas in the oil increases with increasing pressure and decreasing temperature. Hence the density of the oil, and the volume it occupies, is dependent on the local pressure and temperature. The reference condition at which the oil density is specified is:

Pressure = 14.7 psia
Temperature = 60°F

This is known as stock tank conditions (STC). The oil specific gravity is density of oil at stock tank conditions divided by the density of pure water at stock tank conditions.

$$\text{Oil SG} = \frac{\text{Density oil at STC}}{\text{Density water at STC}}$$

An oil production rate of 1 stock tank barrel per day (1STBD) is that flowrate which when produced to a stock tank at 14.7 psia and 60°F yields 1 barrel (1 bbl i.e. 5.614 ft^3) of oil. Hence, the flowrate in STBD is a measure of the stabilised (degassed oil) that would be yielded if the pipeline flow discharged to a stock tank.

The ratio of volume occupied by liquid at pipeline conditions to that occupied at stock tank conditions is the volume formation factor, B,:

$$B_o = \frac{\text{Pipeline barrels (pbbl)}}{\text{Stock tank barrels (STB)}}$$

Hence if the oil flowrate is Q_{os} (STBD) and the volume formation factor at some pipeline conditions is B_o, then the actual in-situ volume flowrate of oil Q_{oA} in pbbl/day is given by:

$$Q_{oA} = Q_{os} \times B_o$$

The value of B_o usually lies in the range 1 .O to 2.0. A more detailed discussion of the volume formation factor is given in the Physical Properties section of this manual, Section 2.3.

1.1.2 Water cut and water volume formation factor

The water cut is normally defined as the water fraction of the total liquid flow expressed as a percentage:

$$\textbf{Water cut} = \frac{Q_{ws}}{(Q_{os} + Q_{ws})} \times 100$$

where:

Q_{ws} = water production rate in STBD
Q_{os} = oil production rate in STBD

The water volume formation factor, B_w, is the ratio of the volume occupied by water at any pressure and temperature to that occupied at stock tank conditions. Hence the in-situ volume flowrate of water Q_{wA} in pbbl/day is given by:

$$Q_{WA} = B_w \times Q_{ws}$$

The value of B_w is normally close to 1 .O.

The total actual liquid flowrate in a pipeline, Q,, is simply given by:

$$QLA = Q_{oA} + Q_{wA}$$

1.1.3 Producing gas oil ratio and gas specific gravity

The producing gas oil ratio (GOR) is the total quantity of gas produced when reservoir fluids are flashed to stock tank conditions. The units of GOR are standard cubic feet per stock tank barrel (SCF/STB); that is, it is the volume of gas produced at 14.7 psia and 60°F associated with the production of 1 stock tank barrel of oil.

The specific gravity of the gas is the density of the gas produced at stock tank conditions divided by the density of air measured at 14.7 psia and 60°F.

$$\textbf{Gas SG} = \frac{\textbf{Density of stock tank gas}}{\textbf{Density of air at 14.7 psia, 60°F}}$$

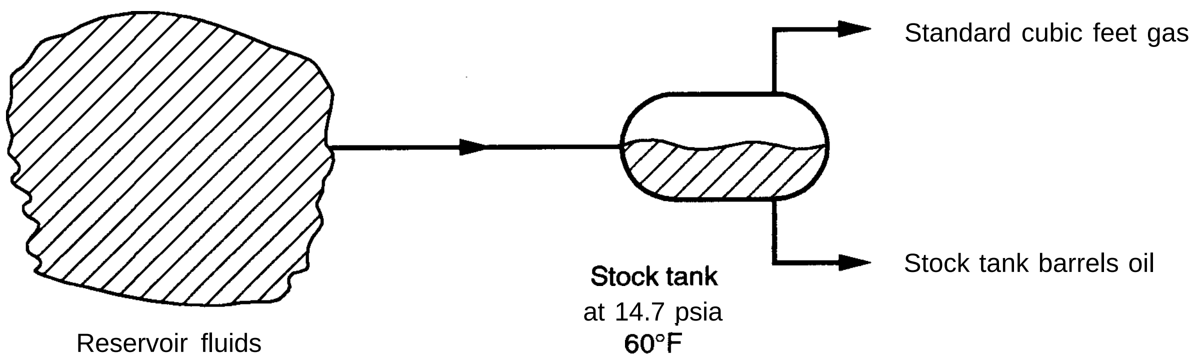


Figure 1.

In practice the GOR of reservoir fluids is dependent on the number of flash stages taken to get down to stock tank conditions. This is discussed in more detail in Section 2.3.

The solution gas oil ratio of oil (R_s) is the amount of gas dissolved in the oil at any pressure and temperature. Its units are SCF/STB. It represents the amount of gas produced when the pressure and temperature of the oil are altered to stock tank conditions. At any temperature and pressure the amount of free gas present in the pipeline is the difference between the values of GOR and R_s. Hence, if the oil flowrate is Q_{os} (STBD), the flowrate of gas at any temperature and pressure (Q_{GS}) is given by:

$$Q_{GS} = (GOR - R_s)$$

The units of gas flowrate are standard cubic feet per day, SCFD. To convert this gas flow into actual cubic feet per day (Q_{GA}) at pipeline conditions, an equation of state (EOS) is used. The simplest EOS is the engineering EOS:

$$PV = n\,z\,RT$$

or

$$\frac{P_1 V_1}{4 \quad T_1} = \frac{P_2 V_2}{z_2 T_2}$$

where:

P = pressure
V = volume
z = compressibility
T = temperature
n = number of moles

State 1 = stock tank conditions, 14.7 psia 60°F
State 2 = pipeline conditions

More information is given in Section 2.3 on GOR and R_s. Equations of state are discussed in more detail in Section 2.2.

1.1.4 In-situ gas and liquid density

The density of oil in a pipeline is less than that of stock tank oil due to the presence of dissolved gas and temperature expansion effects. These factors are taken into account by use of the volume formation factor and solution gas oil ratio. The density of oil at a condition 2 is then given by:

$$\rho_{o2} = \frac{1}{B_{o2}} \left[\rho_{os} + \frac{R_{s2}\,\rho_{gs}}{5.614} \right]$$

where:

ρ_{o2} = oil density at pipeline conditions (lb/ft^3)
ρ_{os} = oil density at stock tank conditions (lb/ft^3)
B_{o2} = oil volume formation factor at pipeline conditions
ρ_{gs} = gas density at stock tank conditions (lb/ft^3)
R_{s2} = solution gas oil ratio at pipeline conditions (SCF/STB)

If water is present the in-situ liquid density, ρ_{L2}, is simply a volume weighted average of the oil and water densities.

$$\rho_{L2} = \frac{\rho_{o2} \times Q_{oA} + \rho_{w2} \times Q_{wA}}{Q_{oA} + Q_{wA}}$$

where:

ρ_{w2} = water density at pipeline conditions
Q_{oA} and Q_{wA} are as defined above.

If the simple engineering EOS is used to correct gas density to pipeline conditions, then in-situ gas density is calculated from:

$$\rho_{g2} = \rho_{go} \times \frac{P_2}{P_o} \times \frac{T_o}{T_2} \times \frac{z_o}{z_2}$$

where:

State 0 refers to stock tank conditions
State 2 refers to pipeline conditions

ρ_g = gas density
P = pressure
T = absolute temperature
z = compressibility

1.1.5 Superficial liquid and gas velocity

The superficial liquid velocity, V_{SL}, is the velocity the liquid would have if it occupied the entire cross sectional area of the pipe. It is thus defined as:

$$V_{SL} = \frac{Q_{LA}}{A_p}$$

where:

Q_{LA} = actual liquid flowrate at pipeline conditions (ft^3/sec)
A_p = cross sectional area of pipe (ft^2)

Similarly the superficial gas velocity, V_{SG}, is the velocity the gas would have if it occupied the entire cross section of the pipe:

$$V_{SG} = \frac{Q_{GA}}{A_p}$$

where:

Q_{GA} = actual gas flowrate at pipeline conditions (ft^3/sec)

1.1.6 Mixture velocity and no slip hold-up

The mixture velocity in 2-phase flow is simply defined by:

$$V_m = V_{SL} + V_{SG}$$

where:

V_m = mixture velocity (ft/sec)

Now as V_{SL} and V_{SG} are given by:

$$V_{SL} = \frac{Q_{LA}}{A_p} \quad \text{and} \quad V_{SG} = \frac{Q_{GA}}{A_p}$$

then:

$$V_m = V_{SL} + V_{SG} = \frac{Q_{LA}}{A_p} + \frac{Q_{GA}}{A_p}$$

So if we define the total actual fluid flowrate Q_T as:

$$Q_T = Q_{LA} + Q_{GA}$$

then:

$$V_m = \frac{Q_T}{A_p}$$

Now in any multiphase system the actual velocities of the liquid and gas, V_L and V,, will be dependent on their respective volume flowrates and the fraction of the pipe area they occupy.

If the fraction of the pipe occupied by liquid and gas is H_L and H_G, then the actual liquid and gas velocities are given by:

$$V_L = \frac{QLA}{H_L\ A_p} = \frac{V_{SL}}{H_L}$$

$$V_G = \frac{Q_{GA}}{H_G.A_p} = \frac{V_{SG}}{H_G}$$

Now if the liquid and gas velocities are equal, we say there is no slippage between the phases:

$$V_L = V_G$$

and so:

$$\frac{V_{SL}}{H_L} = \frac{V_{SG}}{H_G} \quad \text{or} \quad \frac{V_{SL}}{H_L} = \frac{V_{SG}}{(1 - H_L)}$$

If there is no slippage between the phases the liquid fraction is termed the non-slip liquid hold up, λ_L. The above equation then becomes:

$$\frac{V_{SL}}{\lambda_L} = \frac{V_{SG}}{(1 - \lambda_L)}$$

rearranging we obtain:

$$A, = \frac{V_{SL}}{V_{SG} + V_{SL}}$$

thus:

$$\lambda_L = \frac{V_{SL}}{V_m}$$

Hence the non-slip hold-up of liquid is simply the ratio of superficial liquid velocity to mixture velocity.

1.1.7 Slippage and liquid hold-up

In most 2-phase pipelines the gas travels faster than the liquid:

$$V_G > V_L$$

Under these conditions there is said to be slippage between the phases. As the liquid velocity is less than it would be under non-slip conditions, the liquid hold-up, H_L, must increase to a value in excess of the non-slip value λ_L in order to maintain continuity:

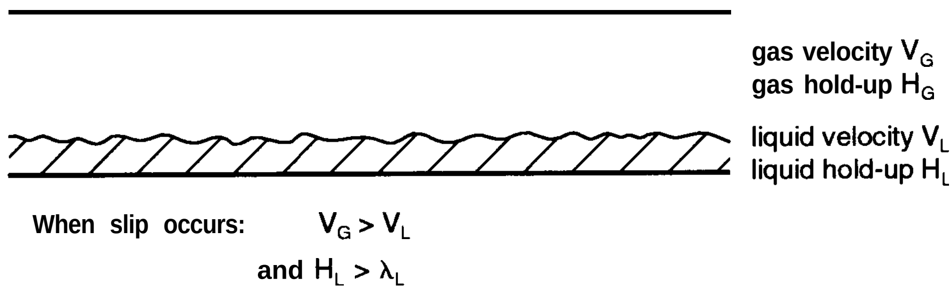


Figure 2.

Note that irrespective of slippage the in-situ liquid flowrate Q_{LA} is always given by:

$$Q_{LA} = V_{SL} \times A_p$$

Liquid hold-up H_L is an important parameter in 2-phase flow. The liquid hold-up at any condition of flow, pressure and temperature has a major influence on:

- The type of flow that will occur (i.e. flow regime, see Section 1.2)
- Pressure **drop**
- The slugcatcher volume required at the reception facilities to handle liquid swept out of the line by a pig.

The value of H_L is normally determined by use of an empirical correlation such as those of Beggs & Brill, and Eaton. The correlations typically provide H_L as a function of V_{SL}, V_{SG}, ρ_L and ρ_g. More information is given on hold-up correlations in Section 3.1.

1.1.8 Pressure drop

The total pressure drop over a pipeline system, ΔP_T is given by:

$$\Delta P_T = AP, + \Delta P_{eL} + \Delta P_{acc}$$

where:

AP, = frictional pressure loss
ΔP_{eL} = pressure **drop** caused by elevational changes
ΔP_{acc} = pressure drop required to accelerate fluids
ΔP_{acc} is usually insignificant and is not discussed here.

(a) Frictional pressure loss

In single phase flow frictional pressure drop is given by:

$$AP, = \frac{f\,L}{d}\,\frac{\rho\,u^2}{2}$$

where:

f = friction factor (Darcy-Weisbach)
L = length of pipe
d = pipe diameter
ρ = fluid density
u = fluid velocity

Note that the friction factor used throughout this manual is that of Darcy-Weisbach. This is 4 times greater than the fanning friction factor, i.e. in laminar flow f is given by:

$$f = \frac{64}{\text{Reynolds No.}}$$

Use of the Darcy-Weisbach friction factor has the advantage that the term f.L/d represents the number of velocity heads lost over any pipe length L.

Losses over fittings are commonly available in the form of velocity heads lost, or loss coefficients, K. Hence friction loss over a fitting is given by:

$$AP, = K\,\frac{\rho u^2}{2}$$

For homogeneous non-slip 2-phase flow (i.e. where liquid and gas are thoroughly mixed and flow at the same velocity) we could express the friction pressure loss as:

$$AP, = f_{ns}\,\frac{L}{d}\,\frac{\rho_{ns}\,V_m^{\,2}}{2}$$

where:

V_m = mixture velocity
ρ_{ns} = no slip mixture velocity

$$\rho_{ns} = \frac{\rho_L \times V_{SL} + \rho_g \times V_{SG}}{V_m}$$

where;

f_{ns} = friction based on a mixture Reynolds number, Re.

$$Re = \frac{V_m \times \rho_{ns} \times d}{\mu_{ns}}$$

where:

μ_{ns} = no slip mixture viscosity

$$\mu_{ns} = \frac{V_{SL} \times \mu_L + V_{SG} \times \mu_G}{V_m}$$

where:

μ_L = in-situ oil viscosity (live oil viscosity)
μ_G = in-situ gas viscosity

For truly homogeneous non-slip 2-phase flow the above equation will give a reasonable estimate of AP,. However, in most multiphase systems slippage occurs between the two phases. As a consequence energy is lost as a result of interfacial shear between the phases, and depending on flow regime, by other mechanisms. The overall effect is that the total frictional pressure loss is greater than for homogeneous conditions.

The frictional pressure loss in 2-phase flow is normally calculated using an equation of the form:

$$\Delta P_f = f_{tp} \frac{L}{d} \frac{\rho_{ns} V_m^2}{2}$$

where:

f_{tp} = 2-phase friction factor.

In most 2-phase correlations f_{tp} is calculated using the following equation:

$$f_{tp} = m_{tp}\ f_{ns}$$

where:

m_{tp} = an empirically determined Z-phase multiplier
f_{ns} = friction factor, normally calculated on basis of homogeneous flow

The various pressure drop correlations use different methods for determining f_{tp}. These are discussed in more detail in Section 3.1.

(b) Pressure drop due to elevational changes

The elevational pressure drop is given by:

$$\Delta P_{eL} = \rho_s\ L\ g\ \sin\theta$$

where:

L = segment length
g = acceleration due to gravity
θ = angle of segment to horizontal
ρ_s = in-situ density of 2-phase mixture (i.e. accounting for slip)

$$\rho_s = \rho_L\ H_L + \rho_g\ H_g$$

where:

H_L = in-situ liquid hold-up
H_g = in-situ gas hold-up $(1 - H_L)$

Clearly knowledge of liquid hold-up is vital in determining ΔP_{eL}

Section 3.1 of this manual discusses pressure drop and hold-up correlations in more detail and gives recommendations of the best correlations for crude oil-gas systems.

1.2 Flow Regimes

1.2.1 Horizontal and Near Horizontal Flow

When liquid and gas flow concurrently in a pipe they can distribute themselves into any one of a number of flow patterns depending on their respective flow rates, physical properties, pipe size and inclination. Figure 3 (below) shows the principal flow patterns, or flow regimes, occurring in horizontal or slightly inclined pipelines.

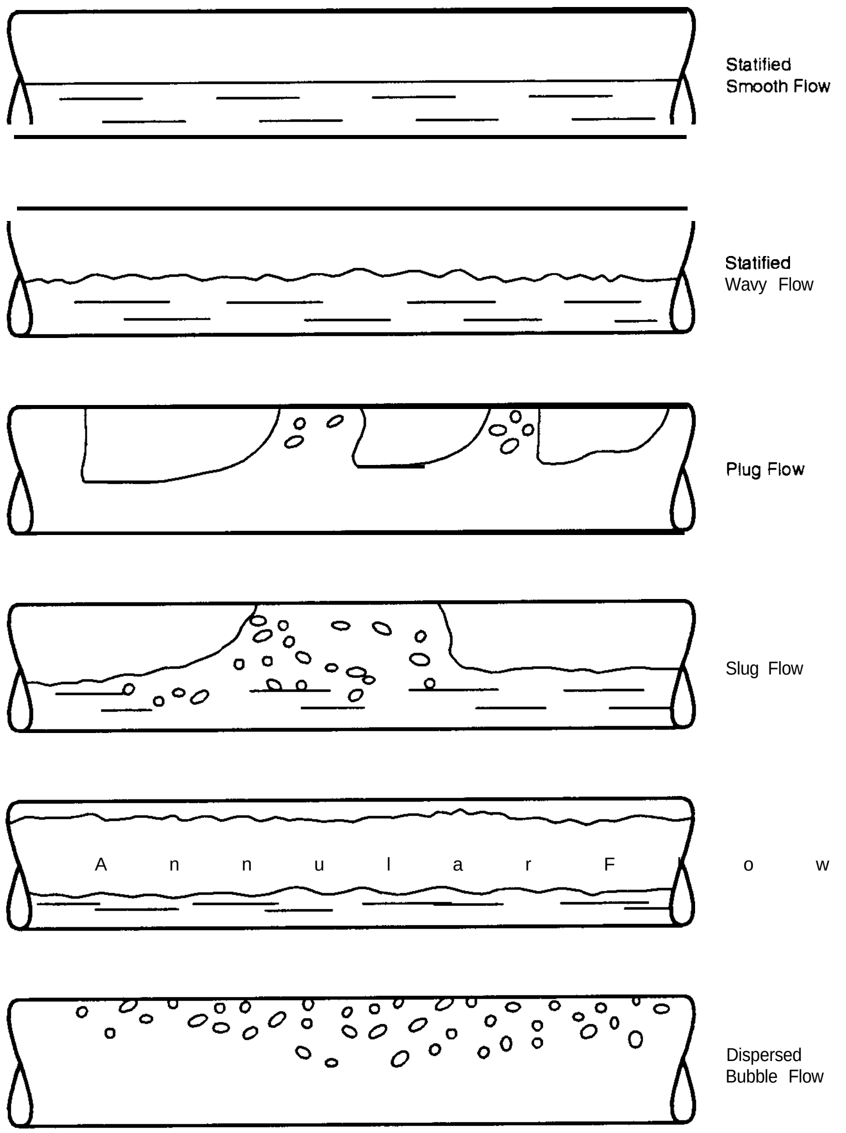


Figure 3. Flow regimes for a horizontal pipe.

These are briefly described below:

(a) Stratified Smooth Flow

At low gas and liquid flowrates the phases segregate with the liquid flowing along the bottom, and the gas flowing through the upper part of the pipe. The interface between the phases is smooth. This type of flow rarely occurs in the field.

(b) Stratified Wavy Flow (Wavy Flow)

As the gas velocity is increased waves form on the liquid surface. The liquid partly climbs the pipe wall to form a crescent.

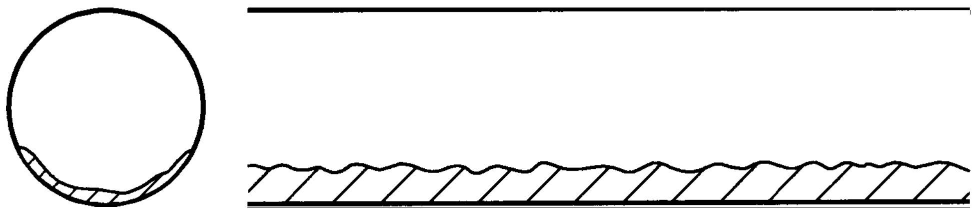


Figure 4. Waves on surface. Liquid climbs pipe wall to form a crescent.

This type of flow is common in gas-condensate systems.

The wave height, and level to which the liquid film climbs the surface of the pipe, increase as the gas velocity is raised. Some liquid is also torn off from the waves to form droplets which travel for a distance in the gas core before being deposited back into the liquid flow.

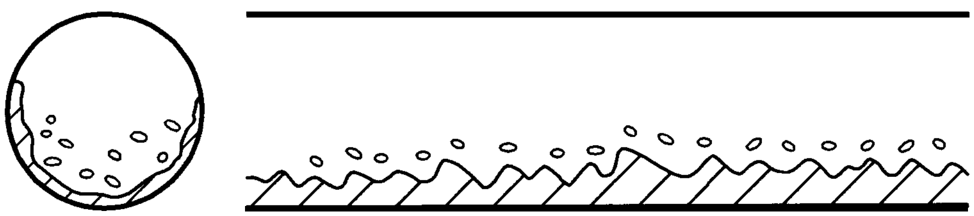


Figure 5. Liquid climbs further up wall. Some liquid entrained in gas core

(Note that in a horizontal line stratified smooth and stratified wavy flow give rise to approximately steady production of gas and liquid. However, where the line undergoes changes in inclination, and where there is a sufficient quantity of liquid present, the liquid accumulation in low points of the system may produce a blockage causing the flow to stall. The liquid will then be intermittently expelled from the dip as a slug. This condition, commonly referred to as surging, is not normally experienced in low liquid loading gas-condensate systems, but can occur in low velocity crude oil-gas systems such as at Wytch Farm and Welton).

(c) Annular Flow (Annular-Mist)

As the gas velocity is further increased there comes a point where the liquid film forms a complete annular ring around the surface of the pipe. The liquid film is obviously thickest at the bottom of the pipe. Some liquid is entrained as a mist in the gas core.

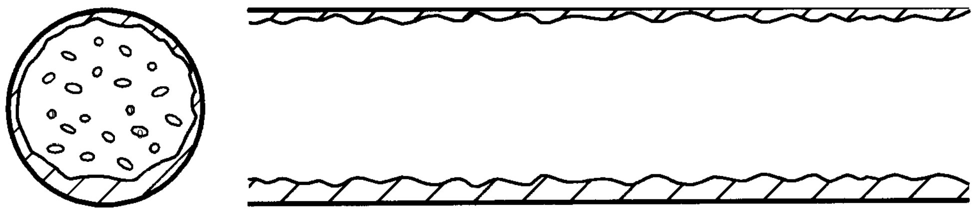


Figure 6. Liquid forms complete annular ring, thickest at bottom. Some liquid entrained in core

This type of flow can occur in gas-condensate systems at the top end of their throughput range. The onset of annular flow marks the transition to a condition where the pipe wall is fully wetted. This transition to annular flow thus has implications for corrosion inhibition.

(d) Plug Flow (Elongated Bubble)

Plug flow occurs at low gas and moderate liquid velocities and is characterised by bubbles of gas distributed in a liquid continuum. Plug flow can occur in low GOR crude oil-gas systems, such as the Wytch Farm Bridport flowlines, and also in moderate to high GOR systems when operated at high pressures.

This type of flow does not normally cause any process upsets as the gas plugs are relatively short and are produced regularly.

(e) Slug Flow

Slug flow occurs over a wide range of conditions and is hence often encountered in multiphase lines. In slug flow the fluids are ordered as alternate slugs of liquid and bubbles of vapour.

The variation in liquid and gas production rates associated with slug flow can cause upsets at the process plant. The extent of the upsets will depend to a large extent on the length of the slugs. Note that plug and slug flow are often jointly termed intermittent flow.

(f) Dispersed Bubble Flow

Dispersed bubble flow occurs at high flowrates when the flow induced turbulence causes the liquid and gas to become well mixed. This type of flow is sometimes referred to as froth flow.

There is some evidence available that suggests that foaming agents can be used to change a slug flow condition to frothy dispersed bubble flow by altering the gas/liquid surface tension. Hence the chemical nature of the fluids can actually have a bearing on the flow regime by interacting with the physical forces which tend to separate the liquid and gas.

The transition between each flow regime is usually presented in the form of a two-dimensional flowmap, see Figure 7 (below). The usual co-ordinates of such a map are superficial liquid velocity (effectively liquid volume flowrate) and superficial gas velocity.

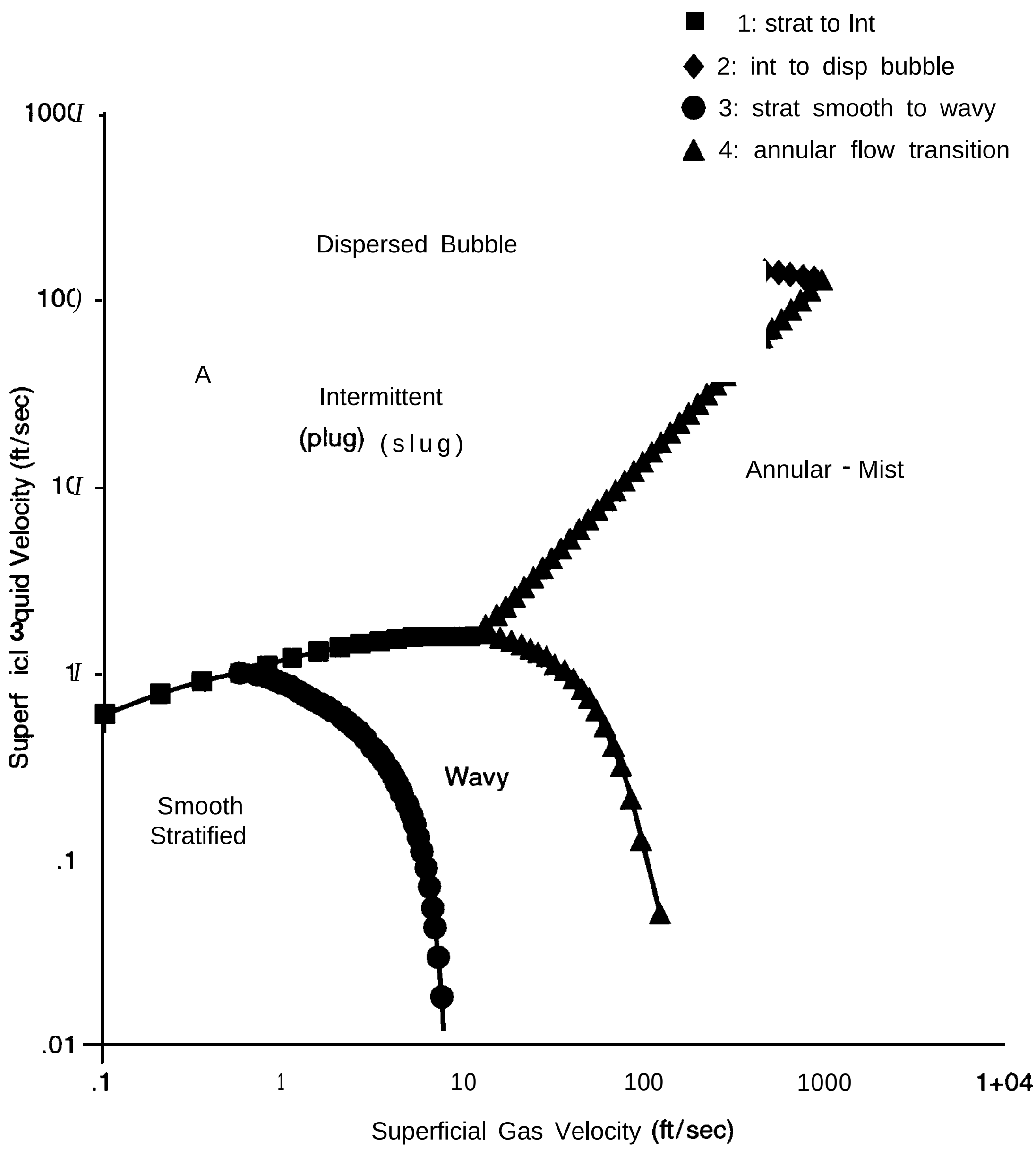


Figure 7. Taitel-Dukler horizontal flowmap.

The flowmap shown in Figure 7 (above) was produced by the industry standard method of Taitel-Dukler. BP's field experience has shown discrepancies between Taitel-Dukler predictions and measured data. BP Exploration is developing new mechanistic models for determining flow regimes. Work on the transition marking the onset of wall wetting is now complete and this model should now be used in design.

1.2.2 Vertical Flow

in vertical flow gravity acts to produce symmetrical flow about the centre-line so that there is no equivalent of stratified or wavy flow.

Figure 8 (below) shows the principal flow regimes occurring in vertical flow.

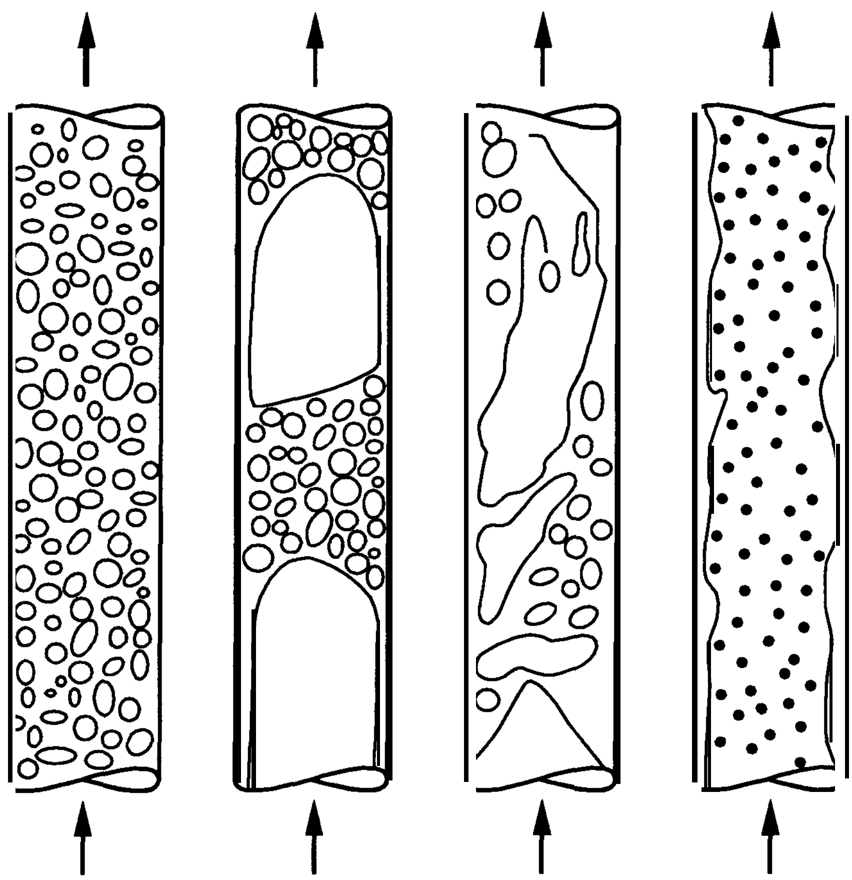


Figure 8. Flow patterns for upward vertical flow.

(a) Bubble Flow

In bubble flow the gas phase is distributed more or less uniformly in the form of discrete bubbles in a continuous liquid phase.

(b) Slug Flow

In the slug flow pattern the two fluids redistribute axially so that at any cross-section the flow rates of liquid and gas vary with time. The gas flows largely in a "Taylor bubble", which occupies most of the pipe's cross-sectional area and can vary in length from the tube diameter to over a hundred diameters. Between the Taylor bubble and the wall, a thin liquid film flows downward.

(c) Churn Flow

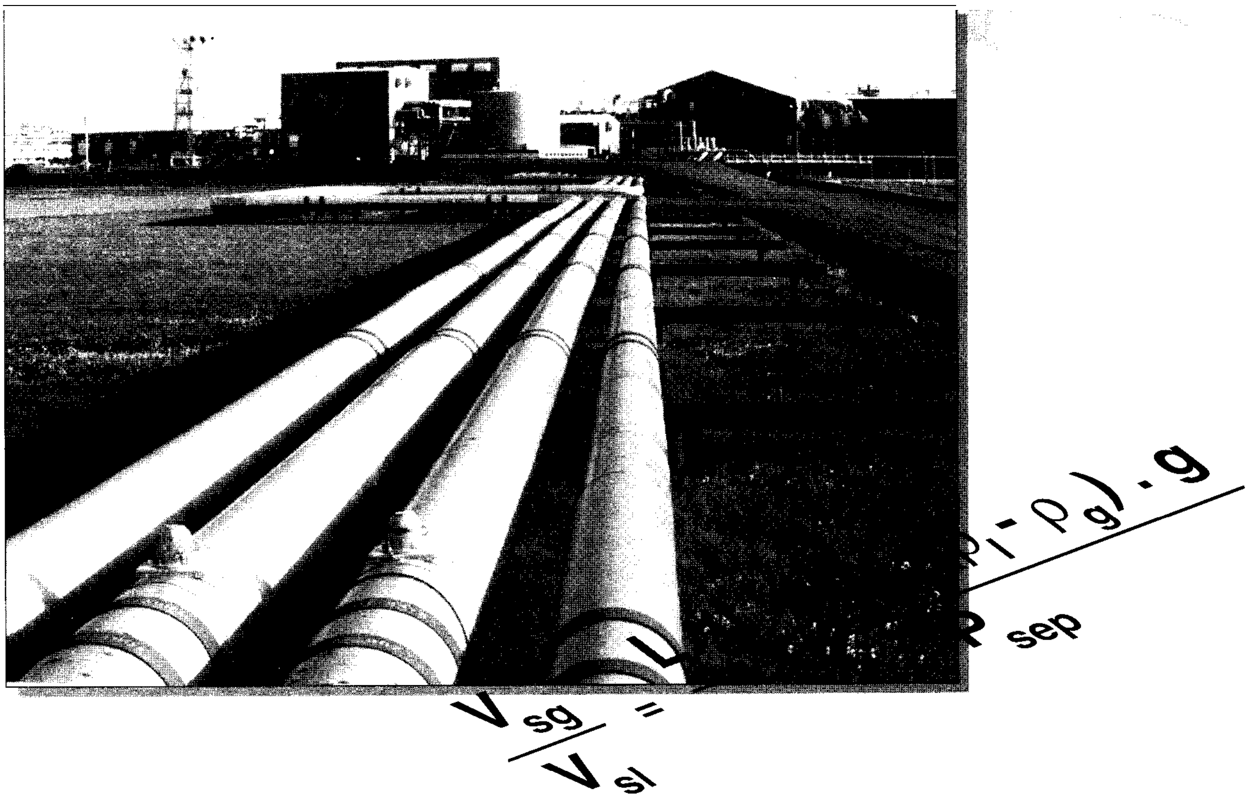
Churn flow is somewhat similar to slug flow. It is, however, much more chaotic, frothy, and disordered. The bullet-shaped Taylor bubble becomes narrow, and its shape is distorted. The continuity of the liquid in the slug between successive Taylor bubbles is repeatedly destroyed by a high local gas concentration in the slug. As this happens and the liquid falls, this liquid accumulates, forms a bridge, and is again lifted by the gas.

(d) Annular Flow

Annular flow is characterised by continuity in the axial direction of the gas phase in the core. Liquid flows upward, both as a thin film and as droplets dispersed in the gas. Except at the highest flow rates, the liquid appears to also flow as large fast- moving lumps that are intermittent in nature - either large roll waves travelling over the film or a high local concentration of droplets.

Section 2. Physical Properties

Prediction of Gas and Liquid Physical Properties



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2.1 Introduction

Multi-component phase behaviour is a complex phenomenon which requires accurate determination if two-phase pressure loss, hold-up and flow regime are to be determined with any degree of confidence.

In practice, experimentally determined phase behaviour is often scarce and one has to resort to some method of prediction. There are two approaches commonly employed in the prediction of hydrocarbon phase behaviour. These are the “black oil” method, which assumes that only two components i.e. gas and liquid, make up the mixture, and the so called “compositional” approach in which each hydrocarbon component is taken into account. The methods have their own relative merits and are discussed below.

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2.2 Compositional Models

Hydrocarbon mixtures display a phase behaviour that is generally represented in the form of a P-T **phase** diagram, thus:

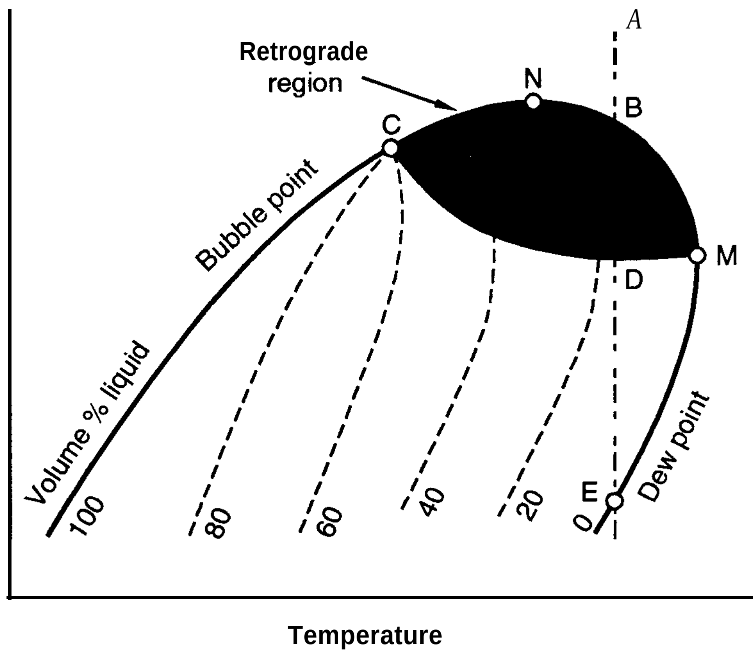

Figure 1.

Such a diagram is produced for one particular mixture composition. Phase diagrams are determined by use of equilibrium vapour ratios, K (or K values). The K value is defined as:

$$K_i = \frac{y_i}{x_i}$$

where

y_i = mole fraction of component i in the vapour phase
x_i = mole fraction of component i in the liquid phase

The K value for any component is dependent on temperature and pressure, and also, at high pressures, on the fluid composition.

There are several terms used to define the location of various points on the phase envelope.

Cricondenbar

Maximum pressure at which liquid and vapour may co-exist in equilibrium (point N).

Cricondentherm

Maximum temperature at which liquid and vapour may co-exist in equilibrium (point M).

Retrograde Region

The area inside the phase envelope where condensation of liquid occurs on lowering pressure or increasing temperature.

Quality Lines

Those lines showing constant gas volume weight or mol percentages. These lines intersect at the critical point (C) and are essentially parallel to the bubble and dew point curves.

Line ABDE on the above diagram represents a retrograde condensation process occurring, for example, in a gas-condensate transport line as pressure drops due to friction. Point A represents the single phase fluid outside the phase envelope. As pressure is lowered, point B is reached where condensation begins. As pressure is lowered further, more liquid forms because of the change in the slope of the quality lines. The retrograde area is governed by the inflection points of such lines. As the process continues outside the retrograde area, less and less liquid forms until the dew point is reached (point E). Below E no liquid forms.

In a compositional model the predictions of gas and liquid physical properties are performed through use of an equation of state, EOS. Any equation correlating pressure (P), volume (V) and temperature (T) is known as an EOS. For an ideal gas the EOS is simply :

$$PV = n\ RT$$

where:

n = number of moles of gas
R = Universal gas constant.

To account for the non-ideality of most practical systems the above equation is modified to include various correlating constants. The Peng-Robinson (PR) EOS, for example, is given by:

$$P = \frac{RT}{V - b} - \frac{a(T)}{V(V+b) + b(V-b)}$$

where for any mixture :

$b = \Sigma\ y_i\ b_i$
y_i = mole mole fraction of component i
b_i = empirical constant for component i
$a = \Sigma_i\ \Sigma_j\ y_i\ y_j\ (1 - BN_{ij})\ a_i\ a_j$
BN_{ij} = empirically determined interaction parameter for the two components i and j.
a_i, a, = empirical constants for components i and j. These are a function of temperature.

A number of data generation methods are available in BP's flowsheeting program, GENESIS. When running BP's multiphase flow program MULTIFLO, with a compositional description of the fluids, the GENESIS module FLODAT is first run to produce the phase envelope and physical property data. When running FLODAT the user is asked for two data generators.

1. The first is for K value determination, i.e. for generating vapour liquid equilibrium data.

2. The second is for the generation of thermal properties and densities.

   In the absence of other recommendations, the most suitable data generators for typical gas-condensate systems are:

1. iP-R (Peng-Robinson with interactive parameters) for K values

2. Z-J (Zudkvitch-Joffe) for thermal properties and densities.

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2.3 Black Oil Models

The black oil approach to the prediction of phase behaviour ignores the fluid composition and simply considers the mixture as consisting of a gas and liquid phase in which the gas may be dissolved in the liquid. The basic assumption of a black oil model is that increasing system pressure (and reducing temperature) cause more gas to dissolve in the liquid phases, and, conversely, decreasing system pressure (and increasing temperature) cause gas to evaporate from the liquid phase. It was noted in Section 2.2 that retrograde condensation involves the conversion of gas to liquid on reducing pressure. This is contrary to the fundamental assumption of the black oil model and so the black oil approach is only valid for systems operating at conditions far removed from the retrograde region.

Consider the diagrams below which illustrate a typical expansion process from a reservoir A to a separator B for:

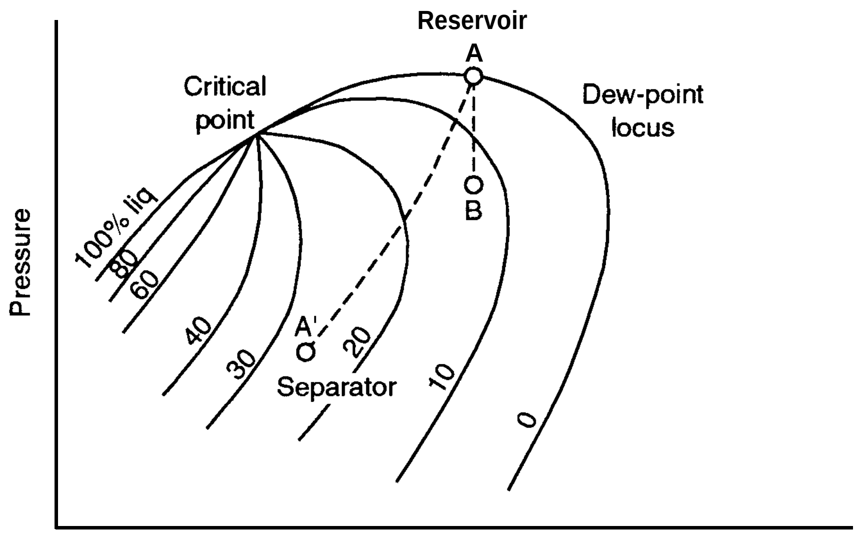

Figure 2. A gas reservoir

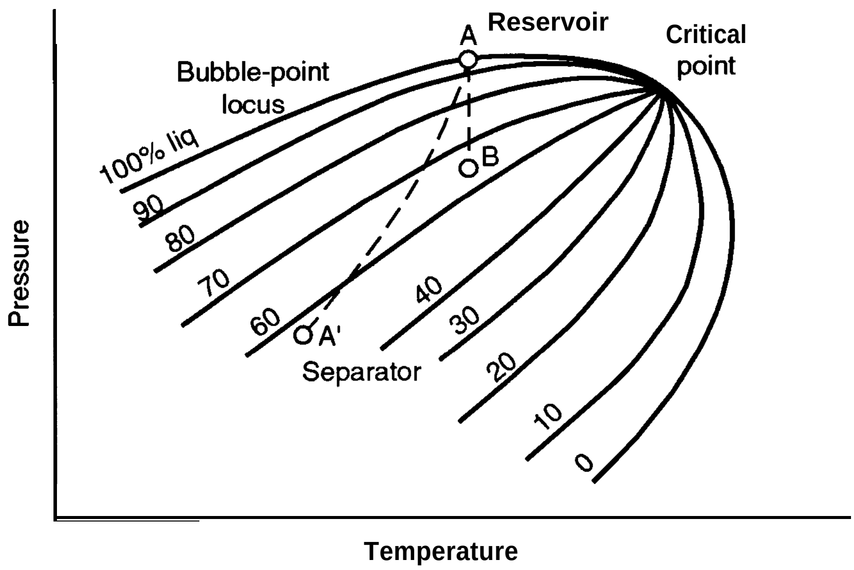

Figure 3. A crude oil reservoir

For the liquid reservoir, lower diagram, the reservoir condition is well to the left of the critical point and so the expansion process involves the continual evolution of gas, i.e. the operating point moves steadily across the quality lines to a condition of ever decreasing liquid content. This type of process would be adequately represented by a black oil model.

For the gas reservoir, upper diagram, the reservoir condition lies to the right of the critical point so that on expansion, (reducing pressure) the operating point moves across the quality lines to a condition of increasing liquid content, i.e. retrograde condensation. This process could not be represented by a black oil model.

As a general guide a black oil model should be adequate for describing crude oil-gas systems, while a compositional model is necessary to describe wet-gas, gas-condensate and dense phase systems.

The black oil model employs certain concepts and nomenclature which require definition. These are discussed briefly below:

2.3.1 Producing Gas Oil Ratio (GOR)

This is the quantity of gas evolved when reservoir fluids are flashed to stock tank conditions. The units are standard cubic feet of gas per stock tank barrel of oil (SCF/STB) measured at 14.7 psia and 60°F.

The GOR of a crude is obtained by experimental testing. However, the GOR will vary depending on how many flash stages are employed to get down to stock tank conditions. The normal convention is to calculate GOR from the sum of gas volumes evolved from a multistage flash procedure (normally this involves 2 or 3 flash stages). This more closely represents conditions in the field with the pressure and temperature conditions chosen for the first stage flash approximating to conditions likely to be experienced in the first stage separator in the field.

2.3.2 Solution Gas Oil Ratio (Rs)

This is the quantity of gas dissolved in the oil at any temperature and pressure. It represents the quantity of gas that would be evolved from the oil if its temperature and pressure were altered to stock tank conditions, 14.7 psia and 60°F. Hence, by definition the Rs of stock tank oil is zero.

The Rs crude at its bubble point is equal to the producing GOR of the reservoir fluids.

The volume of free gas present at any pressure and temperature is the difference between the GOR and the Rs. The volume of free gas is corrected for pressure,temperature and compressibility to compute the actual in-situ volume of gas and hence superficial gas velocity. Rs can be evaluated from standard correlations such as Glaso or Standing. These correlations require as input the oil and gas gravity and the pressure and temperature conditions.

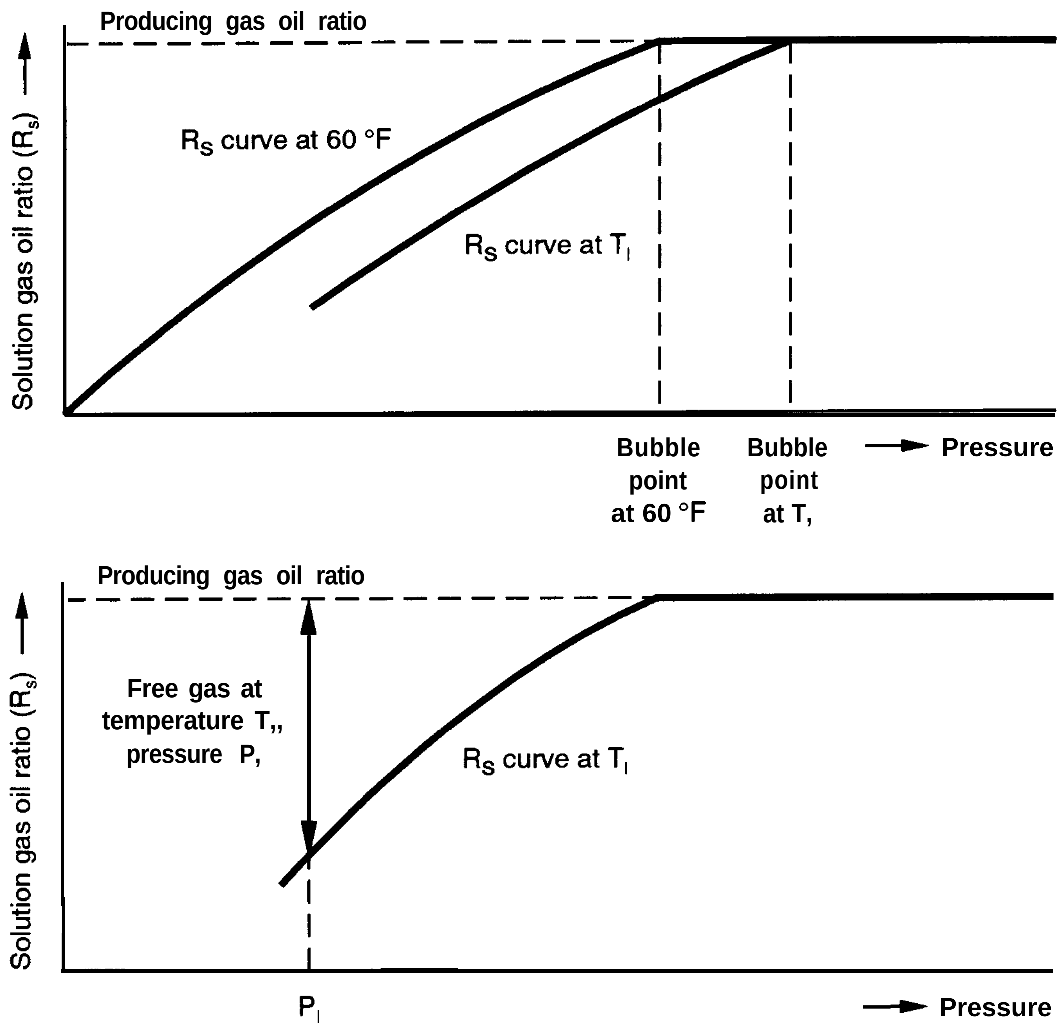

Figure 4.

2.3.3 Volume Formation Factor (Bo)

The volume formation factor is the ratio of the volume occupied by oil at any pressure and temperature to the volume occupied at stock tank conditions. The units are pipeline barrels per stock tank barrel (pbbl/STB). The volume formation factor of stock tank oil is thus 1 .O. Through use of Bo the volume flowrate and density of the liquid phase can be calculated, (see Section 1.1).

Standard correlations are available to compute Bo. These require as input the oil and gas density, the Rs of the liquid at the conditions of interest, and the pressure and temperature.

2.3.4 Live Oil Viscosity (Ov)

The viscosity of the oil in a 2-phase pipeline depends on the stock tank oil viscosity (dead oil viscosity), the solution gas oil ratio at the conditions of interest, and the pressure and temperature. Correlations are available to compute the live oil viscosity.

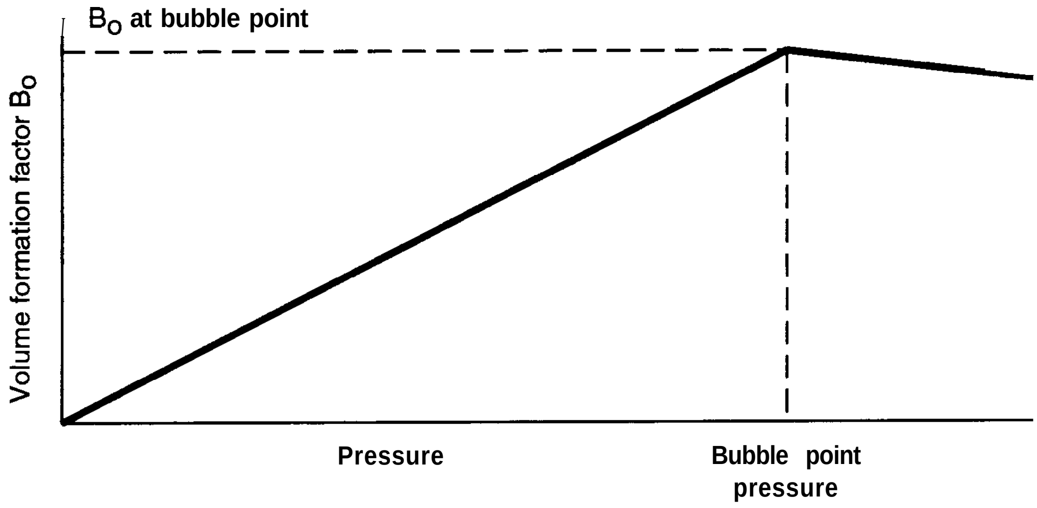

Figure 5.

The correlations available for Rs, Bo and Ov will yield approximate values only and where laboratory or field data is available, these should be used to adjust and tune these correlations. The way in which the correlations are tuned will depend on the quantity of field data available.

Details of the way in which field or laboratory data are used to tune the correlations is given in Appendix 8C of the MULTILFO User Manual.

The minimum physical property information required to run a black oil model is:

(a) Stock tank oil gravity

(b) Gas gravity

There is often some confusion about the definition of gas gravity and hence uncertainty about the value of this data item.

The majority of the correlations provided in MULTIFLO are based upon multistage 2-3 stages) separation, and the gas gravity used should always be the total gravity based upon the weighted average gravity from each stage, viz:

$$\textbf{Total gas gravity} = \frac{i = 1,n\ \sum(sgi*Gi)}{i = 1,n\ \sum Gi}$$

where:

sgi = gravity of the ith separator stage off-gas
Gi = free gas GOR at the ith separator stage
n = number of stages in the separator train with the final stage at stock tank conditions.

(c) Total producing GOR

This should be taken as the sum of the gas volumes evolved from each stage of a multistage flash.

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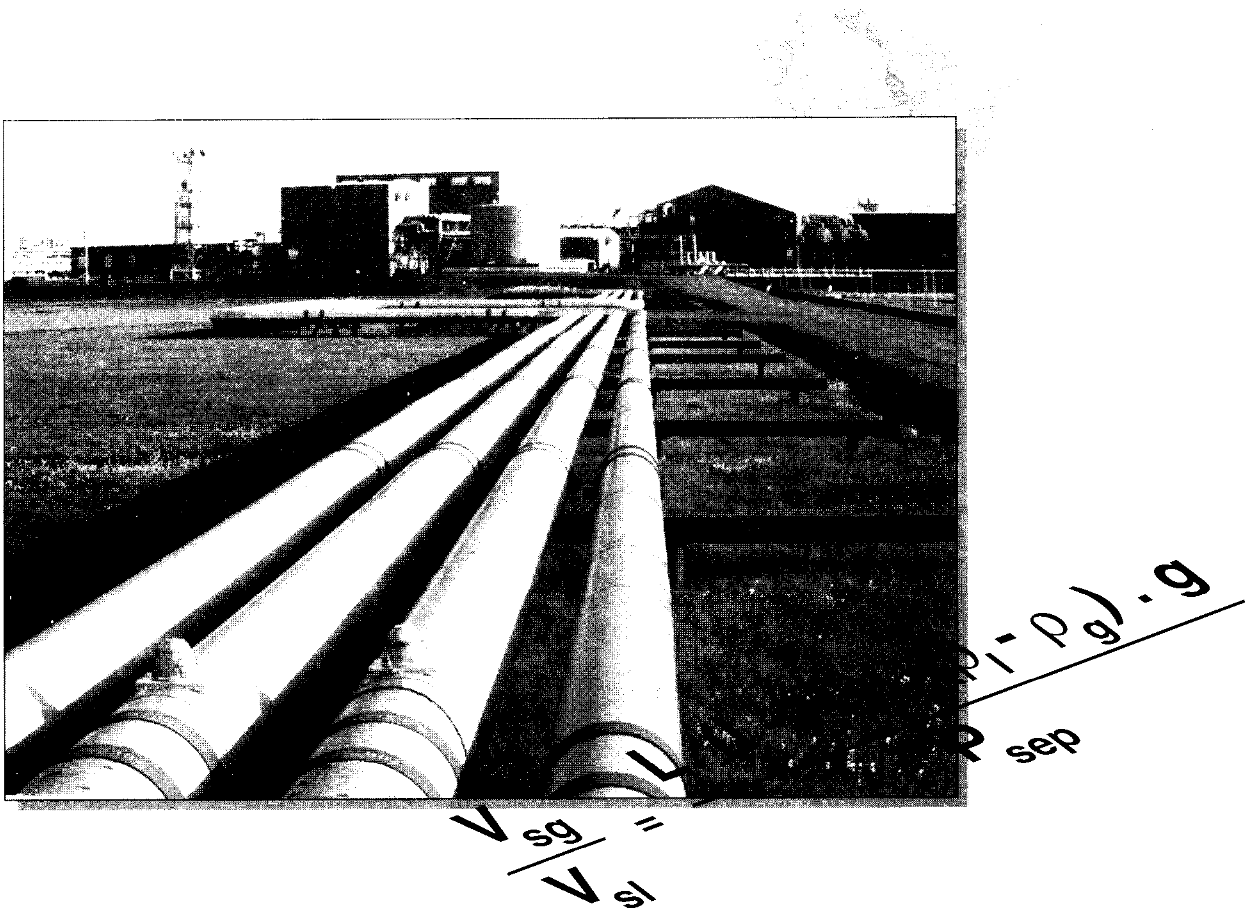
Section 3. Oil / Gas Pipeline Design

Design of Multiphase Crude Oil and Gas Pipeline Systems



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3 Design of Multiphase Crude Oil / Gas Pipeline Systems

The purpose of this section is to detail the various factors that have to be considered when designing multiphase crude oil/gas pipeline systems. Such pipelines may be onshore, such as at Wytch Farm and Prudhoe Bay, or offshore, such as the SE Forties inter-platform line, and the Magnus satellite well flowlines.

3.1 Line Sizing and Pressure Drop

The first stage of any design study is the specification of a line size to provide the required capacity within the available pressure constraints. However, in addition to sizing a line for the available pressure drop, consideration must be given to ensuring that the flowing velocities provide reasonable operating conditions.

A multiphase pipeline is generally used to transport total wellhead fluids (oil, gas and water) from a wellsite, or a minimum facilities platform (MFP), to a remote processing plant. The pressure at the processing plant is normally known, i.e. the separator pressure. There is generally some information available on the pressure available at the wellhead, or WHFP, from a wellhead pressure study. Hence the pressure drop available for the pipeline is generally known. Note that as the study progresses into detailed design, the wellhead pressure data may be refined and may become available on a year-by-year basis. It may also be possible, and is desirable, to base pressure drop predictions on the bottom hole flowing pressure (BHFP) and the separator pressure as the fixed points in the system.

A line size has to be chosen which will yield a pressure drop equal to, or less than, that available. The total pressure drop, ΔP_T, is given by:

$$AP, = \Delta P_f + \Delta P_{el} + \Delta P_{acc}$$

where:

AP, = pressure drop due to friction
AP,, = elevational pressure drop (hydrostatic head loss)
AP, = pressure drop to acceleration of the fluids

AP, is normally insignificant and will not be discussed here. (The ΔP_{acc} term should always computed in multiphase simulators such as MULTIFLO but only becomes significant for high velocity flows approaching critical conditions).

3.1.1 Frictional Pressure Drop ΔP_f

AP, is given by:

$$AP, = f_{tp} \, L/D \, \rho_{ns} \, Vm^2/2$$

where:

f_{tp} = 2-phase friction factor

L = line segment length (m)
D = line diameter (m)
ρ_{ns} = no-slip density (defined in Section 1 .1) (kg/m^3)
V_m = mixture velocity (defined in Section 1.1) (m/s)

f_{tp} is given by:

$$f_{tp} = f_{ns}\, m_{tp}$$

where:

f_{ns} = no slip friction factor (Darcy-Weisbach) based on homogeneous 2-phase physical properties and Reynolds number (see Section 1 .1).

m_{tp} = a dimensionless 2-phase multiplier used to account for slip between the phases.

Most of the commonly used 2-phase pressure drop methods compute mtp as a function of the calculated in-situ liauid hold-ub. H_L, (for definition of H_L see Section 1 .1). The Beggs and Brill method calculates mtp as follows:

$$m_{tp} = e^s$$

where:

$$s = \frac{\ln(y)}{(-0.052 + 3.18 \ln(y) - 0.872 [\ln(y)]^2 + 0.0185 [\ln(y)]^4\}}$$

$y = \lambda_L / H_L^2$

λ_L = no slip hold-up, $\dfrac{V_{sl}}{V_{sl} + V_{sg}}$ (see Section 1 .1)

H_L = liquid hold-up

The Beggs and Brill correlation for H_L in a horizontal line gives:

$$H_L = \frac{a\, \lambda_L^b}{Fr^c}$$

where:

Fr = Froude number, V_m^2 / gD

a, b, and c are constants provided for each of the 3 flow regimes recognised by the Beggs & Brill method; segregated, intermittent, and distributed.

The Beggs and Brill method also has a term to correct hold-up for the effect of inclination:

$$H_L(\theta) = \varphi H_L$$

where:

φ = is provided as a function of inclination (θ) and flow regime.

Full details of the Beggs and Brill pressure drop and hold-up correlations, and other commonly used correlations, are given in (1).

3.1.2 Elevational Pressure Drop (P_{el})

Once the liquid hold-up, H_L, has been evaluated the elevational pressure drop term is simply calculated from :

$$\Delta P_{el} = g\, h\, [\rho_l H_L + \rho_g(1 - H_L)]$$

where:

AP,, = elevational pressure drop (N/m*)
g = acceleration due to gravity (9.81 m/s')
h = height increment (m)
ρ_l = in-situ liquid density (kg/m^3)
ρ_g = in-situ gas density (kg/m^3)

3.1.3 Recommended Pressure Drop Methods

Multiphase design computer programs, such as MULTIFLO, generally provide a wide range of the most commonly used correlations for pressure loss and hold-up.

The Multiphase Flow group have tested the accuracy of these correlations using field data collected from a number of sites, (2), (3) and (4).

The Beggs and Brill pressure drop method incorporating the Beggs and Brill hold-up correlations, was generally found to be the most reliable of the commonly used methods. It normally over predicts pressure drop by between 0–30%, and hence it provides a degree of conservation.

However, for hilly terrain pipeline systems the situation is less clear. For the Wytch Farm in-field lines, the original Beggs and Brill method was found to under predict pressure loss. Studies revealed that this was due to an over-prediction of pressure recovery in the downhill sections, (3). A modified version of Beggs and Brill has been implemented in MULTIFLO which calculates elevational pressure drop in downward sloping sections assuming gas head recovery only (i.e. it neglects liquid head recovery). This modified version of Beggs and Brill gave very good agreement with data collected at Wytch Farm.

Data collected from the Cusiana Long Term Test (LTT), and the Kutubu hilly terrain pipelines indicate that pressure recovery in downward sloping lines is a complex function of flow regime and flowrate. A new computer model is currently being developed and tested to enable pressure loss to be more accurately evaluated for hilly terrain pipeline systems. The model tracks slug lengths through the pipeline and by taking account of slug decay in downhill sections it can provide a more realistic estimate of the pressure recovery term. This development is discussed further in Appendix 3D.

Data collected on the Forties Echo - Forties Alpha (FE - FA) inter-platform multiphase lines suggested that the best correlations to use for flow in the risers are:

- Upflow Risers – Orkiszewski
  (generally over-predict pressure loss)

- Downflow Risers – homogeneous
  (tends to slightly under-predict pressure recovery)

In summary, the recommended pressure drop correlations for multiphase crude oil/gas lines are:

(a) Horizontal and near horizontal lines
Modified Beggs and Brill, gas head recovery only in downhill sections. For hilly terrain pipelines the most accurate determination of pressure loss will be obtained using the new BP model which is currently under development and test.

(b) Upflow Risers – Orkiszewski

(c) Downflow Risers – homogeneous

A recent study conducted by the National Multiphase Flow Database Project tested the accuracy of the most commonly used pressure drop correlations against crude oil/gas field data supplied by members, (5). It concluded that the Orkiszewski vertical flow correlation in combination with the Beggs and Brill correlation for horizontal flow were the most consistently accurate pressure drop methods available. Hence apart from the hilly terrain pipeline considerations (and the National Database has little hilly pipeline data) the Database Project's findings are in line with our own conclusions.

3.1.4 Flowing Velocities

Having selected a line size to meet throughput and pressure drop constraints, it is important to check whether acceptable flowing conditions exist within the pipeline.

The pressure gradient in a crude oil/gas multiphase line normally lies in the range 0.2-2.0 psi/100 ft. However, a more useful means of assessing whether flowing conditions are acceptable is to check whether fluid velocities lie within certain limits.

At the low end of the normal throughput range the actual (not superficial) liquid velocity, V_L, should ideally be greater than ca. 3 ft/s. This will ensure that sand and water are continuously transported with the liquid phase and not allowed to drop out and accumulate at the bottom of the pipe. Details of the calculation procedure recommended to accurately assess the critical sand transport velocity will be given in Section 12.

Note that the actual liquid velocity, V_L, is given by:

$$V_l = V_{sl} / H_L$$

where:

V_{sl} = superficial liquid velocity
H_L = liquid hold-up.

At the maximum throughput conditions the mixture velocity, V_m should not exceed the erosional velocity limit, V_e.

The current industry standard method of determining Ve is through use of the relationship given in API 14E:

$$V_e = C / \sqrt{\rho_m}$$

where:

V_e = maximum acceptable mixture velocity to avoid excessive erosion (ft/s)
ρ_m = no-slip mixture density (lb/ft^3)
C = a constant, given in API 14E as 100 for carbon steel

This relationship is considered to provide an inadequate description of the erosion process. R&D is currently underway to provide a more satisfactory means of determining V_e.

A review by the Materials and Inspection Engineering Group, ESS, advises that provided that the fluid does not have a high sand content, a value of C equal to 135 can be used for carbon steel lines, (6).

Materials and Inspection Engineering Group, MIE, have recently issued a comprehensive guideline on erosion in single and multiphase pipelines, (10).

In some systems flowing velocities need to be limited by a requirement to avoid removal of corrosion inhibitor. MIE recommend a maximum wall shear rate to avoid inhibitor stripping of 100 N/m^2, (11).

For duplex stainless steel lines a C factor of 236 is recommended. It is pointed out that these modified C factors were based on limited data and that critical areas of pipework should always be regularly monitored using NDT and corrosion probes as appropriate.

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3.2 Temperature Drop and Temperature Related Problems

3.2.1 General

In most multiphase simulators, such as MULTIFLO, heat transfer rates are calculated using homogeneous mixture properties. Hence no attempt is made to account for the position of the liquid and gas in the pipe. This approach has been found to be satisfactory for multiphase pipelines operating under normal transportation conditions. For example, good agreement was found between predicted and actual temperature drop for the FE-FA 6" and 12" flowlines.

In some circumstances the homogeneous heat transfer model may not be adequate. Hence when investigating heat flow from a pipe, and valves, to a multiphase fluid during a rupture, knowledge of the location of the liquid film may be of great importance. Under these circumstances it would be necessary to use a sophisticated transient model.

3.2.2 High Temperature Related Problems

Transportation of high temperature wellhead fluids (in excess of ca. 90°C) may give rise to problems in meeting the maximum temperature specification of the pipe coating material. Approximate maximum temperatures for various pipe coatings are given below.

Coating	Maximum Temperature ("C)
FBE (fusion bonded epoxy)	90
Coal Tar Enamel (coal tar wrap)	
(i) on subsea lines	70
(ii) on lines buried on land or at sea	50
Neoprene	ca. 100

The coating specialists in the Materials and Inspection Engineering Group should be consulted if there is any possibility of operating near the maximum allowable coating temperature. This group can also advise on alternative coatings for any particular application.

CO, corrosion of pipelines increases with temperature but reaches a maximum at about 60°C. In some cases it is necessary to cool the fluids prior to entry to the pipeline to reduce corrosion rates to a level consistent with the use of carbon steel pipework.

An alternative to providing cooling offshore is to use corrosion resistant pipe for the initial sections of the pipeline. At some distance downstream of the start of the line natural cooling in the sea will reduce fluid temperatures to a level where carbon steel pipe can be used.

The Materials and Inspection Engineering Group, ESS, should be consulted for advice on corrosion rates, corrosion inhibitor policy, and acceptable line inlet temperatures.

3.2.3 Low Temper&we Related Problems

The main areas of concern with regard to low temperature operation of multiphase lines are:

- **Wax deposition**
- **Hydrate formation**
- **High viscosities, in some crudes, and under some circumstances, in oil/water emulsions.**

(a) Wax Deposition

In subsea and onshore multiphase pipelines, fluid temperatures may drop below the wax appearance point with the result that small particles of wax form. Some of these wax particles are carried along in the liquid and some are deposited on the pipe walls. In fact because the walls of the pipeline are generally colder than the bulk fluid, wax deposition onto the pipe wall may occur while the bulk pipeline fluids temperature is still somewhat above the wax appearance point.

The pipe wall temperature is not printed in the MULTIFLO output, but can be simply calculated in the following way.

For each pipe section MULTIFLO prints out the following heat transfer coefficients (Btu/hr °F ft^2).

INTFLM – internal film coefficient
EXTMED – external film coefficient
INSLTN – heat transfer coefficient for the pipe coating or insulation
PIPEWL – pipe wall heat transfer coefficient
RADTN – contribution to heat transfer from radiation
OVRALL – overall heat transfer coefficient

(All these heat transfer coefficients are based on the pipe internal diameter)

When radiation effects are negligible the overall heat transfer coefficient is given by:

$$\frac{1}{\text{OVRALL}} = \frac{1}{\text{INTFLM}} + \frac{1}{\text{EXTMED}} + \frac{1}{\text{INSLTN}} + \frac{1}{\text{PIPEWL}}$$

When radiation contributes to heat transfer, OVRALL is given by:

$$\text{OVRALL} = \frac{1}{(1/\text{INTFLM}) + (1/\text{EXTMED}) + (1/\text{INSLTN}) + (1/\text{PIPEWL})} + \text{RADTN}$$

The heat flow through the pipe can be calculated from:

$$H = U\,A\,(T_i - T_o)$$

where:

H = heat flow (Btu per hr per foot length of pipe)
U = overall heat transfer coefficient, OVRALL (Btu/hr °F ft^2, based **on** pipe id)
A = surface area of pipe ID (ft^2 per foot length of pipe)
T_i = bulk temperature of fluid in pipe (“F)
T_o = bulk temperature of fluid external to pipe (°F)

Through any section of pipe:

$$H = h A (T_a - T_b)$$

where:

h = heat transfer coefficient for layer or film (EXTMED, etc.)
T_a = temperature of one side of layer
T_b = temperature at other side of layer

If we draw a section of the pipe as:

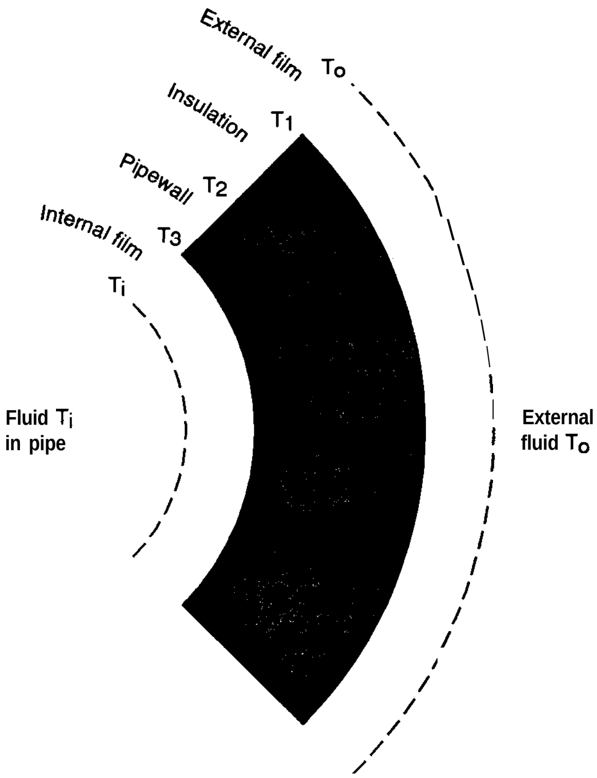

Figure 1. Pipe cross-section

Then starting from outside of pipe, we have:

$$T_, = T_o + \frac{H}{\text{EXTMED } A}$$

where:

T, = temperature at outside surface of insulation

The temperature at inside surface of the pipe wall, T_3, is given by:

$$T_3 = T_i - \frac{H}{INTFLM\ A}$$

It is important to be able to predict wax deposition rates as this may have a significant effect on pipeline operability. Wax deposition may increase to the point where it affects pressure drop. If wax is allowed to deposit in a line over a prolonged period intelligence pigging (for inspection) may become a risky operation because of the danger of blocking the line with a large plug of wax. It is in fact possible to develop procedures to gradually remove the accumulated wax but this can be a prolonged and costly exercise.

Wax appearance points are normally determined by the Production Operations Group (POB) Sunbury (7). POB can also determine wax deposition rates. The extent and nature of wax deposits may dictate a need for regular pigging of a pipeline. It may also affect the choice of how production from a new field is linked into an existing system. For example, a requirement for regular pigging may rule out the use of a subsea tee for a smaller line feeding an existing larger one.

Listed below are wax appearance points for some typical stabilised crudes, and one live crude. The wax appearance point is not generally available for live crudes (although it can be obtained by POB) but it is always lower than that of a stabilised crude.

	Wax Appearance Point	
Crude	**Stabilised Crude (“C)**	**LIVE Crude (“C)**
Forties	32	24
Andrew	28	
Clair (low wax sample)	11	
Clair (high wax sample)	17	
Ivanhoe	36	

It should be noted that although some systems operate below the wax appearance point, their wax deposition rate may be so low that there is no requirement for regular pigging.

Further R&D effort is required to understand many aspects of wax deposition in multiphase systems. A major JIP is currently being set-up to build and collect data from a multiphase rig operating under conditions where wax deposition occurs. The ultimate aim of the JIP will be the development of models to predict wax formation and deposition characteristics in multiphase pipelines.

Wax crystal modifier chemicals are used in some systems to reduce was deposition and hence avoid pipeline operating problems. These chemicals do not prevent waxes forming, but alter the crystal structure to make it more difficult for the wax to adhere to the pipe wall. The wax particles are then more likely to be carried along with the liquid.

Problems relating to wax deposition should be taken up with POB Sunbury at an early stage of any project so that a testing programme can be planned and carried out in time to answer key questions relating to design and operability.

(b) Hydrate Formation

Hydrates are ice like materials which can form under certain conditions of temperature and pressure when water is present. Hydrates formation is promoted by high pressure and low temperature. The formation of hydrates is undesirable because they can obstruct the flow, causing an excessive pressure drop, and, if dislodged, can cause impact damage at bends or in vessels.

Hydrate formation conditions can be determined by the GENESIS program, or looked up in the GPSA charts.

The prevention of hydrate formation in multiphase systems can be accomplished by one of the following means :

- Maintain system pressure below the hydrate point.
- Maintain system temperature above the hydrate formation point.
- Remove all water from the hydrocarbons
  (this would normally require the gaseous phase to be dehydrated).
- Injection of conventional hydrate inhibitors.
- Injection of threshold hydrate inhibitor.

A brief summary of the advantages and disadvantages of the various hydrate inhibitors is given below.

(b1) Methanol

Methanol has the advantage over the glycols (MEG, DEG and TEG) of being able to dissolve formed hydrates. It also has a lower viscosity, and hence better dispersion characteristics than the glycols. Methanol is, however, more hazardous because of its higher vapour pressure and flash point.

If the gas stream from a methanol dosed multiphase stream is treated in a glycol contactor dehydration plant (normally using TEG) some of the methanol vapour may be absorbed in the contactor increasing reboiler duty and loading in the still column. The absorbed methanol vapour will then pass with the water vapour from the column and be vented to atmosphere.

(b2) Monoethylene Glycol (MEG)

MEG will not dissolve formed hydrates. If gas from an MEG dosed stream passes to a glycol contactor at the reception plant, there is a potential problem of MEG degradation. The degradation and boiling point temperatures of MEG are 165°C and 197°C respectively, compared to a typical TEG contactor reboiler operating temperature of ca. 204°C.

(b3) Diethylene Glycol (DEG)

As for MEG but vapour pressure is significantly lower (hence less passes to contactor). The degradation and boiling point temperatures of DEG are 164oC and 245oC respectively. The main disadvantages of DEG is its availability and cost. Also, DEG has a higher viscosity and inferior dispersion characteristics in comparison to MEG.

(b4)Triethylene Glycol (TEG)

There are no compatibility problems with the glycol contactor unit. The main problem with TEG is its high viscosity and hence poor dispersion characteristics.

(b5) Threshold Hydrate Inhibitor

The conventional alcohol and glycol inhibitors discussed above work by altering the thermodynamic phase equilibrium so that hydrate formation occurs at a reduced temperature. These chemicals are typically injected at rates of 1 0-40% by volume on the water phase. Recently a new type of inhibitor has been developed which is effective at very much lower dosage rates. These threshold hydrate inhibitors (THI) do not affect the phase equilibria but alter the kinetics of the hydrate formation process so that the possibility of hydrate formation occurring within the pipeline is reduced to a very low level.

Field and laboratory tests indicate that THI will be effective at concentrations of between 1000 and 5000 ppm. A THI field trial was conducted on the Ravenspurn to Cleeton gas/condensate line in May 1994. The trial showed conclusively that THI prevented hydrate formation for 6 days when it was used in place of methanol. Prior to injection of the THI a blank test was carried out during which neither THI nor methanol was injected. There was clear evidence that hydrates formed and partially blocked the line during this period.

(c) High Viscosity Fluids

As the temperature in a multiphase pipeline reduces so the viscosity of the oil phase increases and this leads to an increased frictional pressure gradient. The viscosity of stabilised and live oil can be determined by POB, Sunbury, as a function of temperature, and these values can be used to tune the viscosity correlations in MULTIFLO.

When oil/water emulsions are present determination of the liquid viscosity is far more complicated. The viscosity of an emulsion depends on the particle size distribution, which is largely determined by the flow history of the fluid and the shear imparted by chokes, pumps, etc. Dispersions that have undergone high shear (e.g. by flowing through a high pressure drop choke) will contain water particles of a relatively small size and this so called “tight” emulsion will have a relatively large viscosity.

Once formed the emulsion will generally exhibit non- Newtonian shear thinning flow behaviour so that its viscosity is dependent on the shear rate in the pipeline, i.e. the prevailing flowrate.

Emulsion viscosities are thus dependent on:

- Temperature
- Water cut

- Flow history, including shear rate experienced when formed
- Flowing shear rate, i.e. flowrate in the pipeline.

In MULTIFLO emulsion viscosities can be estimated using a weighted average of the oil and water. Alternatively the Richardson emulsion viscosity correlation can be used. This gives the user the ability to assess the sensitivity of the pressure drop predictions to various possible emulsion viscosities by altering the constant in the correlation.

When experimental data on emulsion viscosities become available, the constants in the Richardson correlation can be varied so that predicted emulsion viscosities approximately match the actual data for the range of flowrates (shear rates) that will be experienced in the pipeline.

The effect of viscosity on flow regimes and slug characteristics is discussed in Section 3.3.2 below.

3.2.4 Thermal Insulation of Pipelines and Risers

In some circumstances it is practical and cost effective to thermally insulate a pipeline system. Insulation may be provided for any one, or a combination of, the following reasons:

- Avoid or limit wax deposition.
- Avoid or limit hydrate formation.
- Minimise oil emulsion viscosity to avoid excessive pressure drop.
- Maintain fluids at a sufficiently high temperature to avoid, or limit requirement for, feed heating at the separation plant. A certain minimum separator temperature may be necessary to ensure satisfactory oil/water separation. A minimum temperature may also be required to meet the oil vapour pressure specification.

Pipelines and risers are always provided with some form of coating for corrosion protection and these offer varying degrees of thermal insulation. Risers on offshore platforms are normally coated with coal tar enamel or a very thin layer of fusion bonded epoxy (FBE). These offer little thermal insulation. Often risers are coated with ca. 12 mm of Neoprene to provide corrosion protection in the splash zone. Where thermal insulation is required this coating may be used over the full length of the riser.

Pipelines are normally protected against corrosion by FBE or coal tar enamel. For an offshore pipeline requiring concrete weight coating (to provide negative buoyancy) coal tar enamel is normally applied to the pipe to provide corrosion protection.

Offshore pipelines smaller than 16 ins. in diameter are generally required to be trenched to avoid being damaged by fishing trawl boards. Even if a pipeline trench is not mechanically backfilled (i.e. not buried) some natural in-fill of the trench normally occurs so that the pipe becomes partially buried. A survey of the Forties 32" MOL showed that most of the pipe had become totally covered in soil. In MULTIFLO the user can model partial burial by specifying the fraction of pipe surface exposed to the sea. The default value is 1 .O, i.e. fully exposed pipe.

Under some conditions a trenched pipeline has to be covered in rocks to provide stability or to protect it from upheaval buckling. This so called rock dumping buries the line providing enhanced thermal insulation. On some occasions pipelines are buried solely to provide enhanced thermal insulation. The FE-FA pipelines were buried for this reason.

In addition to the coatings referred to above, low thermal conductivity insulation materials can be applied to minimise heat loss from a pipeline. Insulation systems can be divided into 2 main categories:

(a) Low strength polymeric based materials requiring protection from the environment by encasing in an outer sleeve.

Examples of such system are:

- Central Cormorant multiphase production lines, which are insulated with polyurethane foam (PUF) encased in an outer steel sleeve. (Note that pipeline coaters are now offering a system for subsea use consisting of PVC foam sandwiched between an ethylene propylene diene monomer (EPDM) corrosion coating and an EPDM outer layer providing protection from the environment).
- Miller Landline. This is insulated with PUF encased in an outer layer of polyurethane.

(b) Insulation capable of withstanding the environment.

Examples are:

- Neoprene, as used on the FE-FA flowlines.
- EPDM, as used on the Don-Thistle flowline.
- Syntactic materials, such as syntactic PUF as used on Central Brae.

For further details on pipeline insulation materials available, costs and installation, please contact the Pipeline Group of XFE, BPX (Ian Parker).

Listed below are the thermal conductivity of the coatings and insulation materials previously referred to:

Material	**Thermal Conductivity** **Btu (hr °F ft)**	**W (mK)**
Concrete (saturated low density)	1.66	2.87
Concrete (saturated high density)	1.18(1)	2.04
FBE	0.114	0.2
Coal Tar Enamel	0.114	0.2
Neoprene/EPDM	0.156(2)	027
Polyurethane Foam (PUF)	0.017(3)	0.03
PVC Foam	0.023 - 0.04(3)	0.04 - 0.7
Syntactic PU	0.07(3)	0.12

Notes

(1) Most pipelines now use high density concrete.
(2) Ageing tests on Neoprene carried out in 1985 showed that the thermal conductivity rises to ca. 0.38 W/mK as water absorption occurs.
(3) Values depend on water depth, temperature and age.

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3.3 Flow Regimes

3.3.1 Discussion

Detailed descriptions of the various flow regimes occurring in horizontal and vertical flow were given in Section 1.2 of this manual.

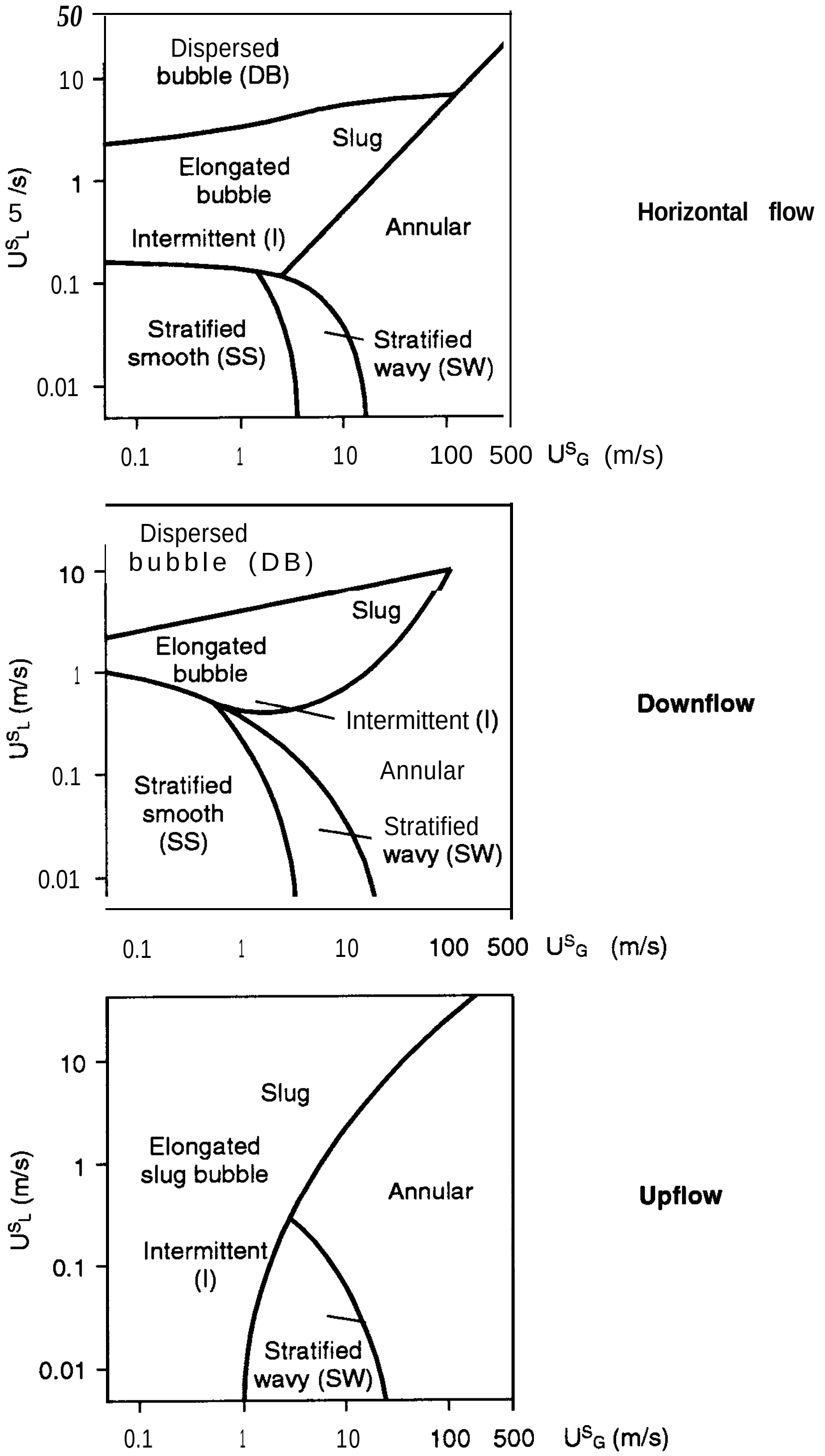

Figure 2. Taitel-Dukler Flowmaps

The transition between such flow regimes is usually presented in the form of a two dimensional flowmap. The usual co-ordinates of such a map are superficial liquid velocity (effectively liquid volume flowrate) and superficial gas velocity. The early flowmaps were derived from flow observations. These empirically based flowmaps are not particularly accurate for systems where the fluid properties, pipe size and inclination are different from those for which the flowmap was originally produced.

More recently flowmaps have been developed which incorporate physical models for each transition process. Models of this type, such as that of Taitel-Dukler (T-D), attempt to account for the effects of pipe diameter, pipe inclination and gas and liquid density and viscosity. In Figure 2 flowmaps are presented for a horizontal pipe, one with a small downhill inclination, and one with a small upward inclination. These flowmaps, which were produced using the T-D procedure, show how small inclinations significantly affect the position of the flow regime transitions. For a downward sloping line the area of the stratified flow regime increases. For an upwardly inclined line the stratified flow region shrinks and as the slope is further increased, disappears entirely.

One of the main reasons for producing a flowmap is to determine the conditions under which slug flow occurs. It can be seen in Figure 2 that the slug flow region occupies a fairly central position on the map, i.e. slug flow occurs at moderate gas and liquid flowrates. This type of slug flow is commonly encountered in pipelines carrying live crude oil such as the Magnus and Buchan satellite well flowlines, the Forties Echo to Alpha in-field flowlines, and the Wytch Farm in-field lines.

3.3.2 Effect of Viscosity on Flow Regimes

Experimental work has been performed at Sunbury to investigate the effect of viscosity on flow regimes and slug characteristics. The studies, performed on a 2” air - water/glycerol rig, showed that the slug flow region of the flowmap expands with increasing viscosity. This increase in the area of the slug flow region is in close agreement with the predictions of the Taitel-Dukler flowmap.

No data has been collected on the effect of viscosity on slug characteristics. However, the BP model correctly predicts a higher equilibrium liquid hold-up with increasing viscosity. As a higher hold-up in practice results in more frequent slugging, we can confidently predict that increasing liquid viscosity will give rise to smaller, more frequent slugs.

3.3.3 Recommended Flowmap

In the BPE Report “Summary of Field Data on Slugging Multiphase Systems” (8), it was commented that flow regime data collected by BP had shown that the original (T-D) flowmap was “unsafe” in that in many instances it predicted the flow to be annular when in fact slug flow was observed. The Barnea modified T-D flowmap predicts that the annular to slug transition occurs at higher gas velocities than the original T-D and so has a larger slug flow region. All the tests where slug flow was observed at Prudhoe Bay were predicted to be in slug flow by the Barnea modified T-D flowmap. On some occasions, however, annular flow was experienced when slug flow was predicted. The Barnea modified T-D flowmap can thus be thought of as providing a safe, or conservative, prediction of flow regime.

The Barnea modified T-D flowmap is now available in MULTIFLO and is the recommended method of evaluating flow regime in crude oil/gas multiphase pipelines.

3.3.4 Effect of Pipeline Geometry and Risers on Flow Regime

The type of slug flow indicated by flow regime maps arises as a result of fluid dynamic instability. It is frequently described as normal slug flow since it can occur in perfectly horizontal lines and does not depend for its generation on line topography or throughput changes.

Slugs may, however, be generated by other mechanisms. In particular when stratified flow occurs in a pipeline which undergoes changes in inclination, liquid may accumulate at low points to form a temporary blockage. Eventually the pressure builds up behind the blockage causing the liquid to be expelled and transported through the line as a slug. This type of flow is termed terrain, or geometry dependent slug flow.

when a flowline terminates in a riser a particular type of terrain slugging may occur where some, or all, of the riser fills with liquid. This phenomenon, known as severe slugging, imparts greater perturbations on the process plant, and larger forces on pipe bends, than does normal slug flow. Severe slugging occurs when stratified flow occurs in the flowline, and the ratio of gas to liquid flowrate is below some critical value such that the rate of pressure build-up upstream of the blockage at the riser base is insufficient to overcome the increase in hydrostatic head as the slug grows in the riser. (Severe slugging is discussed in detail in Section 3.4.2 and in Appendix 3C).

The conditions leading to the occurrence of severe slugging have been determined from experimental tests at Sunbury.

The Multiphase Flow Group at Sunbury have developed a mechanistic model which describes the conditions leading to the occurrence of severe slugging. This model has been validated with data obtained from the Sunbury experimental facility. The mathematical expression developed defines a critical liquid velocity above which severe slugging will occur:

$$V_{sl} > \frac{V_{sg}\, P_{sep}}{[L\,(\rho_l - \rho_g)\, g\,(1 - H_L)\sin\beta]}$$

where:

P_{sep} = separation pressure (N/m^2)
L = line length upstream of the riser(m)
ρ_l = liquid density (kg/m^3)
ρ_g = gas density (kg/m^3)
H_L = average liquid hold-up in line
g = acceleration due to gravity (9.81 m/s^2)
β = inclination of riser to horizontal (i.e. sin β is usually 1)

This expression, in a slightly different form, was later published by Shell following studies carried out on an experimental facility at the KSLA Laboratories. Shell defined a dimensionless number π_{ss} as:

$$\pi_{ss} = \frac{V_{sg}\, \rho}{[V_{sl}\, g\, L\, \rho_L\,(1 - H_L)]}$$

Hence when $\pi_{ss} > 1$, severe slugging is not experienced.

The expression developed by BP for severe slugging can be drawn on the T-D flowmap in MULTIFLO, and this then provides a flow regime map for the complete flowline-riser system. When the flow regime is predicted to be stratified-wavy, but lies to the right of the severe slugging line (i.e. at gas velocities in excess of that required for severe slugging), then wavy flow will occur in the flowline, and a churn or annular type flow will be experienced in the riser.

When the flow regime is anything other than stratified or stratified-wavy flow in the flowline, then that type of flow will persist through the riser. In particular if normal slug flow occurs in the flowline, then the slugs will continue on through the riser. Field data gathered from the Forties Echo-Alpha inter-field pipelines have shown that at moderate flowrates, slugs, once formed in the flowline, do persist essentially unaltered through the riser. However, at lower flowrates it was observed that although slug flow was maintained through the riser, the slugs underwent significant deceleration in the riser. This type of flow, termed slug deceleration, is discussed further in Section 3.4.1 below.

3.4 Slugging Flows

3.4.1 Normal or Hydrodynamic Slugging

It was discussed in section 3.3 that normal slug flow occurs at moderate gas and liquid flowrates and hence is commonly encountered in multiphase pipelines. The intermittent nature of this type of flow means that the separation plant at the downstream end of a slugging multi-phase line will experience variations in liquid and gas flow.

In order to be able to design reception facilities to accommodate slug flow, and to design pipe supports to handle the forces associated with slug flow, it is necessary to predict both of the following:

- **Slug volume (slug length and liquid content)**
- **Slug velocity**

(a) Slug Volume (Slug Length and Hold-up)

A considerable amount of R&D work has been carried out world-wide to improve the under-standing of slug flow. It was apparent from early experimental work that slug lengths observed in small scale laboratory rigs could not be scaled up to field conditions.

In the late 1970s the operators of the Prudhoe Bay Field, Alaska, (PBU) were faced with the problem of designing numerous large diameter multiphase flowlines (24" in diameter) and the separators into which the fluids pass. They commissioned fluid flow specialists at Tulsa University (Dr. J.P. Brill) to carry out a large programme of work which included the collection of slug flow data from existing lines at PBU. The largest of these was a 16" 3-mile pipeline. From the data collected at PBU, together with some information on smaller flowlines, and laboratory scale multiphase systems, a correlation was developed to predict mean and maximum slug length. This correlation was published by Brill et al in 1979 :

$$\ln(L_m) = -2.663 + 5.441\,[\ln(d)]^{0.5} + 0.059\,[\ln(V_m)]$$

where:

L_m = mean slug length (ft)
d = pipe diameter (in)
V_m = mixture velocity (ft/sec)

The PBU data suggested that slug lengths follow a log normal distribution, i.e. slugs distributed about the mean length through the following relationship:

$$\ln(L_s) = \sigma^* zp + \ln(L_m)$$

where:

L_s = slug length observation

σ = standard deviation (approximately 0.5)
zp = standard normal distribution function (3.08 for 0.1% probability)

Based on the 16" PBU data, Brill et al claimed that the longest slug likely to be observed equates approximately to the 0.1% probability slug. The calculated 0.1% probability slug is approximately 4.7 times longer than the calculated mean slug.

The above slug length correlation has become the industry standard design method and has been used extensively by oil companies and contractors since 1979.

In 1980/81 the Tulsa University workers returned to PBU to gather data on a new 24" line that ran parallel to the 16" line that had been monitored previously. They found that in general slugs were not as large as had been predicted using their original correlation. BP revised the original slug length correlation to take account of the new 24" data. It is this modified correlation which is available in MULTIFLO. It is referred to as the Brill method as the correlation was kept in the same form as Brill's original equation:

$$\ln(L_m) = -3.579 + 7.075\,[\ln(d)]^{0.5} + 0.059\,[\ln(V_m)] - 0.7712\,[\ln(d)]$$

The above correlation is heavily biased by the PBU data. Most of this data was collected on a 16" and 24" line which run parallel to one another and hence have the same length and geometry. The correlation thus shows no dependence on line geometry or fluid type, and only a very weak dependence on flowrate. It was widely felt that the correlation could prove unreliable when applied to other types of systems. Consequently, under BPX, XTC, sponsorship the Multiphase Flow Group commenced a programme of work aimed at gathering slug flow data on a wide range of multiphase systems.

Since 1985 the Multiphase Flow Group have collected data from the following sites.

Wytch Farm	(Sherwood)	6 and 10"
Wytch Farm	(Bridport)	4"
Magnus Satellite Well Flowlines		6"
Forties Echo-Alpha in-field lines		6" and 12"
Prudhoe Bay WOA	(1987)	24"
Prudhoe Bay WOA and EOA	(1989)	24"
Kuparuk	(1990)	10-24"
Don-Thistle		8"

A summary of the data collected on all these tests, apart from the last 2, is available in (8).

The data collected by BP covers a wide range of pipe sizes, lengths, geometries, fluid properties, and water cuts. The data has been compiled into a database vastly greater than anything previously available. Using this database, the Multiphase Flow Group have developed a semi-mechanistic correlation for slug frequency. By applying one of the available slug flow models, knowledge of slug frequency yields the mean slug length. This is discussed in Appendix 3A.

No data has yet been collected from slugging hilly terrain pipelines. A new slug model has recently been developed by the Multiphase Flow Group to track slug sizes throughout a hilly terrain pipeline system. This model was referred to in Section 3.1.3, and is discussed further in Appendix 3D. It is intended to gather data from the Cusiana in-field flowline system. The data collected from this hilly terrain system will be used to validate and develop the new model.

In parallel with this data gathering exercise, the Multiphase Flow Group have carried out an extensive programme of work to determine how slug lengths are distributed about the mean. This work has involved analysis of data collected on experimental rigs at Sunbury as well as data collected from other experimental facilities. Using the limited PBU data Brill et al had concluded that slug lengths follow a log-normal distribution. Hence the Brill correlation always calculates the maximum slug length to be 4.7 times the mean. The Multiphase Group have found that the distribution of slug lengths actually varies with the location of the pipeline operating point on the flowmap. As a consequence the ratio of maximum to mean slug length varies with flowing conditions. Near the stratified wavy boundary the maximum slug length may be ca. 4-5 times the mean. However, near the elongated bubble transition maximum slug lengths are only ca. 2 times the mean. Work on quantifying slug length distribution is continuing at Sunbury.

The combination of a method for determining slug frequency, a slug flow model, and a slug length distribution model provides a means of predicting mean and maximum slug lengths. This new approach is available in MULTIFLO as the "RCS Mechanistic Model".

A more detailed explanation of the slug model is given in Appendix 3A.

Prediction of the liquid volume associated with slug flow requires knowledge of the liquid content of the slugs (slug hold-up), as well as the slug length. The slug hold-up method currently used within the RCS slugging model is that due to Gregory. The Gregory model is based on laboratory scale experiments and predicts slug hold-up as a function of mixture velocity only. BP has found that in practice slug hold-up is strongly dependent on water cut as well as mixture velocity. Hence at Prudhoe Bay slug hold-ups have increased from ca. 0.3 to 0.9 as water cuts have risen from 0 to 50%. Similar significant increases in slug hold-up have been observed on the FE-FA 6" test line when running wells of different water cuts. Further work is being conducted in this area in order to develop a more reliable design method.

(b) Slug Velocity and Forces due to Slugging

(b1) Horizontal Lines

When slug flow occurs in an essentially horizontal line the mean velocity of the liquid in the body of the slug is equal to the mixture velocity, V,, hence :

$$V_s = V_m = V_{sg} + V_{sl}$$

V_s = **mean slug velocity**

It is this velocity, V,, which should be used when evaluating the forces imposed by a slug as it travels through a bend.

A detailed discussion of how to calculate loads associated with slug flow is given in Section 11 of this manual.

(b2) Pipeline Risers

It was mentioned in Section 3.3 that although a slug will progress through a riser at the downstream end of a pipeline, it will tend to decelerate as the line upstream of the slug packs up to provide the pressure to overcome the increasing hydrostatic head in the riser.

As the slug leaves the riser the hydrostatic head loss in the riser reduces so that the upstream gas bubble expands and accelerates the slug into the process plant. This phenomenon is shown schematically in Figure 3.

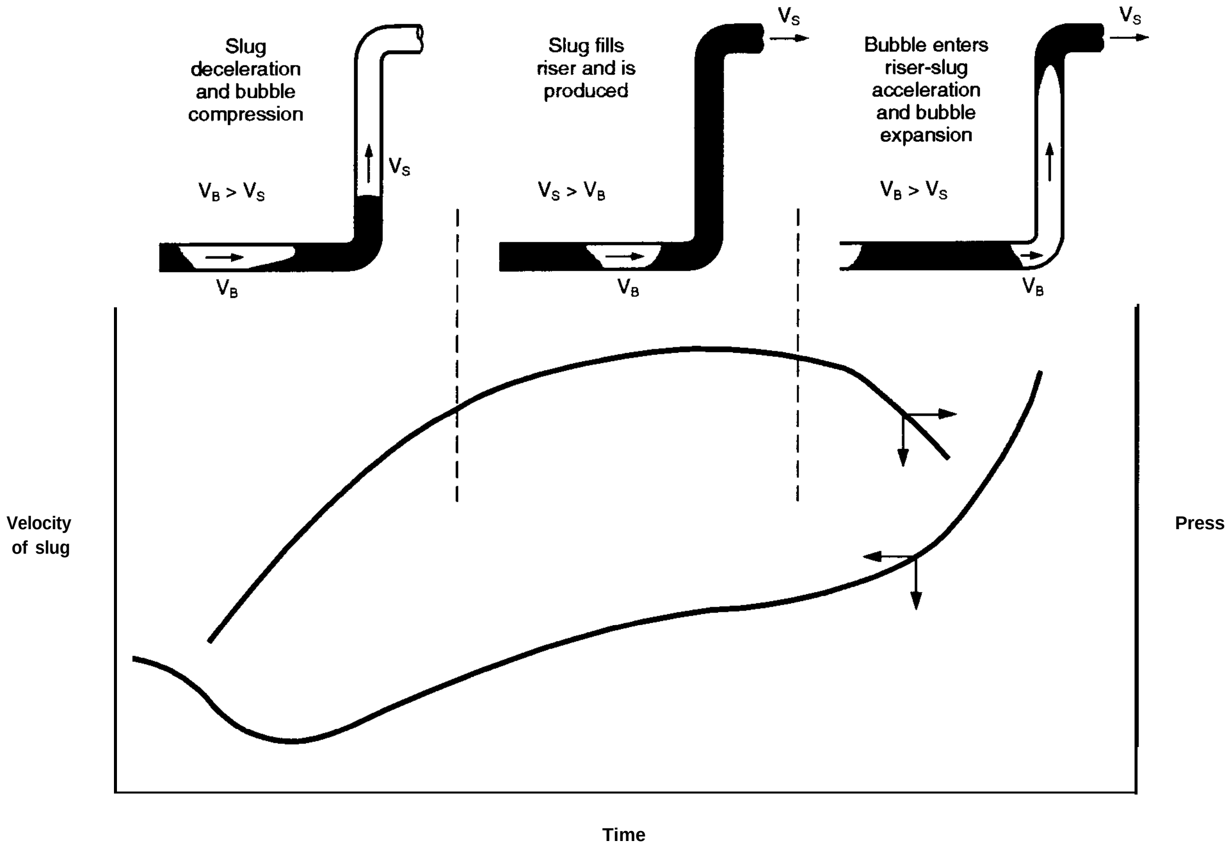

Figure 3. Normal slugging in flowline riser

In order to model the effect of slug flow in a pipeline- riser, BPX has developed a dynamic program known as NORMSL. This can be used to determine the velocity of the slugs as they pass through the riser and into the topsides pipework. In order to assess the effect of slugging on the topsides process plant the output file from NORMSL can be used as the input to a dynamic simulation model of the topsides plant.

Process dynamic simulation work in BPX is often performed using the SPEED-UP program. (This type of work is performed by the Process Simulation Group of ESS, Sunbury). NORMSL and SPEED-UP have now been directly linked so that pressure changes in the process plant arising from the production of slugs or gas bubbles are fed back directly into the pipeline-riser model. This feedback effect becomes more significant as the height of the riser increases, i.e. for deep water developments.

3.4.2 Terrain/Geometry Dependent Slugging

(a) Hilly Terrain Pipelines

It was explained in Section 3.3 that when stratified flow occurs in a pipeline liquid may accumulate at low points to form a temporary blockage. Gas pressure builds up behind the blockage causing the liquid to be expelled as a slug. Such terrain induced slugging is clearly dependent on the geometry of the pipeline as well as the flowing conditions.

As the available design methods for this phenomenon were inadequate and known to significantly overpredict slug size, the Multiphase Flow Group carried out physical and theoretical modelling studies to gain a better understanding of the terrain slugging process. These studies have led to the development of a mechanistic model for slug production from dips. The model has been validated with data obtained from the experimental facilities. The new model is discussed in detail in Appendix 3B.

The two most important aspects of the model with regard to pipeline design are:

(al) Critical Gas Velocity

The critical gas velocity above which no liquid accumulates in a dip is evaluated by considering the gas velocity for total liquid removal by the co-current liquid film process.

(a2) Slug Size and Frequency

At gas velocities below the critical value liquid can accumulate in a dip and eventually this leads to the production of a slug. As a slug moves through the uphill section, liquid is shed from its rear and runs back down the slope. If insufficient liquid is available in the preceding film for the slug to scoop up and replace that lost by shedding, then the slug will collapse before reaching the brow of the hill. For a system with a steady liquid inflow, liquid will build up in the dip so that eventually slugs will emerge to pass into the downstream pipework. The model evaluates the frequency of slug production for a particular geometry. With knowledge of the slug frequency, the slug size is determined by calculating the volume of liquid entering the system during the inter-slugging period.

(b) Pipeline-Riser Systems

(b1) Theory

When a pipeline terminates in a riser a particular type of terrain slugging may occur which has been variously termed riser and severe slugging. The conditions giving rise to the occurrence of severe slugging have been outlined in Section 3.3.

A diagrammatic representation of severe slugging is given in Figure 4. A detailed description of the severe slugging phenomenon plus details of the BP test rig studies is given in Appendix 3C. A brief description of severe slugging is given below:

When a stratified flow occurs in a pipeline and the ratio of gas to liquid flowrate is below some critical value, a liquid blockage will form at the base of the riser. When such a blockage occurs liquid accumulates at the base of the riser while gas is trapped within the flowline. The liquid slug now formed at the junction between the flowline and the riser will continue to grow if the rate of hydrostatic head increase in the riser, corresponding to the rate at which liquid arrives from the flowline, is greater than the rate of gas pressure increase in the flowline. Liquid accumulation continues until the riser is full of liquid (slug generation).

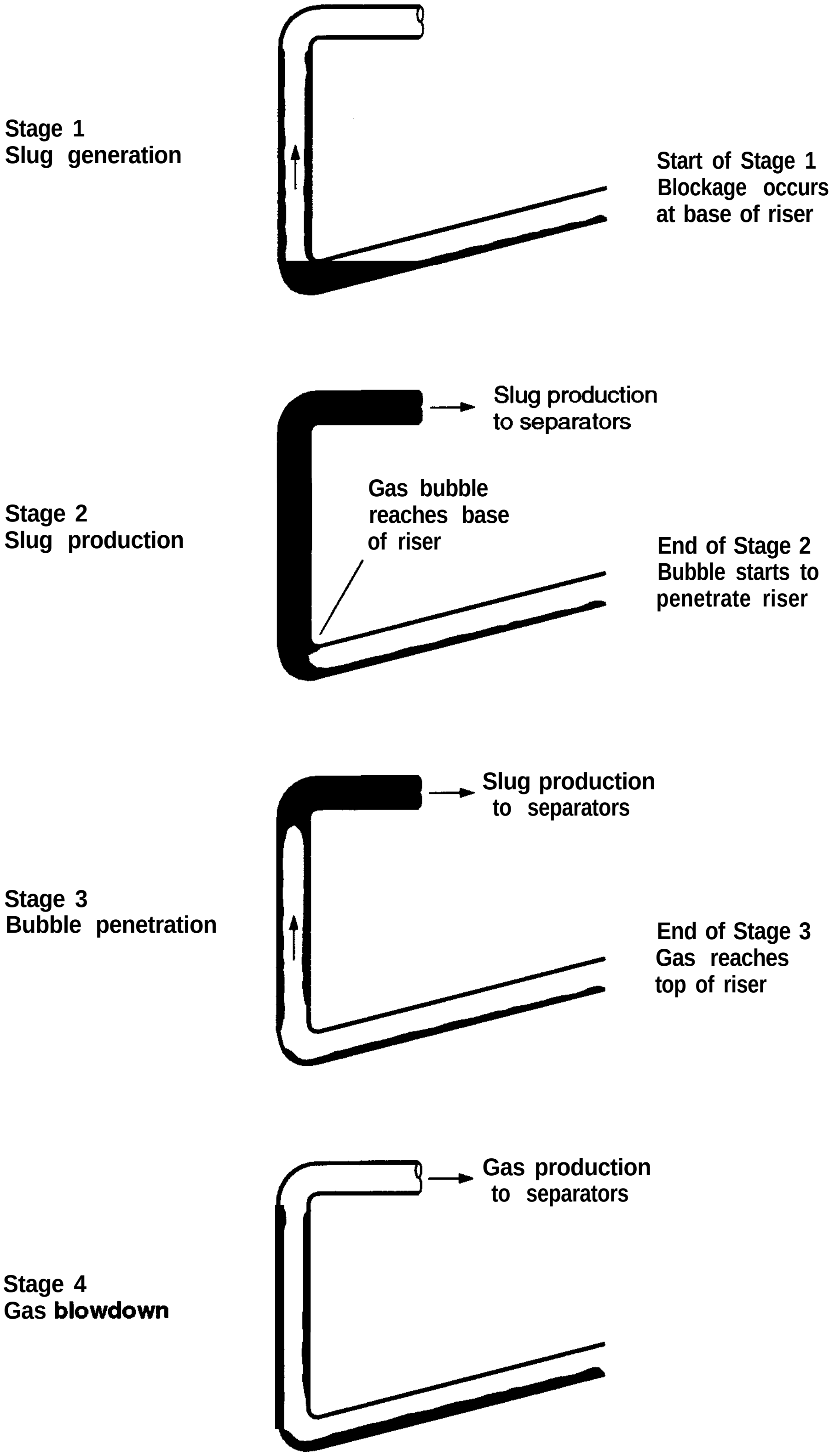

Figure 4. Severe slugging in flowline riser

At this point the hydrostatic head loss over the riser reaches a maximum value and the slug begins to be pushed slowly from the flowline (slug production).

Once the gas slug interface enters the riser the hydrostatic head decreases rapidly while the expanding gas bubble accelerates the bulk of the liquid from the system (bubble penetration).

This stage continues until the gas bubble enters the separator. Gas production rapidly rises to a maximum value and then declines steadily as the line depressures (gas blowdown).

As the gas flowrate reduces, any liquid held up in the riser falls back and accumulates together with liquid arriving from the flowline, to form a blockage at the base of the riser. The cycle is then repeated.

The occurrence of severe slugging is generally associated with flowlines which slope downwards to the base of the riser. However, a similar surging cycle can be produced in horizontal and slightly upwardly inclined pipelines.

(b2) Modelling

Extensive physical modelling has been carried out by the Multiphase Flow Group to investigate and confirm previous descriptions of severe slugging and to investigate means of eliminating the phenomenon, see Appendix 3C. Much of this work was sponsored by the SE Forties project.

Of particular interest to the SE Forties Project was the use of riser gas injection to eliminate severe slugging both at low throughputs and at start-up. Tests showed that gas injection would reduce the severity of the severe slugging cycle and that in sufficient quantity it would completely eliminate the phenomenon.

A computer model of severe slugging was purchased from the Tulsa University Fluid flow Projects (TUFFP). This program evaluates liquid and gas production rates throughout the severe slugging cycle and so by feeding the results directly into a dynamic model of the topsides plant, the effects of slugging on the process facilities can be evaluated. The TUFFP model was modified at Sunbury to include topsides pipework. The program could then be used to evaluate the forces exerted on the riser and topsides pipework. The program was further modified to include the effect of riser gas injection. The predicted quantities of gas required to eliminate severe slugging were found to be in good agreement with the data collected from the BPX Sunbury experimental facility.

It was found that the Taitel-Dukler-Barnea criterion for the onset of annular flow provided a good estimate of the quantity of gas required to eliminate severe slugging. This method predicts that the transition to annular flow occurs at superficial gas velocities in excess of a critical value given by:

$$V_{sg} \geqslant \frac{3.1\ [s\, g\, (\rho_l - \rho_g)]^{1/4}}{\rho_g^{0.5}}$$

where:

s = surface tension (N/m)
g = 9.81 (m/sec^2)

ρ_l = liquid density (kg/m^3)
ρ_g = gas density (kg/m^3)

The modelling work showed that partial closure of a choke positioned near the top of the riser would also eliminate severe slugging.

The choke adds sufficient frictional pressure drop so that the system pressure loss becomes dominated by friction, rather than by hydrostatic head loss as is the case for the unchoked pipeline-riser. In order to establish a stable, friction dominated system, the choke needs to provide a pressure drop comparable to the hydrostatic head loss over the riser when full of liquid.

(b3) Operational Experience

(b3.1) General

The SE Forties field is the principal BP-operated site where severe slugging was considered in the design stage. A number of the Magnus satellite wells may also have exhibited severe slugging had their maximum production rates been reduced to low values (their arrangement is different to SE Forties in that the chokes are at the top of the riser - severe slugging would not occur unless the riser- top chokes were almost fully open).

There is one 12” line and one 6” line between Forties Echo minimum facilities platform and Forties Alpha. There are eleven wells on Forties Echo, now mainly assisted by electric submersible pump. They have chokes on Forties Echo. Therefore once the flows from the wells are co-mingled into the 12” line (or flowing individually or co-mingled through the 6” line), there is the possibility of severe slugging at low flowrate.

Low flowrate considerations obviously may also include start-up, when one or more severe slugging cycles may be experienced before a higher steady-state flowrate is reached.

In April 1988 a major field test was undertaken on the Forties Alpha platform mainly to investigate the behaviour of the 6” line over a range of flowrates, and from different wells that had a range of water cut. The aim was to keep reducing the flowrate from one well at a time through the 6” line until the onset of severe slugging. However, due to a fear that choking back the wells too far might kill them, it was not possible to reduce the production rates sufficiently to reach genuine severe slugging during the tests. Despite this limitation, there was a marked change in the character of the flow pattern, from steady low liquid hold-up slugging to a regular surging, as the flowrates were reduced.

Experience with the 12” line is limited to start-ups. Operators present at first oil through the line reported heavy surging. During the 1988 tests three start-ups were monitored, with several large surges during each. However, because of the steadily increasing input flowrates it was not possible to assess the accuracy of the predictive tools, either for the occurrence of severe slugging, or for the cycle time.

(b3.2) Elimination of Severe Slugging

The Forties Echo-Forties Alpha line has two methods for ameliorating the effects of severe slugging - riser-base gas injection, and riser-top choking. The theory behind these 2 methods is discussed in Appendix 3C.

The design intention for SE Forties was that riser gas injection would be the principal means of eliminating or ameliorating severe slugging at low production rates and at start-up.

Gas is re-cycled after NGL extraction through a 3” line down to the base of the 12” riser, and a 2” line for the 6” riser. The gas injection system capacity is sufficient to put the risers into annular-mist flow at the highest gas rates.

As mentioned above it was not possible to undertake any ‘steady-state’ severe slugging tests during the 1988 field work. However, the use of gas injection to ameliorate the low flowrate period of a line start-up was demonstrated on both diameter lines.

The full gas injection procedure was used during the third of the 1988 test sequence of three start-ups of the 12” line, with the maximum recommended gas injection rate (180,000 Sm3/day) into the 12” line flowing before the first production fluids were introduced at the Echo platform. Whilst there was still an initial surging the gas injection did serve to reduce the duration of the main surge, and also the number of cycles.

This benefit of gas injection during start-up was also demonstrated whilst trying to bring on a weak well through the 6” test line. The well began to flow, but as liquid reached the base of the riser at Forties Alpha the additional back pressure due to the increasing liquid head was sufficient to cause the well to stop flowing. To overcome this the gas injection system was switched on in the 6” line (up to 60,000 Sm^3/day). The liquid head build-up was greatly reduced, thereby minimising any back pressure and allowing the well to build up to its maximum production rate.

The Forties Echo risers at FA were also fitted with throttling valves, as a back-up to the gas injection system. Use of the throttling valves was seen as having the disadvantage of requiring a large pressure drop to be effective. However, installation of ESPs on Forties Echo has meant that pressure is available to overcome the throttling valve pressure loss, and this combined with the fact that gas is often not available when FE is started up has meant that use of the throttling valve on the 12” riser has become the standard means of avoiding severe slugging at start-up. As water cut has risen, surging at start-up has become more pronounced such that even with the throttling valve shut-in to its minimum stop (25% open) the system was regularly tripping out on high separator level at start-up.

Forties have recently reduced the minimum stop position to 10% and this has enabled operations to bring FE on line with minimal surging and without high level separator trips. Data collected during the November 1993 field trials illustrated the relatively steady nature of a start-up of the 12” line when shutting in the throttling valve to its minimum stop condition (9).

3.4.3 Slugging Produced by Transient Effects

In addition to the mechanisms already discussed, slugs may be produced as a result of transient effects such as pressure or flowrate changes. For example, if a line operating in stratified flow is subject to an increase in gas flowrate, or total production rate, one or more slugs may be produced as the equilibrium liquid level drops towards a new steady state condition.

BP has access to two general purpose transient multiphase programs. These are PLAC, developed at Harwell as part of a Joint Industry Project, and TRANFLO, developed at Sunbury and now available through Harwell. OLGA is another commercially available transient simulator. Current thoughts on the relative merits and applicability of these codes, together with a detailed discussion of transient multiphase modelling is given in Section 7 of this manual.

3.4.4 The Effect of Slugging Flows on Process Plant

(a) Control Philosophies

It has already been mentioned that the intermittent nature of slug flow results in the separation plant experiencing variation in liquid and gas flow. A separator, or slugcatcher subject to slugging conditions will thus experience variations in:

- Liquid level
- Gas pressure
- Liquid and gas outflow.

The exact nature of these variations will depend on the way in which separator/slugcatcher controls are set up.

In some cases the first stage separator is set up on level control. If the control system is set to respond rapidly to changes in level, the liquid level may rise only slightly, but the liquid outflow will increase sharply. Under these circumstances the liquid disposal system (valves and pumps), and the downstream liquid handling plant will have to deal with rapid variations in liquid flowrate. The problem with this type of system is that the disposal pumps and downstream liquid processing plant need to be sized for flowrates well in excess of nominal maximum production rate. In addition, the separator liquid residence time is reduced during periods of slug production so that oil/water separation efficiency may deteriorate.

The usual way in which slugs are catered for is through use of a large separator, or slugcatcher, where the increase in liquid flowrate during slug production is handled by allowing the liquid level to rise. Provided the liquid level remains within set limits the liquid disposal rate may be kept constant. The vessel is then said to be on flow control. Once the liquid level rises above a certain point the liquid flow controller has to be reset to a higher value to avoid high level shutdowns. The advantage of this type of system is that the flowrate of liquid to the downstream plant is maintained near constant. The disadvantage is the large size of vessel required to accommodate the slugs.

The two types of control philosophy discussed, level control and flow control, represent the two extreme methods of operation. In practice some blend of the two may provide the solution for any particular application. In order to optimise the liquid level and disposal control systems a dynamic model of the plant should be developed during design. Dynamic modelling is further discussed below.

In addition to handling the variations in liquid flowrate during slugging, the reception facilities will have to be designed to cater for the large variation in gas flowrate during slugging.

In fact generally, it is the gas plant which is more susceptible to operational upsets during slugging than the liquid plant. As with control of the liquid level, there are two philosophies for handling gas production from the separator or slugcatcher.

(al) Fixed pressure control

In this system the excess gas flowrate occurring during production of the gas bubble passes rapidly through the separator to the downstream gas plant.

(a2) Allow Pressure in Separator to Rise

Here at least some of the excess gas production is held back in the separator as the pressure is allowed to rise. This type of system will impart a smaller perturbation on the downstream gas plant than one with fixed pressure control.

In this case the separator is acting as a gas accumulator. Clearly this type of system requires the normal operating pressure of the separator to be set at some level below the design pressure.

(b) Dynamic Simulation of Process Plant

In order to accurately assess the effect of slugging on the process plant it is necessary to take the output from a slugging model and input this into a dynamic model of the process plant. BPX have used the dynamic simulation package SPEED-UP on a number of recent developments studies including Bruce WAD, Cyrus-Andrew, and Cusiana. This program is used and developed by the Process Simulation Group of ESS Sunbury. SPEED-UP can be used to model a wide range of items or process plant including pumps, compressors, valves, and all their associated control systems.

A slugging model, such as NORMSL, see Section 3.4.1, produces output in terms of gas and liquid production rates with time. In order to accurately assess the interaction between the process plant and the slugging pipeline, NORMSL has been interfaced with SPEED-UP. In this way any pressure changes in the separator or slugcatcher, caused by slugging, are fed back directly into the pipeline model where they may affect the production rate of subsequent slugs and bubbles.

The dynamic model can be used to assess the ability of a preliminary design to handle slugging conditions. If the model suggests that slugging flow will create unacceptable plant conditions it may be possible to produce an operable system by modifying or changing the control system. On other occasions new or larger vessels may be required.

The Process Simulation Group use the combined NORMSUSPEED-UP model to assess the relative merits of various active slug control schemes, such as the use of separator inlet throttling. They are also conducting field trials to validate the findings of such modelling activities.

The PLOT team of BPX Aberdeen use a dynamic simulation package to assist in optimising platform topsides plant conditions and assessing the effect on the process plant of changes in feed rate and operating conditions. PLOT have recently incorporated the BPX slugging model NORMSL into their dynamic model of the Bruce topsides to enable them to assess the effect of slugging from the Bruce Western area on the Bruce process plant.

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3A Slug Flow Design Method

3A.1 Introduction

As described in the main body of the design guidelines (Section 3.4.1 (a)) the pioneering field work on steady-state slug flow was carried out by Professor Jim Brill (Tulsa University) at Prudhoe Bay in 1978 and 1980 (1). The resulting correlation for slug length was derived using statistical regression techniques on gas and liquid flowrate, and diameter.

In the mid 1980's BP began their own programme of field data collection on lines in slug flow. The 'Field Data' section of the bibliography of BP Research reports on multiphase flow lists the relevant titles. It quickly became evident that the original Brill equation (and the modification due to Norris) was very conservative in predicting both mean and maximum slug lengths over a variety of flowrates and diameters. This provided the incentive to develop an improved method that sought to take into account a greater number of the relevant variables than the original Brill method, and to do so using a mechanistically based approach.

BP Exploration have been involved in the development and implementation of this method, which was installed into the MULTIFLO computer programme early in 1990, thereby making it generally available within the BP group. In addition, a number of papers, covering test rig and field data collection and the development of part of the slug flow method, have been published by BP in the open literature.

3A.2 Key Variables in the Prediction of Slug Characteristics

As stated in Section 3.4.1 (a) the key parameters required for the assessment of the performance of separation and associated downstream facilities are the average slug volume and frequency, and the greatest likely slug volume and its frequency of occurrence. To assess the design of pipework supports the combination of slug velocity, frequency and liquid holdup is important.

These parameters are all influenced, to a greater or lesser extent, by the following variables:

V_{sg} = gas superficial velocity
V_{sl} = liquid superficial velocity
D = pipeline diameter
L = pipeline length
θ = pipeline inclination
ρ_g = gas density
ρ_l = liquid density
μ_g = gas viscosity
μ_l = liquid viscosity
σ_l = surface tension

The aim of the work described here has been to develop a mechanistically-base method for slug characterisation that reflects the contributions of many of these different variables. This gives much greater confidence in the method output over that of a straight correlation based on statistical regression of limited using only one or two variables.

3A.3 Development of the Average Slug Frequency Method

The physical process causing normal slug flow in (near) horizontal pipe flow is the formation and growth of a wave on the surface of a stratified liquid film. This wave results from a flow of gas over the liquid film. The wave may grow until it reaches the top of the pipe at which point the gas flow is sealed behind the incipient slug. As the gas velocity was much greater than the wave, there is then an acceleration of the wave/incipient slug up to a velocity approaching that of the bulk gas. This increased wave/slug velocity results in liquid in the film ahead of the wave/slug being picked up and incorporated into the slug body. Liquid is shed off the back of the newly formed slug, but at a slower rate than the pickup rate. The slug keeps growing in this way until it reaches the end of the liquid film shed by the slug in front (with the film shed from a slug being less deep than the equilibrium stratified liquid film). A stable length is then maintained.

The fundamental parameters determining the average slug length are therefore the slug formation frequency, the depth of the equilibrium stratified liquid film ahead of the slug, the behaviour of the liquid film shed from the rear of the slug, and the liquid holdup in the slug body.

Of these four parameters the one with the least reliable prediction method, before the BP R&D programme, was the slug frequency. Correlations based on small diameter and/or a limited range of fluid properties have been available for some time, but none performs particularly well against the field data collected by BP since 1985.

What has now been developed is a relationship between a dimensionless slug frequency and the equilibrium stratified liquid holdup (H_{le}) that would result from the prevailing gas and liquid flowrates and physical properties if slugs were not being produced (2). The exact nature of the relationship has not yet been derived mechanistically, and so a correlating line is used whilst further work on the mathematics is being undertaken. However, this is a big step further on than the simple regression techniques, which did not aim to describe the physics behind the phenomena.

The dimensionless frequency has the form:

$$\frac{F_s D}{V_m}$$

where:

F_s = the actual slug frequency
V_g = in-situ gas velocity
V_l = in-situ liquid velocity

H_{le}, V_g and V_l are calculated from the equations (due to Taitel and Dukler (3)) for determining the depth of the equilibrium stratified liquid film given the gas and liquid flowrates, pipeline geometry, and fluid properties.

Gas phase

$$A_g \left[\frac{dP}{dx}\right]_g = -\tau_{wg} S_g - \tau_i S_i - A_g \rho_g g \sin\theta$$

Liquid phase

$$A_l \left[\frac{dP}{dx} \right]_l = -\tau_{wl} S_l + \tau_i S_i - A_l \rho_l g \sin\theta$$

where:

$A_l = A H_{le}$
$A_g = A (1 - H_{le})$
S_g = gas wall contact length
S_l = liquid wall contact length
τ_{wg} = gas shear stress to the pipe wall
τ_{wl} = liquid shear stress to the pipe wall
τ_i = inter-facial shear stress between gas and liquid

At equilibrium

$$\left[\frac{dP}{dx} \right]_g = \left[\frac{dP}{dx} \right]_l$$

The two equations may then be solved for H_{le}.

The form of the slug frequency method that correlates the dimensionless frequency against the equilibrium stratified liquid holdup, is given below. This method was published in (7):

$$\left[\frac{F_s D}{V_m} \right]' = -24.729 + 0.00766\, e^{(9.91209 \cdot Hle')} + 24.721\, e^{(0.20524 \cdot Hle')}$$

where:

$$\left[\frac{F_s D}{V_m} \right]' = \frac{F_s D}{V_m} (1 - 0.05\, V_{sg}) \,.\, D^{0.3}$$

$$H_{le}' = H_{le} \left[1 - \frac{0.068}{V_{sl}} \right]$$

D in m

V_m, V_{sg} and V_{sl} in m/s

This relationship covers both published and in-house data from test rigs and producing oilfields. The diameter range is from 1.5 to 24 inches, with fluids such as air and water, nitrogen and diesel, and actual produced reservoir fluids involved in the studies.

The inter-facial friction factor, f_i, used in the development of the correlation, was that due to Andritsos and Hanratty i.e.

For $V_{sg} \leq 1.5$ m/s

$$f_i = f_g$$

For $V_{sg} \geq 1.5$ m/s

$$f_i = f_g [1 + (V_{sg}/1.5 - 1)\ 0.75]$$

Further work is underway to revise the slug frequency method by the inclusion of additional test rig and field data, and by reviewing the selection of the inter-facial friction method.

3A.4 Slug Holdup Method

Currently there is one widely recognised method for the prediction of liquid holdup in slugs, due to Gregory (4). This is presented as a function of mixture velocity, shown below:

$$H_{ls} = \left[\frac{1}{1 + (V_m / 28.4)^{1.39}} \right] \qquad \text{(with } V_m \text{ in ft/s)}$$

This correlation gives reasonable performance with two-phase mixtures (e.g. air and water, or dry crude and associated gas). However, when there is water in the crude the slug liquid holdup tends to be much higher than predicted.

This does not affect the calculations of average slug volume (as the slug model is a mass balance determined by the slug frequency, which is calculated independently of slug holdup). However, calculations of slug length must always be considered in conjunction with the predicted slug liquid holdup.

The slug holdup becomes an important parameter when considering pipework loads on bends and supports (see relevant section of the Design Guidelines).

3A.5 Use of the Creare Slug Model

Models of slug flow are essentially mass balances of gas and liquid flow through the slug/bubble unit. The main difficulty is encountered in predicting the behaviour of liquid being shed from the back of the slug.

The principal workers in this area have been Dukler and Hubbard (5), who wrote a model for slug flow in 1975. The model requires slug frequency and slug holdup as inputs. The model was revised by Crowley of Creare inc. during 1986 to improve the solution algorithms (6).

The solution process involves first calculating the length of the gas bubble/liquid film section. Three variables are involved, and they are interdependent (therefore requiring solution of two simultaneous equations) - the gas bubble length (L_b), the minimum film velocity (V_{cmin}) and the minimum film holdup (H_{fmin}). The gas bubble length is calculated as a function of the minimum film velocity. The detailed equations for all inclinations are available from BP Research. For horizontal flow the solution sequence is given below:

V_{sf} = velocity of slug front

L_s = mean length of slugs

Slug frequency (F_s) and holdup (H_{ls}) are obtained from the methods described in the preceding sections.

$$V_{sf} = 1.3\, V_m \qquad \text{(with } V_m \text{ in m/s)}$$

$$L_b = \frac{e}{a V_m}\left[1 - \frac{V_m}{V_{cmin}}\right]$$

where:

$$e = -\rho_l \left[\frac{\pi D^2}{4}\right](V_{sf} - V_m)$$

$$a = \left[0.046 \left(\frac{\rho_l D V_m}{\mu_l}\right)^{-0.2} \left(\frac{\pi D \rho_l}{4}\right)\right]$$

$$V_{cmin} = \left[V_{sf} - \frac{H_{ls}(V_{sf} - V_m)}{H_{fmin}}\right]$$

From the above equations L_b can be expressed as a function of H_{fmin}. The second independent equation also expresses the gas bubble length as a function of the minimum liquid film fraction:

$$L_b = \left(\frac{V_{sf}}{F_s}\right) - \left[\frac{V_{sf}}{F_s\, 1.3\,(H_{ls} - H_{fmin})}\right]\left[\left(\frac{V_{sl}}{V_m}\right) - H_{fmin} + 0.3\,(H_{ls} - H_{fmin})\right]$$

The two equations are then solved for L_b.

Finally the mean slug length is obtained from:

$$L_s = \left[\frac{V_s}{F_s} - L_b\right]$$

This is the output of slug length in MULTIFLO.

3A.6 Development of the Slug Length Distribution Model

Since the work by Brill there has been an assumption that the distribution of slug lengths usually follows a log-normal distribution. The large amount of test rig and field data collected by BP indicates that this is not valid as a general assumption. The log-normal case tends to be the extreme form of distribution applicable – extreme in the sense that the ratio of 0.001 probability slug length to mean slug length is higher than other distributions that have since been considered for slug flow. The implication of this is the possibility of over-sizing slugcatcher/separator vessels.

Other distributions that are applicable under varying circumstances are normal, lambda and inverse Gaussian. Each distribution method requires the determination or assumption of one or more parameters, in addition to the predicted mean value.

Currently the lambda distribution with a power of 0.8 is used. This gives a 0.001 to mean ratio of 2.6, as opposed to the ratio of 4.12 for the log-normal distribution (with the assumption of a standard deviation of 0.5).

Work is currently underway on the prediction of which distribution (and the values of parameters required to apply that distribution) is best applied under any given set of conditions.

3A.7 Development of the Maximum Possible Slug Length Method

Under some conditions it may be possible to have only one slug in a line. This could result from stratified flow that could easily be disturbed by, say, a slugging well to form a flowline slug, or from very low frequency slug flow, or a relatively short pipeline. Having only one slug in a line is a special case that requires some additional analysis.

The mean slug length prediction described above assumes that each slug grows until it reaches the film shed from the slug ahead. If the time period between slug generation is significantly longer than the residence time of a slug in the line then the slug may not reach the film shed by the slug ahead of it. The slug length will then be dependent on the additional factor of pipeline length.

The procedure for calculating the 'maximum possible slug length' begins with the calculation of equilibrium stratified liquid holdup (H_{le}), as described earlier. The equations below are then used to calculate the maximum length.

V_s = velocity of liquid in slug body
V_{bf} = velocity of slug rear
V_{lf} = velocity of liquid in equilibrium stratified liquid film
L = pipeline length
L_{smax} = length of maximum possible slug

$$V_{sf} = \frac{V_s H_{ls} - V_{lf} H_{le}}{H_{ls} - H_{le}}$$

$$V_{bf} = 1.2 V_s + 0.35 (g D)^{0.5}$$
$$T = L/V_s$$
$$L_{smax} = (V_{sf} - V_{bf}) . T$$

This method is sensitive to the value of slug liquid holdup, but can be used as a good estimate of the maximum possible slug length that could be obtained in a situation with only one slug in a line. It is not currently available in MULTIFLO, but can be accessed via the Multiphase Flow Group, BP Exploration, Sunbury.

3A.8 Implementation and Use of the Slug Flow Design Method

There are two ways of using the BP slug flow design method. The first of these is via MULTIFLO, which is developed and supported by the Multiphase Flow Group and runs on a VAX computer. This group may be contacted to carry out design studies, or your local VAX cluster may be used to run the programme.

Contact: Phil Sugarman (MSMail Sugarman,Phil or phone +44 1932 762882)

The alternative is to run the model on a Hewlett-Packard workstation. This version contains a larger number of options for calculating slug holdup, frequency and length than are at present installed in MULTIFLO.

Contact: Trevor Hill (MSMail Hill, Trevor or phone +44 1932 763298)

3A.9 Anticipated Future Developments

1. The formulation of an improved method for slug liquid holdup prediction that takes into account water cut as well as gas and liquid superficial velocities and line diameter.

2. The broadening of the slug frequency method to cover inclined lines, and high viscosity liquids.

3. The completion of the slug length distribution model that will include the effect of gas and liquid flowrate on the distributions.

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3. A Model for Predicting Flow Regime Transitions in Horizontal and Near-Horizontal Gas-Liquid Flow,
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6. State of the Art Report on Multiphase Methods for Oil and Gas Pipelines,
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3B Slug Generation Dips

3B.1 Introduction

A significant effort has been made by BP, and by other oil companies and research establishments, to understand and characterise steady-state slug flow. Such a flow occurs with steady input gas and liquid flowrates and in pipelines that are normally assumed to be horizontal. In these cases slugs are formed from hydrodynamic waves, caused by a gas flow over a stratified liquid film, which rise to bridge the pipe. This work is covered in Appendix 3A.

Even for non-horizontal lines, or lines that change inclination, there is available to BP a qualitative assessment of the difference in behaviour from a straight horizontal pipe, at flowrates that would produce slug flow in the horizontal pipe.

A second classification of slug formation is required for systems with a low liquid loading and an uneven pipeline topography. If the pipeline were horizontal then the liquid flowrate would not normally be sufficient for slug flow to occur. However, if there is a dip at any point in the line it may be possible for liquid to accumulate until there is a sufficient quantity to produce a slug out of the end of the upward sloping line leading away from the dip.

This appendix addresses that possibility. It is based on work done largely for the Miller gas export line. This line normally runs in dense phase flow. However, in a depressurisation situation liquid will drop out. A prediction of possible slug sizes arriving at the downstream end of the line was required to enable an assessment of slugcatcher size to be made.

3B.2 Description and Theory

At low liquid flowrates, liquid transport out of a dip will definitely occur at a gas rate sufficient to produce annular-mist flow. This appendix concentrates on the liquid removal that occurs at much lower gas rates than the annular-mist transition when the dip angle is relatively shallow. The liquid may be removed as slugs or as a film.

The study began with experimentation. A volume of liquid was put into a shallow angle dip, with no further liquid inflow into the dip, and with no gas flow. As the gas flow was increased the liquid in the dip was disturbed - initially to form waves, and subsequently to form slugs if sufficient liquid was in the dip at the start. As the gas rate was increased further the liquid was produced out of the upward sloping line leading out of the dip, in the form of a film, slugs or drops, or a combination of these.

A number of quantities are defined in the analysis:

Critical gas velocity
That gas velocity above which all liquid is produced from the dip.

Maximum stable liquid accumulation
The maximum volume of liquid that can be retained in a pipeline dip at a given gas superficial velocity (i.e. no liquid production out of the end of the upward sloping pipeline segment).

Maximum random slug size
The largest liquid slug volume likely to be produced from a dip under steady-state operation.

(a) Critical Gas Velocity

This variable is calculated from the momentum balance equations for an equilibrium stratified film (due to Taitel and Dukler).

Gas phase

$$A_g \left[\frac{dP}{dx} \right]_g = -\tau_{wg} S_g - \tau_i S_i - A_g \rho_g g \sin(\theta)$$

Liquid phase

$$A_l \left[\frac{dP}{dx} \right]_l = -\tau_{wl} S_l + \tau_i S_i - A_l \rho_l g \sin(\theta)$$

At equilibrium

$$\left[\frac{dP}{dx} \right]_g = \left[\frac{dP}{dx} \right]_l$$

The equations may be re-arranged to give:

$$\frac{\tau_{wl} S_l}{A_l} = \tau_i S_i \left[\frac{1}{A_l} + \frac{1}{A_g} \right] + \frac{\tau_{wg} S_g}{A_g} - (\rho_l - \rho_g) g \sin(\theta)$$

For given fluid properties and pipe characteristics, the right hand side of the above equation is dependent on only the gas velocity and film depth.

The solution requirement is a gas velocity (V_{gscrit}) such that, at any liquid film depth in the upward sloping line downstream of the dip, the film flow direction is co-current with the gas. For the 2 inch air-water test rig the predicted value of V_{gscrit} corresponded closely to the observations.

One of the key parameters required during the calculation procedure is the gas-liquid inter-facial friction factor. This variable has an element of uncertainty in its prediction. Numerous efforts at a better definition have been, and are being, made.

(b) Maximum stable liquid accumulation

The magnitude of this quantity is dependent on the prevailing gas velocity, as well as the geometry of the pipeline. The calculation procedure is based round the growth and decay of slugs in the uphill section of line. The largest amount of liquid contained in a dip occurs when the uphill section is in slug flow, but with the slugs decaying just before the end of the section.

If the theoretical average slug decay rate over the uphill section length is equal to, or greater than, the average slug growth rate over the same section, then any slugs formed will decay by the section end. The total volume of liquid in the section will pass through each slug in this limiting case. Therefore, assuming that the liquid holdup in the slug is constant along the line, and that the slug rear velocity may be approximated by a steady-state correlation, the maximum liquid accumulation (Q_{eqm}) may be calculated as:

$$Q_{eqm} = \left[\frac{(0.16\ V_m + 0.59)}{(1.16\ V_m + 0.59)} \right] A\, H_{ls}\, L_p$$

H_{ls} may be evaluated using a steady-state correlation based on gas and liquid superficial velocities, A and L_p are the pipe section cross-sectional area and length of the uphill section respectively. This equation is valid for gas superficial velocities less than V_{gscrit}.

A rapid increase in gas velocity, or the passage of a pig through the dip, have the potential to sweep out the bulk of this maximum liquid accumulation.

(c) Slug frequency

Unlike the case described above, real flowlines generally have a liquid inflow into the dip from an upstream source. In this situation a balance will be set up in which liquid will be produced from the end of the uphill section of pipeline at the same rate (on average) as the liquid inflow into the dip. However, the liquid will not always be produced steadily. Sometimes it will be produced intermittently in the form of slugs. If this is the case it is necessary to calculate the slug formation frequency in order to be able to predict the volume of the slugs leaving the uphill section.

Slugs are formed at or just after the point of minimum pipeline elevation from the liquid film collected there as a result of both liquid inflow from the upstream source and the return of liquid shed from preceding slugs. Calculation of the slug frequency is dependent on these two liquid flowrates, the gas flowrate, and the depth of liquid at the pipeline low point required before slugs will form.

The assumptions currently made are that the liquid inflow from upstream is constant, and that the surface of the liquid accumulation in the dip is horizontal. However, it will be possible to develop the method to take into account cases when these assumptions do not hold.

The liquid flowrate draining back into the dip from a previously formed slug is calculated from a mass and momentum balance on the liquid film shed from the back of a slug. This flowrate is zero for a short period after slug formation, and then increases to a steady value as the distance of the back of the slug from the dip increases.

Slug formation will, on average, occur when the depth of liquid in the low point of the dip exceeds a certain critical value (dependent on the gas flowrate). This critical depth is calculated using the procedure developed by Taitel and Dukler for predicting the onset of slug flow from a stratified film.

Slug frequency is then determined by integrating the flowrate into the dip over time, converting that volume to a liquid depth, and stopping the procedure when the critical liquid depth is reached. The slug formation frequency is the inverse of the time required to reach the critical depth.

(d) Slug lengths leaving uphill section

The volume of slugs persisting to, and leaving the end of, the uphill section will, on average, equal the volume of liquid entering the dip over the time period between the formation of slugs.

It is, however, important to note that a real system has a degree of fluctuation about these average values. The length of slugs leaving the dip will be distributed about the mean value. A succession of short slugs would mean an increasing excess of liquid over the maximum stable liquid accumulation. A slug much longer than the average could then result, which decreases the liquid inventory in the system back down to the maximum stable accumulation.

A factor of 10 times the average is recommended as the safety margin in estimating the largest slug likely to be produced at the prevailing gas and liquid flowrates.

(e) Complicated topographies

The above discussion has centred on a dip with a steady inflow of liquid. In a real line it is possible that there will be several hills and dips, with unsteady liquid production from one dip to the next. The methods described above may be used to track the liquid movement through the system. However, the dependence of the output on the variation in slug length must be emphasised.

3B.3 Availability of Methods

The implementation of this method currently is a programme written for Hewlett-Packard series 200/300 computers running BASIC 5.0, available via Team 121 (Multiphase Flow) at the BP Exploration, Sunbury.

Principal Reference

A Mechanistic Model for the Prediction of Slug Length in Gas/Condensate Lines,
Wood, D G, RCS Branch Report 123 895 dated 255.90.

3C Severe Slugging

3C.1 Introduction

The possibility of significant surging occurring at low flowrates through a pipeline ending in a riser was first identified in the early 1980's. Work at Tulsa University by Schmidt, and in the field by Yocum, had shown that if the pipeline to the base of the riser was inclined downwards, and if the flowrates were low enough, then there could be a liquid accumulation at the base of the riser that would result in the severe slugging cycle described in the main body of the text.

At about the same time BP were planning the development of the SE Forties reservoir, by running flowlines from above the reservoir (originally from subsea templates, and finally from a minimum facilities platform) back to the existing Forties Alpha platform. These schemes involved flowlines sloping downward (about 0.1") to the base of a 120 metre high riser. Concern was expressed about the possibility of riser induced severe slugging, which led to experimental and theoretical investigations by BP Engineering and BP Research. The results of these investigations, and subsequent experience with the SE Forties flowlines, provide the basis for this appendix on severe slugging. The detailed reports covering the work are listed under 'Severe Slugging' in the bibliography of BP multiphase flow reports.

3C.2 Description of Severe Slugging

The four basic stages of severe slugging have already been described in the text. They are:

- Slug Generation
- Slug Production
- Bubble Penetration
- Gas Blowdown

A 50 mm diameter pipeline-riser system was constructed at the BP Sunbury site to investigate severe slugging. The pipeline section was 50 metres long and the riser 15 metres high. A video of this test rig, showing the different multiphase flow regimes, including severe slugging, is available, entitled Slugs in the Pipeline'.

Assuming that bridging (which occurs at the start of the slug generation stage of the cycle, when liquid completely blocks the bend to the pipeline-riser) has occurred it is relatively straightforward to describe the stability of the bridge (and hence whether severe slugging will occur at the given flowrates and pipeline inclination and length), and also the cycle time of stages (1) to (3). Characterisation of the fallback (during gas blowdown), and the prediction of the volume of liquid left at the base of the riser immediately after fallback and bridging, is more difficult, thus introducing a greater degree of uncertainty into the calculations of overall cycle time. The calculation procedure to determine the occurrence of severe slugging is described below.

The equation that describes whether or not severe slugging can occur at the given combination of gas and liquid flowrates, pipeline diameter, inclination and length, is derived as indicated below, and involves the following variables:

L = pipeline length (m)
L_s = length of liquid back-up from riser base (m)
D = pipeline diameter (m)
A = pipe cross-sectional area (m^2)
H = riser height (m)
β = angle of riser to horizontal (degrees)
h = length of liquid column in riser (m)
ρ_g = gas density at separator pressure (kg/m^3)
ρ_l = liquid density (kg/m^3)
V_{sg} = superficial gas velocity at separator pressure (m/s)
V_{sl} = superficial liquid velocity (m/s)
g = acceleration due to gravity (m/s^2)
M = gas molecular weight (kg/kg mol)
n = number of kg moles of gas in pipeline
R = universal gas constant (J/K/kg mol)
P = pipeline pressure (N/m^2)
P_{sep} = separator pressure (N/m^2)
T = pipeline temperature (K)
H_L = liquid hold-up in pipeline
V_g = gas volume in pipeline (m^3)

The gas and liquid mass flowrates into the start of the flowline are assumed to be constant. At the boundary of the severe slugging regime the liquid inflow to the start of the line exactly balances the rate at which liquid fills the riser after bridging (i.e. L_s = 0). The second condition for determination of the boundary is that the rate of gas pressure increase in the pipeline upstream of the bridge is produced exactly by the rate of gas inflow. The rate of gas pressure increase is equal to the rate of liquid head increase in the riser. The value of liquid hold-up in the pipeline is calculated from open channel flow equations, as the in-situ gas velocity is very low during this stage of the severe slugging cycle.

$$\frac{dP}{dt} = (\rho_l - \rho_g)\, g \,\frac{dh}{dt} \sin\beta \qquad (1)$$

$$\frac{dP}{dt} = \frac{d(nM)}{dt}\,\frac{RT}{M\,V_g} \qquad (2)$$

$$Vg = L\,(1 - H_L)\,A \qquad (3)$$

$$\frac{d\,(n\,M)}{dt} = V_{sg}\,\rho_g\,A \qquad (4)$$

$$\frac{dh}{dt} = V_{sl} \qquad (5)$$

Substituting (3) and (4) into (2), and then (2) and (5) into (1), and re- arranging, gives the equation quoted in the text, given that:

$$P_{sep} = \frac{\rho_g R T}{M} \qquad \text{and } \sin \beta = 1$$

$$\frac{V_{sg}}{V_{sl}} = \frac{L (1 - H_l) (\rho_l - \rho_g) g}{P_{sep}} \qquad (6)$$

If this equation shows LHS $\leq$ RHS then severe slugging will occur if the pipeline slopes downwards to the riser, and if the flow in the pipeline is predicted to be stratified at the prevailing in-situ superficial gas and liquid velocities.

This compared well with the observed boundary on the test rig, and also gave good predictions of cycle times (stages (1) to (2)) when equation (1) was integrated over the height of the riser.

The bubble penetration stage may also be modelled simply. If LHS < RHS in equation (6) then there will be liquid backed up the flowline (i.e. $L_s > 0$), as well as a full riser (h = H). Once the liquid column in the riser has reached the top of the riser (maximum liquid head at the base of the riser) the gas inflow at the start of the pipeline is no longer required to boost the pipeline gas pressure. The gas flow therefore begins to push the accumulated liquid along the flowline at a velocity that may be calculated from the inlet gas mass flowrate, the maximum pipeline pressure, and the liquid film hold-up.

$$\frac{dL_s}{dt} = V_{sg} \frac{P_{sep}}{(P_{sep} + \rho_l \, g \, H)} \frac{1}{(1 - H_L)}$$

The rate of liquid production from the top of the riser at this stage is simply:

$$= V_{sl} + \frac{dL_s}{dt} (1 - H_L)$$

Once the bubble front reaches the base of the riser there begins a period in which the liquid column in the riser is decreased in length, and is accelerated by the ever-increasing imbalance between the upstream gas pressure (which remains relatively high) and the decreasing liquid head. The acceleration of the liquid column in the riser is given by:

$$= \frac{g (H - h)}{h}$$

which obviously approaches infinity as the bottom of the liquid column reaches the top of the riser. Therefore there is a requirement to model downstream pipework and restrictions, which determine the upper limit of the liquid production rate.

An empirical approach was used during the test rig studies to estimate the liquid left in the riser, after bubble penetration and gas blowdown, at the time of bridging. This liquid would then fall back to the bottom of the riser to produce a liquid column of a height that is used as the starting point in the integration of:

$$\frac{dh}{dt}$$

to obtain a cycle time.

Obviously the field examples are more complicated - the gas may not obey the ideal gas law, there may be gas transfer into and out of solution in the liquid with pressure change, the pipeline might not be of a constant inclination, and the length of the line might be such that there is a variation of pressure and liquid hold-up along the line.

The simulator used to make calculations for SE Forties, whilst being founded on the physical principles described above, also seeks to take into account some of the complications not present on the test rig. However, equation (6) still provides a rule of thumb for part of the severe slugging boundary for more complex systems.

3C.3 Test Rig Studies

The test rig described above was used to develop a flow regime map for the pipeline-riser, flowing air and water with the separator at atmospheric pressure. Good agreement with the theoretical boundary, and with the theoretical cycle times was found. This gave the necessary confidence in our basic understanding of the phenomenon to enable some of the more detailed requirements of the field case to then be worked on in the VAX-based simulator.

Measurements were made of the liquid flowrate into the separator during the bubble penetration stage. As described above, the mathematical models need to have some pipework or restriction downstream of the riser-top to make the simulations realistic otherwise there would be an infinite liquid production rate as the last portion of liquid was produced from the riser.

The test rig was also used to investigate the flow behaviour at low flowrates in systems with a horizontal line and an upward sloping line upstream of the riser-base. Both of these cases also produced significant surging, but not the classic severe slugging in which liquid completely blocks the pipeline-riser bend. In these cases there was a liquid accumulation, but with gas bubbling through at all times. This observation is significant, however, especially for deep water systems (e.g. Gulf of Mexico) where there is the possibility of a significant liquid head build-up in the riser at low flowrates even if the pipeline slopes upward to the base of the riser.

3C.4 Ways to Eliminate Severe Slugging

There are a number of ways to reduce the extent of, or completely eliminate, severe slugging resulting from low flowrates in a pipeline-riser system.

(a) Pipeline Orientation

The first is to ensure that the pipeline route slopes up to the base of the riser, at least over a distance immediately upstream of the riser that is several times the riser height. This may mean a slightly longer pipeline route than a direct line, but the extent of the surging at low flowrates will be reduced, as full severe slugging cannot occur with an upward sloping line to the riser base.

(b) Gas *Injection*

The second technique is to have the facility to inject extra gas into the base of the riser. Gas injection into the base of the riser continually lifts liquid out of the riser, preventing the build-up of liquid and subsequent seal to the gas flow.

Test rig studies were carried out on riser-base gas injection after the work described above on the severe slugging cycle. The principle was demonstrated successfully. Even gas injection rates that are not sufficient to lift liquid continually out of the riser are of benefit as the gas reduces the maximum head achievable in the riser, therefore reducing the cycle time and severity of the severe slugging cycle. The amount of gas injection needed to prevent severe slugging is such that the combined injected and produced gas flowrate up the riser must put the riser into the churn flow regime at the given liquid production rate.

Studies were also carried out to optimise the design of the gas injection tee, to minimise erosion. The final design comprised a jacket around the riser into which the injection gas was directed. The gas then flowed into the riser through eight holes spaced equally around the riser. The holes were at the bottom of the jacketed length of riser.

Thirdly, gas injection into the start of the pipeline may be used to move the pipeline out of the severe slugging regime, by increasing the ratio of Vsg to Vsl above the critical value defined in equation (6). This either requires surplus gas at the pipeline start, or the addition of a gas line out from the pipeline end.

(c) Choking

A fourth method is to install a control valve on the top of the riser. On shutting this valve a position is reached whereby the frictional pressure drop across the valve acts to stabilise the gas-liquid flow up the riser. Any acceleration of liquid up the riser due to a decrease in liquid head in the riser (caused, say, by a gas bubble entering the base of the riser), is counteracted by the increase in frictional pressure drop across the valve as the liquid accelerates. The penalty of this way of eliminating severe slugging is that the pressure drop across the valve will be of the order of a riser height of liquid, thus imposing a significant extra back-pressure on the system at all times.

A secondary effect of the valve-induced back-pressure, as well as stabilising the riser, may be that the flow regime in the pipeline section is moved out of stratified flow towards bubble flow. This latter effect, if it was predicted to be sufficient to prevent severe slugging, could also be achieved solely by raising the separator pressure.

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3D Slug Flow in Hilly Terrain

Figure 5 shows a method for predicting slug characteristics in hilly terrain system

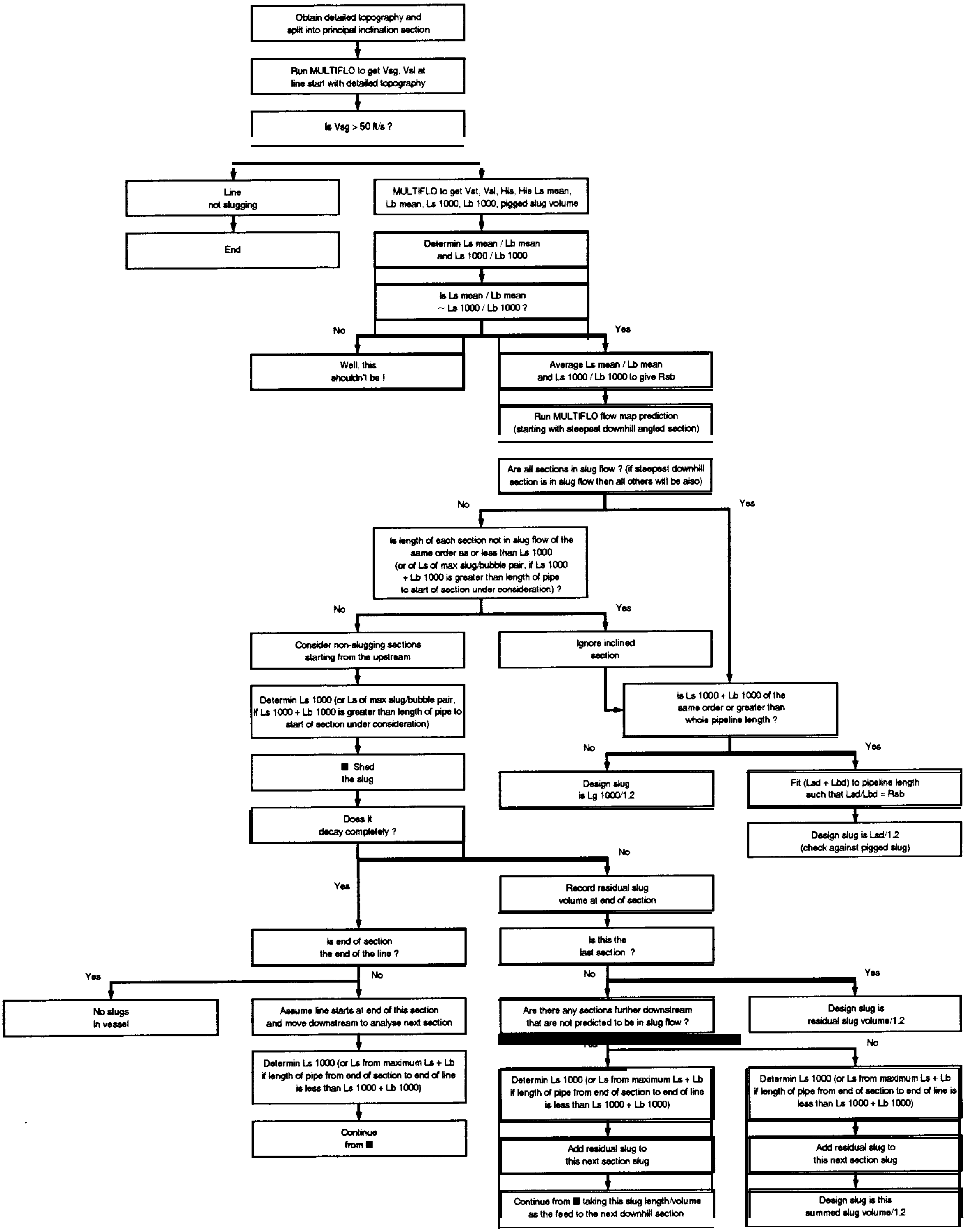


Figure 5. Hilly Terrain Slug Sizing Procedure

The following comments relate to this method, pointing out the sources of information and some of the areas in which further work is required:

1. L_s mean, L_b mean, L_s1000 and L_b1000 determined from MULTIFLO (in which the equilibrium holdup (H_{le}) method uses Blasias for gas and liquid friction factors, $F_i = F_g$ for inter-facial friction factor, and (V_g - V_l) for inter-facial velocity).

2. The frequency correlation which results in the the above lengths was derived for near horizontal only. Test rig studies, performed recently, show that above about 7 m/s V_{sg} the inclination (up to 2 degrees) has little effect, so not conservative at these velocities. However, for lower V_{sg}'s, down to 5 m/s, but same GLR, the rig is showing slug frequency not very dependent on V_{sl}, staying roughly at the 7 m/s V_{sg} (and associated V_{sl}) value. For horizontal lines slug frequency is highly proportional to V_{sl}. Application of the horizontal method to upwardly inclined lines will be conservative.

   i.e. if slug frequencies in upward lines are higher than horizontal at the same flowrates, which the test rig shows, then our horizontal slug lengths used in the attached logic tree will be conservative. Improvements to the prediction of slug frequency in inclined lines is a subject of our current R&D activity.

3. Hle and Fs for shedding calculations are calculated using Chen for gas and liquid friction factors, Fi=Fg for interfacial friction factor, and (V_g-V_l) for inter-facial velocity. This results in slight differences between MULTIFLO F_s and that used in the shedding spreadsheet.

4. Shedding model assumes that there is the equilibrium stratified flow in the downhill section. This is conservative, unless the slugs are very close together (in which case they wouldn't be 1 :1000 length). This is because the liquid flowrate required to sustain an equilibrium stratified film in the downward section cannot be supplied as a steady from an upward sloping line. The upward line will produce the liquid in slugs, which will decay to give a film. But it is very unlikely that the 1 :1000 slug will arrive at the start of the downhill section immediately after other slugs.

   i.e. slugs will probably shed more than we predict

5. In determining pigged slug volume, MULTIFLO uses Beggs and Brill for upward sections, and Mechan 92 for horizontal and downward. The consistency of this has yet to be verified.

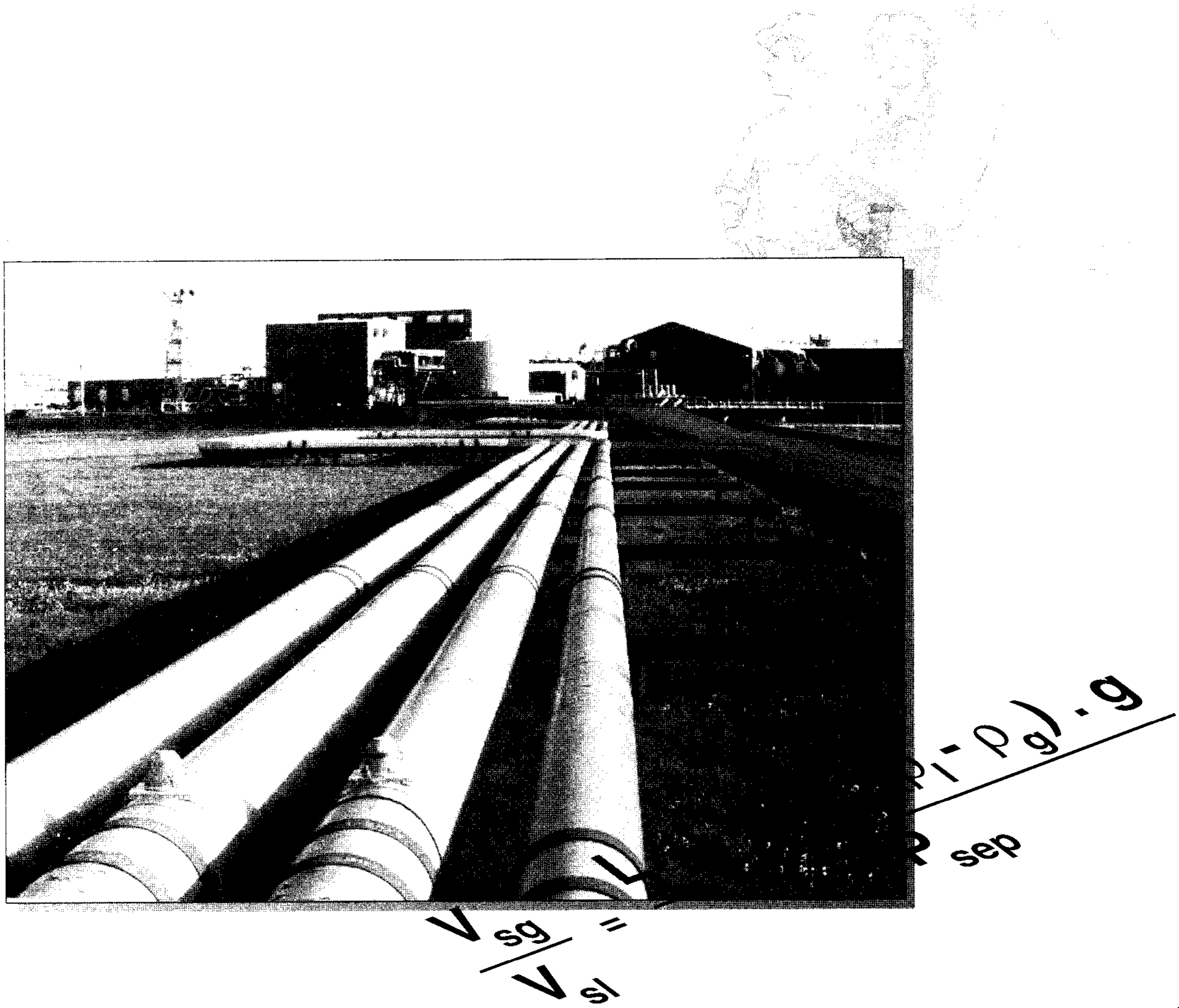
6. Despite 5, we still recommend that the design slug volumes from the logic tree be compared with the MULTIFLO pigged slug volume.

7. Whenever slug volumes are quoted they require a slug length and a slug holdup. These values are the MULTIFLO predictions. Slug lengths should be used in the assessment of each line section until the final delivery into a vessel, when length should be converted to volume.

8. All slug volumes at line end are divided by 1.2 to take into account shedding from the back of the slug as it enters the vessel.

Section 4. Gas/Condensates

Design of Multiphase Gas and Condensate Pipeline Systems



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4.1 Introduction

This section of the Multiphase Design Manual outlines the basic approach required for the design of gas/condensate flowline systems, such as that used for export of gas from the West Sole field to the Easington terminal. The details of the methods available for prediction of pipeline pressure drop, multiphase flow regime and also total and sphered liquid volumes are also included.

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4.2 Basic Design Approach

The criteria on which a gas-condensate system is based are usually a minimum receipt pressure at the downstream terminal, the available wellhead (or platform) pressure and the production profile.

A detailed fluid composition and flowline topography is also required to ensure that reliable results can be obtained using the design methods detailed in this section.

The following is a step by step approach to the typical calculations and considerations required in the design of a gas-condensate system.

4.2.1 Material Selection

The maximum flowrate for any line size in a gas-condensate system is often dictated by erosion rather than pressure constraints. As the erosional velocity limit is dependent on the pipe material, material selection should be conducted at an early stage. It is based on the fraction of acid gases present as well as the operating pressures and temperatures of the flowline. Typically gas with 0.2-2.0 per cent CO_2 at moderate pressures (300-1500 psia) would be transported via carbon steel flowlines with inhibition and a corrosion allowance. Significantly higher CO_2 content or pressures may require the use of corrosion resistant alloys.

The Materials Engineering Group in the company's Shared Resource at Sunbury is responsible for material selection.

4.2.2 Pipeline Sizing

Once the flowline material has been selected the minimum pipeline diameter can be calculated for the given production rates and pressure constraints. This work should be conducted using the Mechanistic Model within the BPX MULTIFLO program. This model is available from Version 11 of MULTIFLO and is described in detail in Section 4.3.

If the turndown in flowrates over the life of the line is large it may be necessary to constrain the production profile in order to reduce the diameter of the flowline. This is because the liquid volume that can accumulate in the system increases rapidly at low throughputs, leading to the necessity for frequent sphering or the requirement for large slug catchers.

4.2.3 Erosion Velocity Limits

Once a flowline diameter and material have been selected the maximum mixture velocity in the flowline must be calculated in order to check that its value does not exceed the erosional velocity limit. The mixture velocity calculation is conducted by the BPX MULTIFLO program. The method used to determine the erosional velocity limit is detailed in Section 3.1.4. It is based on a relationship given in API14E for droplet erosion. If sand is present, particulate erosion may be of greater significance than droplet erosion. Work, sponsored by BP, has been undertaken at Harwell Laboratories to determine the velocity limits for particulate erosion. The conclusions of this work, and a general discussion on erosion issues, is presented in the BP Erosion Guidlines documents, see Section 3, ref (10)

If the erosional velocity limit is exceeded either the line diameter must be increased or the production profile constrained to reduce the maximum mixture velocity. Alternatively a higher grade of pipeline material could be used - duplex stainless steel has a higher resistance to erosion than mild steel (Section 3.1.4).

As well as the erosional velocity limit, API14E also recommends that a line velocity of 60 ft/sec should not be exceeded to ensure that the level of noise emitted by the flow is not excessive.

4.2.4 Slug Catcher Sizing

Once the flowline material and diameter have been chosen, the required capacity of the slug-catcher at the reception facility must be determined. This is based primarily on the volume of liquid that is produced during sphering of the flowline. This sphered liquid volume is dependent on whether or not the line has reached equilibrium - ie. whether, for the given set of operating conditions, the total liquid in the line has reached its maximum value.

For non-equilibrium conditions the total liquid volume produced during sphering is assumed to be equal to the liquid that has entered the line since the launch of the previous sphere. This is dependent on the time interval since the last sphere was launched, the gas flowrate and the liquid loading:

$$V_{prod} = \frac{Q\,\phi\,t}{24} \quad \textbf{(non-equilibrium conditions)}$$

where:

V_{prod} = produced volume (bbls)
Q = gas flowrate (mmscfd)
ϕ = liquid loading (bbls/mmscfd)
t = time interval since previous pig launch (hrs)

The liquid loading is the ratio of liquid to gas and varies with pressure, temperature and composition. For this calculation it is normally calculated for the outlet conditions.

For equilibrium conditions the total liquid hold-up in the flowline is calculated using the Mechanistic Model in MULTIFLO. The volume of the liquid slug produced during sphering is less than the total equilibrium value as liquid flows out of the line during the passage of the sphere. The calculation procedure for determining the sphered slug volume is given in Section *4.4.*

The approximate time period required to reach equilibrium conditions is given by:

$$t_{eqm} = \frac{24\,V_{tot}}{Q\,\phi} \quad \textbf{(non-equilibrium conditions)}$$

where:

t_{eqm} = time interval to reach equilibrium (hrs)

V_{tot} = equilibrium liquid volume (bbls)
Q = gas flowrate (mmscfd)
4 = liquid loading (bbls/mmscfd)

The sphered liquid volumes for a given flowline system under both the equilibrium and non-equilibrium conditions can be plotted against gas flowrate on a design diagram:

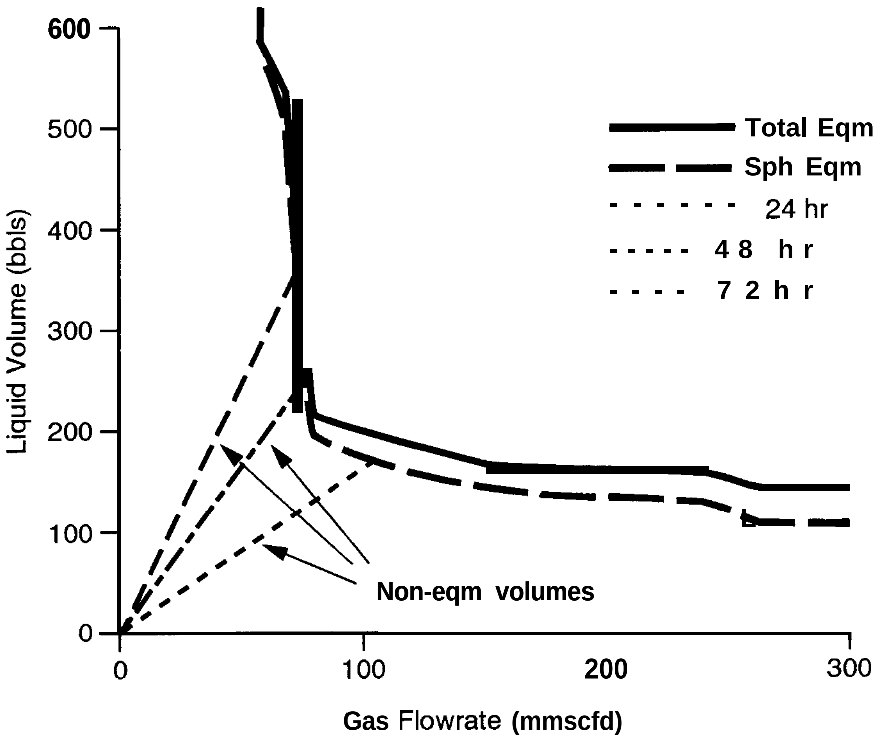

Figure 1.

This figure shows that for equilibrium conditions the sphered liquid volume decreases with gas flowrate, whereas for non-equilibrium conditions, and a set sphering interval, the liquid volume increases with flowrate. The maximum sphered volume can therefore be determined at the intersection of the sphered liquid volume lines for the equilibrium and non-equilibrium conditions. This is often used as a design basis for determining the required working volume of the slug catcher. The location of this intersection moves to a higher flowrate, and therefore a smaller value of the maximum sphered volume, as the sphering frequency is increased. Typically the intersection between the equilibrium and non-equilibrium lines for sphering interval of 24 hours will occur at approximately 50 per cent of the design throughput. If the intersection occurs at a higher flowrate the line is probably oversized. However, this may be unavoidable given the imposed constraints of production profile and allowable pressure drop. At a flowrate greater the value at the intersection, sphering is not necessary to limit the line's liquid inventory. At flowrates less than that at intersection the maximum time interval between spheres can be calculated from:

$$t_{sph} = \frac{24\ V_{sc}}{Q\ 4} \quad \textbf{(non-equilibrium conditions)}$$

where:

t_{sph} = maximum time interval between spheres (hrs)
V_{sc} = slug catcher working volume (bbls)
Q = gas flowrate (mmscfd)
ϕ = liquid loading (bbls/mmscfd)

The design diagram shows that, although the sphered liquid volume tends towards zero as the gas flowrate increases, the sphering operation is still predicted to produce a liquid slug even at high flowrates. However, field data at high flowrates has indicated that above a certain critical gas velocity (V_c) no slug is produced on sphering. An empirical correlation based on field data gives:

$$V_c = 15.2 \left[\sigma \frac{(\rho_l - \rho_g)}{\rho_g^2} \right]^{0.118} D^{0.414}$$

where:

V_c = Critical gas velocity (m/s)
D = pipe diameter (m)
ρ_g = gas density (kg/m^3)
ρ_l = liquid density (kg/m^3)
σ = surface tension (N/m)

The discrepancy between the present model and field measurements at these high flowrates may be due to the model under-predicting the liquid entrainment in the gas flow (see Section 4.3), liquid leakage passed the sphere (see Section 4.4) or an underprediction of the liquid film velocity.

4.2.5 Corrosion Inhibition

Corrosion inhibition, using chemicals such as filming amines, is necessary in gas-condensate systems when the required corrosion allowance for an uninhibited system is excessive. If this is the case a corrosion allowance for the inhibited system can be calculated from the allowance for the uninhibited system and the inhibitor efficiency (typically 85 per cent). If the required corrosion allowance is still excessive a higher grade material must be used for the pipeline. Calculation of corrosion allowance and choice of inhibitor is conducted by the Materials Engineering Group.

Corrosion inhibitors will only perform at their nominal efficiency if they remain in contact with the pipe wall. However, under adverse flow conditions inhibitor stripping can occur. For single phase flow inhibitor effectiveness is only guaranteed in straight, horizontal flowlines below a critical shear stress value. Materials and Inspection Engineering Group in their Corrosion Guidleline Document [See Section 3, ref(11)] recommend a maximum wall shear rate to avoid inhibitor stripping of 100 N/m^2.

The critical shear stress limit is also applied to multiphase systems, where it can be compared with a nominal shear stress value determined from the calculated frictional pressure gradient:

$$\tau_w = \frac{D}{4}\frac{dPf}{dx}$$

where:

τ_w = nominal wall shear stress (Pa)
D = pipe diameter (m)

$\frac{dPf}{dx}$ = frictional pressure gradient (Pa/m)

This relationship between shear stress and frictional pressure gradient is based on homogeneous flow. It always provides a conservative estimate for the actual shear stress in a steady stratified flow - the regime in which gas-condensate flowlines primarily operate. However, the wall shear stress could be substantially higher when slug flow is encountered, on the outside of bends or on the downstream side of weld beads. To estimate the shear stresses in these locations computational fluid dynamics using the FLUENT code can be employed. The Multiphase Flow Skills Group have extensive experience in its application.

At present BP has an ongoing programme of research into the factors which dictate corrosion inhibitor performance in multiphase flow. This work may indicate that inhibitor effectiveness is dependent on factors additional to the shear stress.

As gas-condensate pipelines operate primarily in stratified flow regime, corrosion inhibitor is only deposited on the pipe's top wall if there is a substantial fraction of the liquid entrained as droplets in the the gas phase. However, corrosion will still occur due to the condensation of water. For an uninhibited system the rate of corrosion at the top of the line will be lower than for the of the base of the pipe because, after condensing, the water will rapidly fall back to the stratified liquid layer under gravity. De Waard and Milliams (1) have indicated that a conservative estimate of top of line corrosion is 10 per cent of the overall uninhibited corrosion rate. This has been confirmed by experimental work conducted at Sunbury.

A top of line to bottom of line corrosion rate ratio of 0.1, for an uninhibited system, implies that corrosion considerations will be dictated by the bottom of line rate for inhibited systems, unless the inhibitor efficiency is better than 90 per cent.

4.2.6 Hydrate Formation

At low temperatures and high pressures water and gas can combine to form hydrate crystals. These crystals can grow to block the flowline. To ensure hydrate formation is avoided, the flowline should be designed to operate outside the hydrate formation region throughout its length. Insulation requirements to ensure that the line temperatures is maintained outside this region can be determined using MULITFLO.

If it is not possible to design to avoid the hydrate formation region, hydrate inhibitors such as methanol or glycol can be employed. Inhibitors will often be required at low throughput, and also during a controlled shut-down. If no inhibitor injection is possible prior to shut-down, hydrate plugs may form in the flowline. In these circumstances depressurisation may be required to ensure the line conditions are outside the hydrate region.

The boundary of hydrate formation region, for static equilibrium conditions, can either be determined experimentally at Sunbury or, for a known fluid composition, using the HYD01 module in BP's GENESIS flowsheet simulator. Under flowing conditions, the hydrate region will be smaller than that determined for static equilibrium conditions, and therefore both methods are conservative.

4.2.7 Solids Transport

Sand is rarely encountered in gas-condensate systems. However other solids, such as proppant, may be present following well fracturing operations.

If solids are present the minimum velocity to ensure their continuous transport must be determined. As, in gas-condensate flowlines, the actual liquid velocity will nearly always be less that given in Section 3.1.4 for solids transport (3 ft/s), it is normally assumed that the solids are transported by the gas phase only. However, the transport of solids in gas-condensate systems is, at present, little understood. Developments in this field will be included in the Section 12 of this Multiphase Design Manual.

Section 4.2.3 discusses the effect of solids transport on erosion.

4.2.8 3-Phase Flows

In water wet gas-condensate systems it can normally be assumed that the condensate and water phases are well mixed. However, when the water loading is a substantial proportion of the total liquid loading, the water and condensate can separate out at low flowrates leading to the accumulation of stagnant water at low points in the line. The presence of stagnant water can cause excessive corrosion at these low points. It could also lead to the production of water slugs during normal line operation, flowrate changes or sphering.

The conditions under which water drop-out occurs at low points in flowlines need to be predicted to assess the potential for corrosion and correctly design water handling facilities. The required modelling techniques for 3-phase flows will be discussed in Section 14 of the Multiphase Design Manual.

4.2.9 Transient Operation

Sections 4.2.1 to 4.2.8 discuss the design of gas-condensate flowlines based on operation under steady state conditions. However, the response of the system to start-up, shut-down and flowrate changes must also be considered. For example, liquid slugs are often produced during flowrate increase or line depressurisation. The volume of these transient slugs must be determined to ensure that they can be handled by the downstream separation and processing facilities. If there is insufficient capacity, operating procedures can be introduced to limit the rate of change of flowrate and therefore the size of any produced slugs.

This type of transient analysis can be conducted using transient simulators, such as PLAC, that are currently under development. A full discussion on the use of transient simulators will be included in Section 7 of the BP Multiphase Design Manual.

The compressibility of the fluids in gas-condensate systems can also lead to linepack. For a given outlet pressure the average flowline pressure, and therefore the line's capacity to store gas, increases with flowrate. Thus, an increase in inlet flowrate leads initially to linepack with no immediately impact on the flowrates at the outlet of the line. Equally, if the inlet flowrate is reduced, line depressurisation ensures that the outlet flowrates can be maintained for a period of time that depends on the overall line volume and operating pressure. Linepack or line depressurisation also occurs when the outlet pressure is changed but the inlet flowrate is kept constant.

If the line capacity is sufficient, linepack can be used, in combination with pressure variations at the flowline outlet, to allow steady conditions to be maintained at the inlet whilst supplying the fluctuating daily demand for gas. It also allows the gas supply to be maintained even if the production facility is shut down for short periods.

Line pack and depressurisation can be studied using the TGNET transient code. The Multiphase Skills Group maintain a license to use TGNET and have extensive experience in its application.

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4.3 Mechanistic Model

This section describes the Mechanistic Model for the prediction of pressure drop and total liquid hold-up for systems that have reached equilibrium and are operating under steady state conditions. This model is included within the BPX MULTIFLO program as "Segregated Flow – GRE mechanistic model" (1992). The method is used to initially determine the flow regime in each section of the flowline. The calculation procedure for pressure drop and liquid hold-up (the proportion of the flowline volume or cross-sectional area occupied by the liquid phase) is dependent on this predicted flow regime.

4.3.1 Flow Regime Determination

The flow regime at any point in the flowline is determined primarily using the Taitel-Dukler method (2). This model identifies the following five flow regimes:

(a) Stratified-smooth
(b) Stratified-wavy
(c) Intermittent (or slug)
(d) Annular-mist
(e) Dispersed-bubble

The determination of flow regime is conducted in two stages.

Initially stratified-smooth flow is assumed and an equilibrium liquid level is calculated. This is the stable liquid level that would occur if the flow were stratified with a smooth interface and no droplet entrainment in the gas-phase. It is determined by equating the pressure gradient in the liquid and gas layers.

Once the equilibrium liquid level has been calculated the stability of the liquid film is considered. At low flowrates no waves form on the film surface and the stratified-smooth flow is stable. At higher flowrates stable waves may form which do not grow to block the pipe - the stratified-wavy regime. If unstable wave growth does occur, slugs will form and the flow regime is intermittent. At high gas and low liquid flowrates there may be insufficient liquid in the system for slugs to form. In this case unstable wave growth leads to annular-mist flow. Finally for high liquid rates and low gas rates the gas is entrained as bubbles within the liquid flow -the dispersed-bubble regime. The Taitel-Dukler method models these mechanisms in order to determine the liquid and gas superficial velocities at which each of these transitions occur. The flowrate at the transition is a function of the fluid properties, the flowline diameter and the flowline inclination.

For the mechanistic model several modifications have been made to the original work of Taitel and Dukler. The method for determining the position of the intermittent/annular boundary has been modified following work by Barnea (3). Also, when determining the equilibrium liquid level, the interfacial shear stress term is calculated by setting the value of the inter-facial velocity to be equal to the liquid film velocity and not zero. Finally, work by Landman (4) has shown that there is more than one possible value for the equilibrium liquid level under certain special conditions - ie. low liquid loadings and upwardly inclined flowlines. If this is the case the mechanistic model uses the smallest of the three values, as this is considered to be the stable solution for real systems. The predictions of the stratified/intermittent boundary position may also be improved in the future following work conducted at BP Research (5).

The method of Taitel and Dukler has also been extended to include two additional flow regimes:
(f) Liquid
(g) Gas

and divide the intermittent regime into two sub-regions:

(cl) Normal slug
(c2) Terrain-induced slug

The liquid and gas flow regimes are simply used to identify single phase gas and liquid flows. These are defined to be when the no-slip liquid hold-up is greater than 0.9999 or less than 0.0001. The no-slip liquid hold-up is the volume proportion that would be occupied by the liquid if the flow were fully mixed.

The division of the intermittent regime is only relevant for sections of flowline with positive (upward) inclinations. The influence of the gravity term in uphill sections of flowline leads to the formation of slugs at much lower flowrates than would be the case for a horizontal line. Whereas slugs formed at higher flowrates (normal slugs) will be relatively unaffected by changes in flowline inclination, those formed at low flowrates in uphill sections of flowline (terrain-induced slugs) will not be stable in horizontal or downhill sections. Terrain-induced slugs are also characterised by the counter-current nature of the film that is shed from the rear of each slug. The differences between normal and terrain-induced slug flow leads to the necessity to employ separate methods for the calculation of liquid hold-up and pressure drop.

Gas/condensate systems generally operate in the stratified-smooth, stratified-wavy, annular-mist or gas flow regimes. Terrain-induced slugging may also be encountered in uphill sections of flowline at low flowrates. The remaining flow regimes (normal slug, dispersed bubble and liquid) are only encountered in black-oil systems.

4.3.2 Pressure Drop/Liquid Hold-up Calculations

(a) Stratified and Annular Flow Regimes

The approach used for calculating the pressure drop and liquid hold-up in the stratified-smooth, stratified-wavy and annular-mist flow regimes is identical. In each case a segregated flow is assumed and the liquid hold-up and pressure drop are determined using a similar calculation procedure to that employed when determining the equilibrium liquid level for flow regime identification (Section 4.3.1).

However, for stratified-wavy or annular flows the assumptions of a smooth interface and no liquid entrainment in the gas phase are not correct. To take account of the effects of inter-facial waves and droplet entrainment the pressure drop equations in the gas and liquid layers have been modified.

Inter-facial waves are accounted for by using the inter-facial friction factor equation of Andreussi and Hanratty (6). This leads to the prediction of greater losses through an increased inter-facial stress term. The overall effect of using the Andreussi-Hanratty equation is to increase the predicted pressure drop and decrease the predicted liquid hold-up at high flowrates.

The entrainment of liquid droplets in the gas phase can be accounted for by treating the gas/droplet mixture as a single modified gas phase. The density and viscosity of this gas/droplet phase are calculated assuming that the mixture is homogeneous:

$$\rho_{ge} = C_d \, \rho_l + (1 - C_d) \, \rho_g$$

$$\mu_{ge} = C_d \, \mu_l + (1 - C_d) \, \mu_g$$

where:

C_d = droplet concentration
ρ_g = gas density (kg/m^3)
ρ_l = liquid density (kg/m^3)
ρ_{ge} = density of gas/droplet mixture (kg/m^3)
μ_g = gas viscosity (kg/m/s)
μ_l = liquid viscosity (kg/m/s)
μ_{ge} = viscosity of gas/droplet mixture (kg/m/s)

In these equations the droplet concentration is the ratio of the cross-sectional area occupied by the liquid droplets to that occupied by the gas/droplet mixture. It is related to the entrainment fraction (the ratio of the liquid mass flux transported as droplets to the total liquid mass flux) as follows:

$$C_d = \frac{v_{sl} \, E_m}{v_{sg} + v_{sl} \, E_m}$$

where:

E_m = entrainment fraction
v_{sl} = liquid superficial velocity (m/s)
v_{sg} = gas superficial velocity (m/s)

The treatment of the gas and droplets as a single phase is based on the following assumptions:

- The droplets are transported at the velocity of the gas. Thus drag between the gas and the droplets can be ignored.
- There is no momentum transfer between the gas/droplet phase and the liquid layer due to the droplet entrainment and deposition processes.
- The droplet entrainment and deposition processes do not need to be modelled separately but can be accounted for within the calculation of the entrainment fraction.

Unfortunately there is little entrainment fraction data available for horizontal stratified and annular flow. The method used within the mechanistic model is based on data obtained by Shell on their high pressure Bacton flowloop (7). There is evidence that this method may under-predict the entrainment fraction at high gas flowrates.

The inclusion of entrainment fraction also leads to an increase in pressure drop and a decrease in the liquid hold-up predictions.

(b) Gas, Liquid and Dispersed Bubble Flows

The calculation of the liquid hold-up and pressure drop in the liquid, gas and dispered bubble flow regimes is conducted by assuming the flow is a homogeneous mixture. The flow is treated as a single-phase with the mixture density and viscosity:

$$\rho_m = \frac{v_{sl}\,\rho_l + v_{sg}\,\rho_g}{v_{sl} + v_{sg}} \quad \text{and} \quad \mu_m = \frac{v_{sl}\,\mu_l + v_{sg}\,\mu_g}{v_{sl} + v_{sg}}$$

where:

ρ_m = mixture density (kg/m^3)
ρ_g = gas density (kg/m^3)
ρl = liquid density (kg/m^3)
μ_m = mixture viscosity (kg/m/s)
μ_g = gas viscosity (kg/m/s)
μl = liquid viscosity (kg/m/s)
v_{sl} = liquid superficial velocity (m/s)
v_{sg} = gas superficial velocity (m/s)

(c) Normal Slug Flow

At present no mechanistic model is available for accurately calculating the pressure drop and hold-up in normal slug flow. Thus, the empirical method of Beggs and Brill is used when normal slug flow is predicted (see Section 3.1.3).

(d) Terrain-Induced Slug Flow

If terrain-induced slug flow is predicted the hold-up and pressure drop are calculated from the maximum stable liquid accumulation determined in Appendix 3B.

This maximum stable liquid accumulation is the maximum volume of liquid that can be retained in an uphill section of flowline at a given gas velocity. It is calculated by assuming that slugs form at the start of the uphill section of flowline, pass through the line shedding liquid from their rear and then collapse just before they reach the end. The shed film forms a counter-current liquid film which flows back to the start of the uphill section of flowline where it builds up to form the next slug.

Liquid flowing into the uphill section of flowline leads to the continuation of short slugs into the next section. However, the liquid retained in the uphill section remains equal to the maximum stable. liquid accumulation.

For terrain-induced slugging the liquid hold-up in an uphill section of flowline is simply equal to the ratio of the maximum stable liquid accumulation to the total volume of the pipeline section.

From Appendix 3B this is:

$$H_l = \frac{0.16v_m + 0.59}{1.16v_m + 0.59} H_{ls}$$

where:

H_l = liquid hold-up
H_{ls} = slug hold-up
v_m = mixture velocity (m/s)

The slug hold-up is calculated using a steady state correlation - at present the method of Gregory is employed (Appendix 3A).

The calculation of pressure drop for terrain-induced slugs assumes that the flow is homogeneous with the mixture density and viscosity based on the calculated hold-up:

$$\rho_m = h_l \rho_l + (1 - H_l) \rho_g$$

$$\mu_m = H_l \mu_l + (1 - H_l) \mu_g$$

This approach ensures that a good approximation to the hydrostatic component of the pressure gradient is obtained. This component is normally much greater than the frictional component when terrain-induced slugging is encountered.

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4.4 Sphered Liquid Volume

As discussed in Section 4.2.4 the sphered liquid volume, for a flowline that has not yet achieved equilibrium, can be estimated as the total volume of liquid that has entered the flowline following the launch of the previous sphere.

For the equilibrium case the sphered liquid volume is equal to the total liquid volume in the line, calculated using the Mechanistic Model, less that swept out of the pipeline as a liquid film during the sphere transit time. To determine the sphered liquid volume the growth of the slug in front of the sphere must be calculated, using the mass balance equation, throughout the sphere's passage through the line.

The process involved with the passage of a sphere through the flowline is shown below:

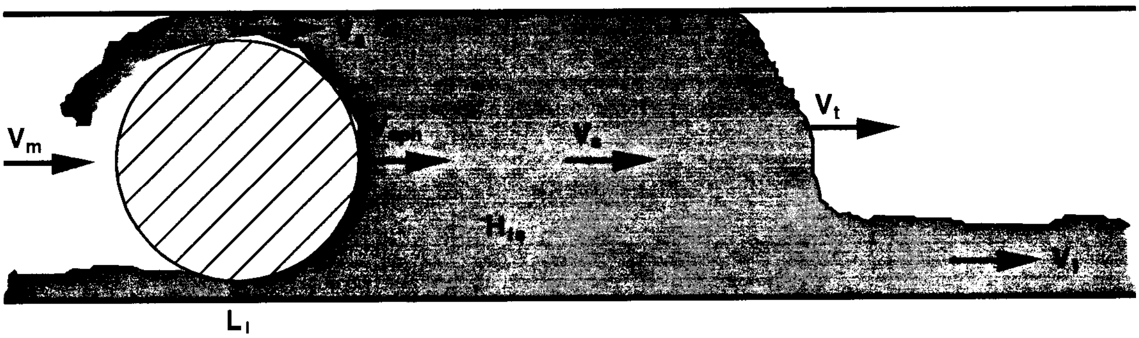

Figure 2. Growth of liquid slug during sphering growth

where:

H, = liquid film hold-up
H_{ls} = sphered slug hold-up
L, = sphere to pipe cross-sectional area ratio
v_a = velocity of liquid at sphere (m/s)
v_l = liquid film velocity (m/s)
v_m = mixture velocity (m/s)
v_s = sphered slug fluid velocity (m/s)
v_{sph} = sphere velocity (m/s)
v, = sphered slug front velocity (m/s)

The mathematical details of the approach required to calculate the sphered slug volume will not be discussed in this design method. However, the general approach is to determine the rate of growth of the sphered slug as it passes through the flowline collecting liquid from the preceding film. Liquid leakage passed a sphere that does not fully block the pipe can also be included.

Thus, a combination of three equations for:

(a) The mass balance at the front of the growing slug

(b) The mass balance at the sphere (if liquid leakage passed the slug is modelled)

(c) The growth of the slug due to the differential velocity between the slug front and sphere

lead to the required method for integrating the slug length as the sphere passes through the flowline.

For the special case of a horizontal flowline, zero pressure drop, no liquid leakage passed the slug and a sphered slug hold-up equal to unity, this equation reduces to:

$$V_{prod} = V_{tot} - V_{out}$$

where:

V_{tot} = total liquid holp-up in the line prior to sphering
V_{out} = liquid volume leaving the line during passage of sphere

This approximate form for calculating the sphered slug volume has been employed in the BPX MULTIFLO program.

4.5 Fluid Composition

The calculation of equilibrium and sphered liquid volumes and, to a lesser extent, that of the pressure and temperature drop in gas-condensate systems is very sensitive to fluid composition. For instance, the liquid loading of the line may double due to small changes in the proportion of the heavier components, leading to an increase in the predicted sphered slug volume of more than a factor of two.

Given the importance of using the correct fluid composition, black oil correlations are not suitable for gas-condensate design. Instead the FLODAT module of the GENESIS flowsheet simu lator should be used to produce a compositional datapack which is then accessed by MULTI-FLO.

To ensure that the correct pipeline composition is used for the production of the compositional datapack any saturated water and offshore processing must be taken into account.

4.5.1 Water Content

Normally the compositional data will be available for the dry wellstream only. However, the gas is normally saturated with water at reservoir conditions. To determine the actual water content of the fluid, GENESIS can be used to add water to the dry wellstream composition. By using the ASAP program in GENESIS, this wet composition can then be flashed at reservoir conditions, to determine the actual water content of the gas.

Although most gas wells have no free water at reservoir conditions, later in field life water breakthrough can occur and free water must be taken into account.

If the gas is dehydrated offshore using a glycol contactor, negligible water remains and the dry wellstream composition can be used.

However, if free water knock-out is used, the water fraction in the flowline is the saturated water content at the pressure and temperature of the water knock-out equipment.

4.5.2 Addition of Inhibitors

The liquid loading of a gas-condensate flowline can be substantially increased by the addition of inhibitors at the upstream end of the flowline. To ensure the correct liquid loading is used for gas-condensate calculations these inhibitors must be added to the composition. Unfortunately, GENESIS cannot correctly partition corrosion inhibitors between the oil and water phases. It must therefore be assumed that the inhibitors are contained entirely within the water phase. This can be carried out by substituting water for the inhibitor in the same volume ratio to the gas, and adding this to the composition determined in 4.51 by using ASAP.

4.5.3 Combined Compositions

Obviously if the production from more than one field is to be transported by a gas-condensate flowline, the combined composition must be used. Thus, compositional datapacks must be prepared for each ratio of flowrates from the two fields.

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4.6 Flowine Topography

The predictions of the Mechanistic Model, and also a second simpler mechanistic type model called Segreg, can be sensitive to the flowline topography at low flowrates. This sensitivity is a real effect and cannot be ignored. Thus, although empirical methods, such as those Oliemans for pressure drop and Eaton for liquid hold-up, are not sensitive to the line topography, this is due to their inability to correctly model the changes in flow regime and liquid level between horizontal and uphill sections of pipeline.

To ensure that the predictions of liquid hold-up and pressure drop are accurate it is therefore important to ensure that as detailed a topography as possible is used. It is generally the case that the liquid hold-up will increase asymptotically to the final value as the number of line segments is increased. Thus, calculations based on a simple approximation to the topography will usually lead to an under-estimate of the liquid hold-up.

The errors incurred by using an over-simplified approximation to the real topography will not be significant for flowlines in which inclinations are small (of the order of ±0.1°, or if gas velocities are high (greater than 30 ft/s). However, if large inclinations or low gas velocities are encountered the errors can be large. An example of this is the North East Frigg flowline. The depth of this flowline varies by up to 90 feet over its 11 mile length, with inclinations of between +3.7° and -3.7". Three topographies were used to model the flowline section, consisting of 1, 9 and 18 pipeline segments respectively. For the lowest flowrate, giving a gas superficial velocity of approximately 2.5 ft/sec, the calculated pressure drops for these three topographies were 35, 57 and 74 psi respectively and the liquid hold-up values were 101, 1284 and 2496 bbls. The measured values, which are themselves only approximate, were 72 psi and 2410 bbls – in close agreement with the most detailed topography used.

It is possible that the predictions would increase further if an even more accurate topography were available. However, as the 18 segment topography covered all the major inclination changes, any further increases in the predictions are likely to be small.

For reference the Olieman pressure drop correlation combined with the Eaton hold-up correlation predict a line pressure drop of 40 psi and liquid hold-up of 500 bbls with the detailed topography.

The North East Frigg example is an extreme case of low flowrate and large upward inclinations. However, it leads to the recommendation that all topography changes which result in a pipeline section that slopes upwards at more than +0.1° should be included in the topography to ensure reliable results.

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4.7 Slug Flow in Normal Operation

Section 4.4 detailed the method employed for calculating the volume of a liquid slug produced on sphering. Slugs may also be produced through the outlet of the line without sphering. These slugs can be caused by changes to the operating conditions of the flowline. If this is the case, the transient simulators briefly mentioned in Section 4.2.9, and discussed in detail in Section 7, can be used to predict the volume of slugs produced in this way.

Slugs can also be produced under steady state conditions if the flowline is predicted to be operating in the slug flow regime. For systems with a high liquid loading (at least 100 bbls/mmscf) normal slug flow is possible. In this case slugs can form in horizontal sections, and are likely to persist throughout the length of the flowline. The BPX slug flow model in MULTIFLO should be used to predict the length of these normal slugs. This method is detailed in Appendix 3A.

When terrain-induced slugging is predicted, slugs will only form in upwardly inclined section of flowline. These slugs are unlikely to persist throughout the length of the line. Instead they will steadily decay and then collapse in horizontal or downwardly inclined sections. However, the production of terrain-induced slugs at the outlet of the line may occur if the flowline is inclined upwards to the reception facilities. This is often the situation for lines between offshore fields and coastal reception terminals. For these circumstances the method based on slug generation in dips outlined in Appendix 3B of this manual should be used.

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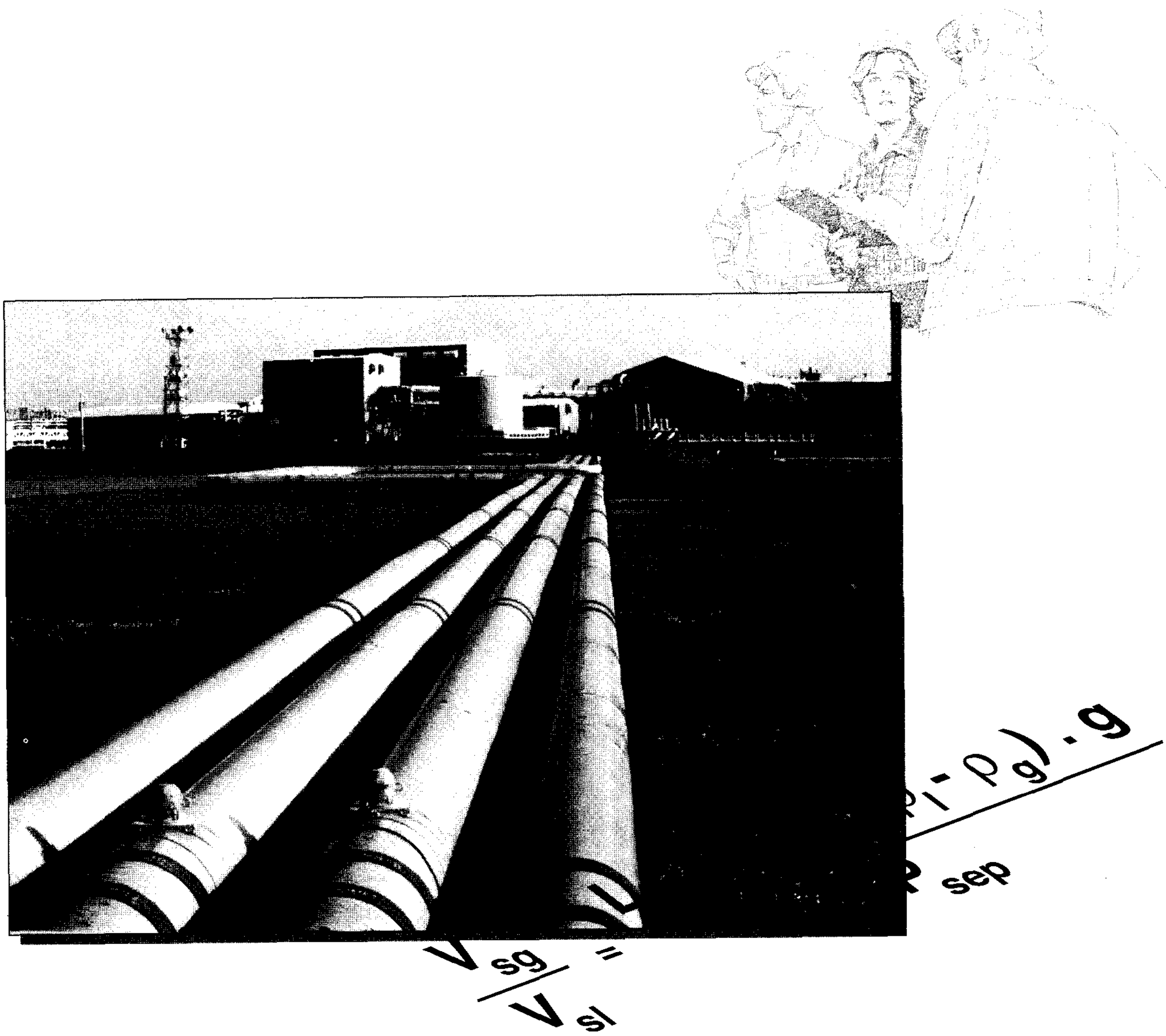
Section 5. Oil Line Water Slugs

Calculation of Pigged Water Slug Volumes from Oil Lines



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5.0 Foreword

This section of the Multiphase Design Manual describes a method for calculating the volume of a water slug produced when pigging an oil line, which contains small quantities of water. The method assumes that the water forms a segregated layer at the bottom of the pipe. This assumption is valid for most oil transport lines but under some high flowrate conditions at least some of the water will be carried along as entrained droplets. Under these circumstances the method given in Section 5.4 below will over predict the pigged water slug volume.

A second calculation procedure has been developed which checks if water will form a segregated layer. It determines the water concentration profile in oil/water pipelines and estimates the water layer depth.

This method has been incorporated into a computer program called WATPROF, which is available from the BPX Multiphase Flow Group. It is discussed in detail in Appendix 5A.

The following input data is required for WATPROF.

D Pipeline diameter (m)
E Pipe roughness (in)
U' Mean fluid velocity (m/s)
C Water cut (%)
μ_o Oil viscosity (Ns/m^2)
ρ_o Oil density (kg/m^3)
ρ_w Water density (kg/m^3)
σ Surface tension (N/m)

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5.1 Introduction

This Multiphase Design Manual describes the phenomena of water hold-up in water/oil multi-phase lines. This note details the steps required for the calculation of the slug volume of a water slug resulting from the pigging of the line.

Free water will tend to be collected at all points in the line at which the velocity of the water phase is retarded relative to that of the oil phase. This behaviour, termed ‘hold-up’, is reflected as an increase in the in-situ volume fraction of water (H,) above its flowing water fraction (C,). Hold-up principally occurs under conditions of stratified flow where the water forms a distinct layer at the bottom of the pipe. Where the concentration of water is relatively low, the separated water resides primarily within the pipe boundary layer. As a result, its average velocity is less that that of the overlying oil. This velocity difference is further increased by the greater inertial stresses acting on the water phase. The presence of uphill sections within a pipeline will further enhance the velocity differences, whilst downhill sections will reduce them. since the pig travels at the average velocity of the line the mechanism of water collection may be likened to one in which the pig ‘overtakes’ the slower moving layer.

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5.2 Nomenclature

The following nomenclature is used for the explanation of the calculation to estimate the water slug volume:

Q Oil flowrate (m^3/s)
C_w Inlet water content
H_w Equilibrium hold-up
t Time since last pig (s)
A Area of line (m^2)
L Length of line (m)
V Slug volume (m^3)
T_e Time for the line to reach equilibrium (s)

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5.3 Equilibrium Hold-up

The water hold-up fraction is influenced by the density and viscosity of both liquids, flowing water fraction, the pipe diameter, mixture flowrate, and the pipe inclination. Of these, the pipe inclination and the flowing water fraction are the most important.

Water hold-up increases with the flowing water fraction and the pipe inclination. As a general rule, when the flowing water fraction increases over 10% the flow regime is no longer stratified and the hold-up will tend towards the flowing water fraction. Water slugs will be created by accumulation and sweep-out, the size of which will mainly depend on the geometry of the line.

The calculation of the hold-up is complicated and is not described in this document but soft ware is currently under development to perform this calculation.

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5.4 Slug Volume

5.4.1 Whole Line at Equilibrium

The main assumption that all these calculations are based on is that no water leaves the line until the whole line has reached equilibrium. Therefore once equilibrium has been reached, the total volume of water in the system is;

$$A \times L \times H_w \qquad \text{Equation 1}$$

At the moment the line reaches equilibrium (the moment water starts to flow from the end of the line) the total water entered since the last pig will be:

$$Q \times C_w \times T_e \qquad \text{Equation 2}$$

Therefore:

$$A \times L \times H_w = Q \times C_w \times T_e \qquad \text{Equation 3}$$

$$T_e = \frac{A \times L \times Hw}{Q \times Cw} \qquad \text{Equation 4}$$

The volume of water produced as a slug will be the water content less the volume of water which flows out naturally:

$$V = A \times L \times Hw - Q \times Cw \times t'' \qquad \text{Equation 5}$$

where:

t" = the time it takes for the pig to pass through the line.

$$t'' = \frac{\text{volume of line}}{\text{volumetric flowrate}} = \frac{A \times L}{Q} \qquad \text{Equation 6}$$

Therefore:

$$V = A \times L \times H_w - Q \times C_w \times \frac{A \times L}{Q} \qquad \text{Equation 7}$$

$$V = A \times L \times (H_w - C_w) \qquad \text{Equation 8}$$

5.4.2 Line Yet to Reach Equilibrium

If the line has not yet reached equilibrium but if a volume of water with a volume greater than V has entered the line the slug volume will be the same as V. This is because no water can flow out of the line until the whole line has reached equilibrium. Therefore even as the pig is travelling along the line building up a water slug in front, new parts of the line will still be reaching equilibrium nearer to the exit of the pipe.

If the volume of water that has entered the line is exactly V when the pig is launched, then no water will leave the pipe until the slug (with volume V) arrives.

The time taken for the 'equilibrium slug volume' (V) to flow into the line is given as;

$$T_f = \frac{A \times L \times (H_w - C_w)}{Q \times C_w}$$ **Equation 9**

If $t < T_f$ then;

$$V = Q \times C_w \times t$$ **Equation 10**

If $t > T_f$ then;

$$V = A \times L \times (H_w - C_w)$$ **Equation 11**

5.4.3 Consecutive Pipelines

Consecutive pipelines means a system where one pipeline (the upstream pipeline) flows directly into a second pipeline (the downstream pipeline). It is assumed that the only production entering the downstream line is from the upstream line. It is assumed that a pig is not launched into the downstream line until a pig has arrived in the pig reciever at the end of the upstream line.

(a) Both Lines Pigged

If the pigging philosophy is such that when one pig arrives at the end of the first line another is launched into the second, then the calculation of slug volumes is simple. Since the calculation is based on volume, the area and length of the first and second line (A_1L_1 and A_2L_2) can be averaged to produce an average area and length (A_{ave} and L_{ave}). Therefore:

$$T_f = \frac{A_{ave} \times L_{ave} \times (H_w - C_w)}{Q \times C_w}$$ **Equation 12**

If $t < T_f$ then:

$$V = Q \times C_w \times t$$ **Equation 13**

If t > T then:

$$V = A_{ave} \times L_{ave} \times (H_w - C_w)$$ **Equation 14**

(b) First Line Pigged Only

If there are two consecutive lines and only the first is pigged, then the calculation of the pig size is based on only the first line in isolation of the second. Any slug which results from the pigging of the first line is assumed to pass through the second line unaffected. Any accumulation of water at the front of the slug of water as it travels along the second line is matched by the deposition of the water behind the slug. The equations 10 and 11 apply with L and A equal to the length and area of the first line.

(c) Second Line Pigged Only

If the second line is pigged without the first being pigged then the assumption that no water leaves the line before the whole line reaches equilibrium is no longer conservative. It should be assumed that no water is held up in the first line and again equations 10 and 11 apply with L and A equal to the length and area of the second line. This is conservative but without more field data it is the only practical method of determining the water slug volume.

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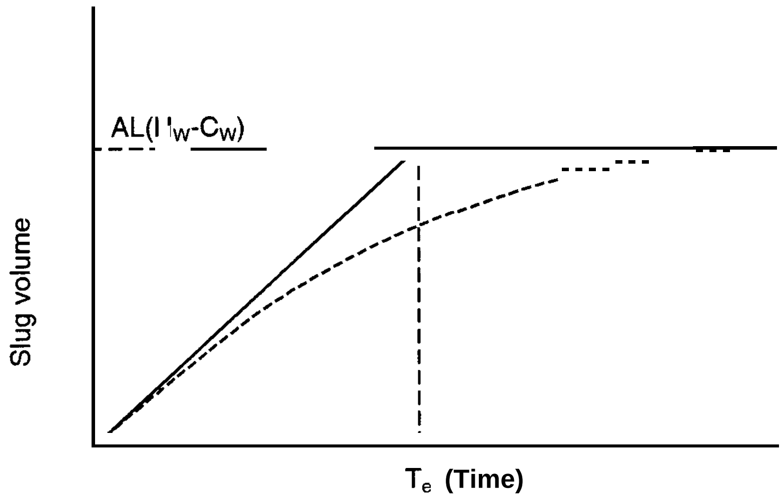
5.5 Conservatism

The assumption that no water leaves the line until equilibrium has been reached is conservative. In practice, water will flow from the line before equilibrium is reached, but it is assumed that the line reaches equilibrium as soon as sufficient water has entered the system. This results in the conservative (i.e. high) estimation of the water hold-up in the line and therefore conservative (i.e. again high) prediction of the water slug volumes when the line is pigged.

This is illustrated below in Figure 1, where pigged slug volume is plotted against time since last pig launch. The hard straight line illustrates the conservative assumption made here, ie liquid accumulates in the line, and does not exit the line, until the equilibrium slug volume (AL (H_w-C_w) has built up in the line in time T,. In practice it takes some time greater than T_e to accumulate this volume of liquid, as shown by the dashed curve.

If there is a significant period of time between each pig and the line has been at equilibrium for a significant period then the estimation of the slug volumes may not be so conservative. This is because the estimation of the equilibrium hold-up is not excessively conservative.



Te = Time for pipe (length L) to reach equilibrium
AL = Volume of pipe
H_W = Equilibrium hold-up
C_W = Inlet water fraction

Figure 1. Pigged slug volume

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5A Appendix: The Water Concentration Profile

5A.1 Nomenclature Used

c(y)	Water concentration at depth y from centreline. (dim.)
d_{95}	Droplet diameter corresponding to 95% volume of droplets smaller than d_{95}. (m)
f	Darcy-Weisbach friction factor.
K	Parameter: ($\equiv w/\zeta u_*$). (dim.)
u*	Friction velocity: ($= u'(f/8)^{0.5}$). (m/s)
u'	Mean fluid velocity. (m/s)
V'	Water:Oil ratio.
w	Water settling velocity. (m/s)
y'	Reduced vertical coordinate, y/R. (dim.)
ζ	Dimensionless diffusivity in pipe flow.
E(k)	Mathematical function dependent on k.

5A.2 Introduction to Karabelas

In steady horizontal pipe flow, the density difference between the dispersed solid or liquid phase and the continuous fluid phase can cause a non-uniform distribution of the dispersion in the pipe cross-section. This is the case for lines carrying oil / water mixtures. Karabelas (1977 and '78), looked at a vertical distribution of dilute suspensions in turbulent pipe flow and at droplet size spectra, generated in turbulent pipe flow of dilute liquid / liquid dispersions.

The WATPROF program uses some of the results derived in these papers. This appendix briefly examines the calculation method and assumptions made throughout.

5A.3 The Calculation and Assumptions

Karabelas (1977), obtained his experimental data for (Equation 1) using dilute suspensions of spherical particles, typically of about 0.5mm in diameter. This is applicable to oil / water dispersions, because usually there is a low concentration of water and consequently a droplet size of only a few millimetres at the maximum, which suggests the droplets will be approximately spherical.

The solution for the concentration distribution can be reduced to:

$$c(y) = 1 + \frac{E(K)}{V'} \times \exp(Kxy') \qquad \text{Equation 1}$$

if the droplet size is nearly uniform.

In this expression $K = W/\zeta u^*$ by definition. This parameter, and especially u^* essentially determine the magnitude of the vertical concentration gradients, with a high value of K indicating that water will tend to settle out.

The term E(K) depends only on K, and its determination is given in Karabelas(l977). V' is the ratio of water to oil in the fluid.

This approach does not take into account droplet coalescence. Hinze(1955), states that when there is a low concentration of the dispersed phase the chances of coalescing are small. Whilst he does not define 'low concentration', droplet coalescence should not be significant in most lines where the water concentration is less than 1 0–15%.

The program considers the water layer to be continuous when $c(y) \leq 0.52$. This value of 0.52 volume fraction is the closest packing density of spheres in a cubic lattice. This will obviously result in an over estimate in the depth of the water layer.

5A.4 Other Factors to Consider

(a) The droplet size, d_{95}, is calculated from a result in Karabelas (1977), and is needed to calculate the water settling velocity. Although this is based on an isotropic and homogeneous flow field, the results yielded are accurate enough for this application. Indeed, a sensitivity check showed that a 1% change in d_{95} would result in a change of 0.7% change in K. The effort required in obtaining a better value can not be justified in this case.

(b) In order to determine the shear velocity, u^*,the friction factor has to be calculated. The Chen equation is used instead of the iterative Colebrook-White implicit equation to do this. It gives satisfactory results over the full range of Reynolds numbers and relative pipe roughnesses.

(c) The value of the dimensionless lateral particle diffusivity, ζ ,is set to 0.25. This is taken from Karabelas (1977) and it is the average value from his experimental work. This is a satisfactory estimate for this calculation.

5A.5 Summary

This model is suitable for oil/water mixtures flowing in long horizontal, or near horizontal, pipelines in a steady state. It is most accurate for small water cuts but still satisfactory up to 15–20%.

No account is taken of the increase in oil velocity that is expected if the water layer is deep, so again the estimate is conservative. Despite this WATPROF is a useful program giving a good indication of water depth in theory.

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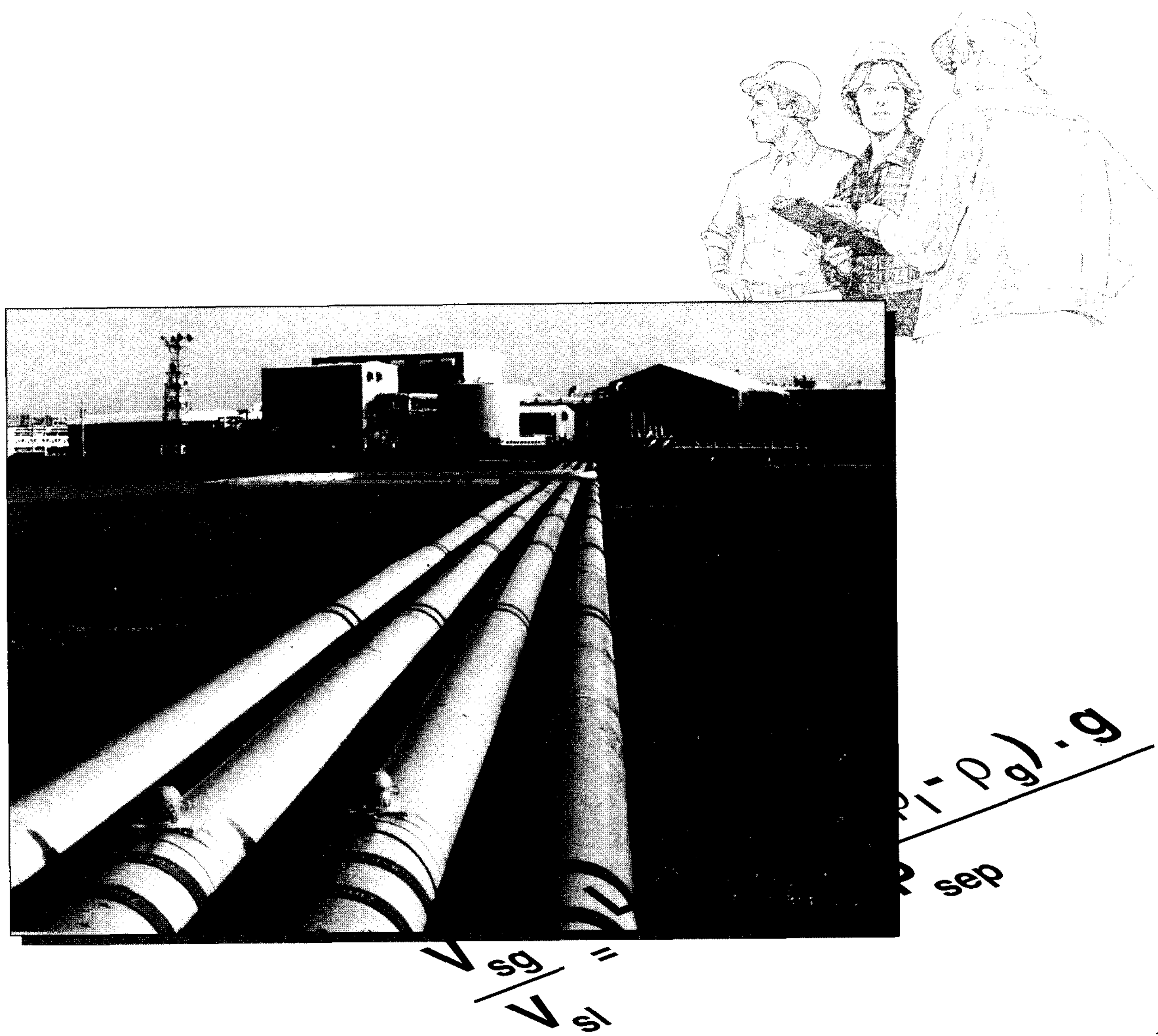
Section 6. Wellhead Temperatures

Prediction of Wellhead Temperatures under Steady State and Transient Conditions



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6.1 Introduction

This section of the Multiphase Design Manual describes why the accurate calculation of the Wellhead Flowing Temperature is important, it also outlines how the WHFT is calculated by BP's steady-state flowline and well simulation program, MULTIFLO. Section 6 is split into two parts, the first describes the factors that influence the WHFT and the theory involved, the second section (Appendix 6A) describes the more practical aspects involved in determining the well head flowing temperature using MULTIFLO.

It is important that the temperature at the wellhead is calculated accurately since it can greatly effect the design of the flowline and/or the separator downstream of the wellhead. For example, the following are influenced by the wellhead temperature:

- The relative vapour/liquid split in the separator or flowline (which may affect the sizing of the flowline or the separator).

- The possibility of hydrate formation.

- Rate of corrosion in the flowline.

- The temperature profile at the inlet to the flowline which effects the type of insulation and burial requirements. If the temperature is high then burial may be required to prevent upheaval buckling.

This design guide discusses the factors which affect the wellhead temperature, such as water cut, flowrate, and completion details. In addition, it summarises the calculation used by the two programs which are most commonly used within BP to calculate Well Head Flowing temperature: MULTIFLO and WELLTEMP:

- MULTIFLO is BP's steady-state two-phase simulation program. It does not rigorously model the heat transfer within the tubing and the casings. Instead, the casings and annuli are approximated to an equivalent thickness of cement (or another insulation material). This method provides accurate results considering that information on the heat transfer through the completion details and the thermal properties of the surrounding rock may not be readily available and/or of questionable accuracy. **MULTIFLO is BP's preferred method of calculating steady-state wellhead flowing temperatures.**

- WELLTEMP is a unsteady-state well simulation program. It rigorously models the temperature loss through the tubing, casing and the annuli with time. It can calculate the temperature profile in the well during start-up and as a result of flowrate changes. **WELLTEMP should not be used to calculate steady-state wellhead flowing temperatures but should only be used to give an indication of the warm-up rates or the thermal response to changes in the operating conditions such as flowrate.**

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6.2 Steady-State (MULTIFLO)

6.2.1 Introduction

This section describes the method used by MULTIFLO to calculate the WHFT. The description of the calculation will allow the reader to determine which factors most influence the WHFT.

6.2.2 Calculated Well Heat Transfer

The heat lost from a flowing well into the surrounding rock matrix is by time dependent heat transfer. As in the case of a flowline, the overall heat transfer coefficient, which determines heat loss from the produced fluids, is obtained from a series combination of three components:

- An Internal Film Coefficient

- A Composite Coefficient for the Tubing/Casing Comprising:
  - the resistance of the tubing and casing walls
  - the resistance of the annulus medium
  - the resistance of the casing cement grouting

- A time dependent coefficient for the surrounding rock matrix.

This summary does not include the contribution to the overall heat transfer coefficient of the heat transfer coefficient for radiation as it's influence is usually small compared with the other coefficients.

6.2.3 Heat Transfer Through Completion

The calculation of the heat transfer rate through the layers becomes quite complicated when a large number of casings are employed. The numerous casings, annuli and cement thicknesses each contribute to the overall resistance to heat transfer. To simplify the calculation, it is possible to approximate the numerous layers to a single layer of cement having the same resistance to heat transfer as the sum of the casings and annuli.

The heat transfer rate, Q, is described below. Where 'j' indicates the properties of the jth layer.

$$Q = \frac{2\pi\Delta T}{\Sigma\,[\ln(D_{J+1} / D_j) / k_j]} \qquad \text{Equation 1}$$

For a single layer of cement with a outer diameter of D,, the equation becomes:

$$Q = \frac{2\pi\Delta T}{\ln(D_c / D_{ci}) / k_c} = \frac{2\pi k_c \Delta T}{\ln(D_c / D_{ci})} \qquad \text{Equation 2}$$

Since Dc is the only unknown and the numerator is the same for both equations the equations can be rewritten as below:

$$D_c = D_{ci} \exp (k_c \Sigma_{j=1}^{n} ([\ln \{D_{J+1} / D_j\} / k_j])$$ **Equation 3**

The TOOLKIT option in MULTIFLO can be used to perform this calculation (use of the TOOLKIT option is described in Appendix A). The heat transfer coefficient through the completion can now be expressed (based on tubing inside diameter) as:

$$h_{comp} = \frac{2k_c}{D_i \ln (D_c / D_{ci})}$$ **Equation 4**

At the bottom of the well there is normally only one tubing, one casing and one thickness of cement (and one annulus). However, nearer the top of the well there may be numerous casings, annulus, different annular fills and cement thicknesses.

6.2.4 Internal Heat Transfer

The internal heat transfer coefficient, hi, assumes the existence of homogeneous two-phase flow. This assumption is usually acceptable since the resistance of the internal film to heat transfer is low compared with the resistance of the casings and the external medium.

The expression used for estimating the film coefficient is the same as that recommended for single phase turbulent flow.

$$h_i = \frac{0.023k_m}{D_i} \times Re^{0.8} \times Pr^{0.3}$$ **Equation 5**

The physical properties are derived on a phase volume weighted basis. The subscript 'm', denotes a mixture property. For example, the mixture viscosity is determined by the following:

$$\mu_m = \mu_l \lambda_1 + \mu_g(1 - \lambda_l)$$ **Equation 6**

(Since heat is usually lost from well fluids the exponent for the Prandtl number [Pr = $C_p \mu_m / k$ J is set to 0.3.)

6.2.5 External Medium

Heat transfer to the earth is essentially transient radial heat conduction, with the effective heat transfer coefficient varying with time. Whilst the analysis of the problem is relatively straightforward, a simple explicit solution is not available. However, approximate asymptotic solutions for 'short' and 'long' flow times can be derived.

The effective heat transfer coefficient, referred to the tubing inner diameter, is given by:

$$h_{earth} = \frac{2K_e}{D_i f(t)}$$ **Equation 7**

where, for short flow times (t<100 hours):

$$f(t) = [(\pi z)^{-0.5} + 0.5 - 0.25(z/\pi)^{0.5} + 0.125z] \times 2/D_{co}$$ **Equation 8**

$$z = \frac{D_{co}^{\ 2}}{\alpha t}$$ **Equation 9**

t = time (hours)

and for long flow times (t>100 hours):

$$f(t) = \ln(D_{co} / \sqrt{\alpha t}) - 0.29$$ **Equation 10**

MULTIFLO will allow the user to specify the flowing time used in the equations above and will therefore roughly calculate the WHFT for short flowing times (ie. warm-up of the well after start-up). However, this method will not provide an accurate prediction of the WHFT during start-up (although a rough indication is given). Therefore, WELLTEMP should be used to calculate well warm-up and not MULTIFLO.

For flowing times of the order of 10,000 hours, the system can reasonably be assumed to be approaching steady-state, and MULTIFLO should be used. For times between 500 (say) and 10,000 hours the well will probably be approaching steady-state, and the 'steady-state' value given by MULTIFLO will be a reasonable approximation.

6.2.6 Geotherm Gradient

The 'undisturbed' temperature of the surrounding soil will vary from the reservoir to the sea/surface. The soil temperature is important since it affects the rate of heat loss from the fluids (see Equation 12). However, the geothermal gradient is not usually known and it is customary to assume that the (unaffected) temperature varies linearly between the reservoir and the seabed.

6.2.7 Overall Heat Transfer

Using the coefficients calculated above the overall heat transfer rate can now be determined:

$$\frac{1}{U} = \frac{1}{h_i} + \frac{1}{h_{comp}} + \frac{1}{h_{earth}}$$ **Equation 11**

The heat transfer coefficients are based on the inside area. The heat loss from the fluid is therefore:

$$Q = UA\Delta T \qquad \text{Equation 12}$$

Both U and the temperature difference vary up the tubing and therefore a calculation is best solved by computer simulation.

6.3 Calculation of Wellhead Temperatures

6.3.1 Effect of Flowrate

Increasing the flowrate will increase the wellhead temperature. Qualitatively, as the flowrate increases, the heat 'lost' in the well (Q) remains roughly constant:

$$Q = UA\Delta T_{lm} \qquad \text{Equation 12}$$

The area is fixed and the overall heat transfer coefficient is relatively insensitive to flowrate since it is mainly controlled by the heat transfer through the surrounding rock. The Tlm will increase slightly since the temperature difference at the surface will increase. The following equation can be used when studying the heat loss from the fluid only:

$$Q = MC_{pm}(T_{in} - T_{out}) \qquad \text{Equation 13}$$

Therefore, rearranging Equation 13 gives:

$$(T_{in} - T_{out}) \propto 1/M \qquad \text{Equation 14}$$

Since Q and C_p can be taken as constant (or relatively insensitive to the flowrate), increasing M (the mass flowrate) will reduce the temperature loss in the tubing. **Therefore, increasing the mass flowrate will resulting in an increase in the WHFT.**

Equation 14 gives only a rough indication of the effect of flowrate since it fails to account for the effect of the geothermal gradient. At very low flowrates the well head temperatures tends towards the sea temperature as the fluid temperature tends towards the geothermal gradient.

6.3.2 Effect of Water Cut

Increasing the water cut increases the specific heat and mixture density of the fluid resulting in an increase in the WHFT. With reference to Equation 13 used above, by assuming that the heat lost from the well (Q) and the mass flowrate (M) to be relatively constant, increasing water cut will raise the specific heat (C_p) which will reduce the temperature loss and therefore result in warmer WHFT.

The predicted influence of the water cut and the flowrate is shown in Figure 1. This is a graph of the steady-state WHFT. Flowrates are shown up to 20 mblpd, or up to about 30 ft/s. The lines are asymptotic, and will approach the reservoir temperature (220°F in this case) at very high flowrates.

The temperature approaches the asymptotic value at low flowrates (5 mblpd or roughly 5 ft/s), and at flowrates above this the WHFT increases slowly.

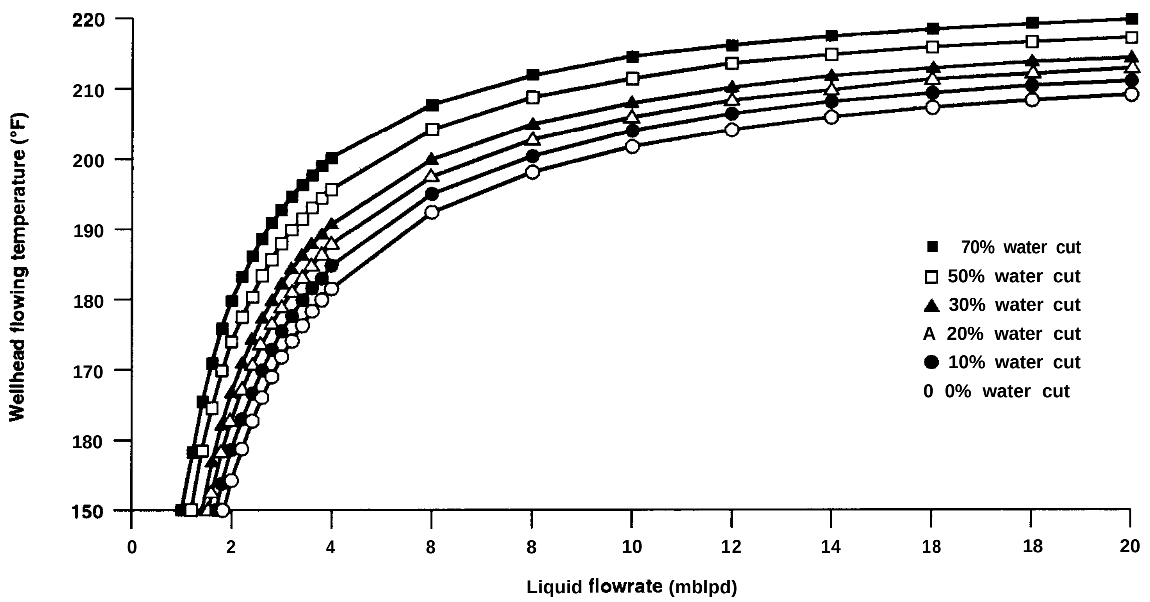

Figure 1. Relationship between WHFT and liquid flowrate and water cut

6.3.3 Effect of Gas Oil Ratio (GOR)

The gas phase in an oil well typically provides a relatively small part of the overall mass of the overall mass of the mixture. Therefore a higher GOR only makes a minor contribution to the overall enthalpy of the system. Since the rate of heat loss is dominated by the casings and the external medium, the increased velocity in the tubing has little effect and the heat loss remains reasonably constant. The effect of the gas oil ratio is less significant than the water cut or the flowrate.

MULTIFLO was used to predict the WHFT for a range of GORs from 100 to 2,000 scf/stb (with the same example as used previously). For a given flowrate, the WHFT only increased by roughly 2°F (1 °C) with increasing GOR. The range in WHFTs reduced with increasing flowrate and at 10 mblpd, the predicted WHFT only varied by 0.13°F between 100 and 2,000 scf/stb.

6.3.4 Effect of Lift Gas

Lift gas greatly complicates the heat transfer model. To rigorously model would require counter-flow calculations for the lift section.

For long well flowing times the overall heat transfer is dominated by the heat transfer through the earth. As a result the WHFT will be reasonably accurate without considering countercurrent flow; provided the mixture temperature of the gas lift gas and produced fluids is calculated. In this simplified approach the temperature profile takes place at the gas lift injection point.

The BP Multiphase program does not calculate the mixture temperature (it assumes gas temperature is the same as the production fluids). As a result MULTIFLO will overpredict WHFT for a well with high gas lift to liquid ratios if the gas lift temperature is significantly lower than the produced fluid temperature.

6.3.5 Reliance in Predicted Results

As mentioned above, the calculation of the WHFT has a major impact in the design of the topsides equipment such as the separator and the heat exchangers.

BP have performed a number of brief studies comparing actual WHFTs against predicted WHFTs for gas wells (West Sole & Bruce) and oil wells (Gyda & Ula). In general, the predicted results differed from the actual results by less than 5°F. In most cases, these calculations were performed using reasonably accurate measurements of flowrate, water cut, and bottom hole flowing pressure and temperature taken while testing the well.

In most cases, the WHFT will be calculated using sketchy data from the appraisal wells which may not be suitable for the whole development. Usually, the most unreliable piece of data is the reservoir temperature and the Bottom Hole Flowing Temperature (BHFT). This not only affects the initial enthalpy of the wells fluids but also affects the geothermal gradient which is usually assumed to be linear between the reservoir and the surface.

Care must be taken when predicting WHFTs for gas wells since, although MULTIFLO accounts for expansion cooling in the tubing, it cannot account for the cooling between the reservoir and the perforations of the tubing. This may lead to the prediction of a high WHFT.

The thermal conductivity of the surrounding rock also has a significant influence on the calculated WHFT. Usually only limited data on the rock type is available and the thermal conductivity is often assumed to be constant. It is advised that when calculating WHFTs, the sensitivity to reservoir temperature, flowrate, water cut, rock thermal conductivity are determined to provide a range of temperatures.

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6.4 Unsteady-State (WELLTEMP)

6.4.1 Time Dependency

Before start-up the rock surrounding the completion will be at (or close to) the geothermal gradient. As the well starts to produce, the surrounding rock will start to warm up. Initially, there is a high AT between the fluid and the surroundings and heat loss is high. With time, the heat loss will reduce as the rock warms up and the WHFT will increase. The rate of this increase is dependent on the flowrate.

Well start-up can be of particular importance. Before the well is started-up, the temperature of the fluids is at the geothermal gradient. On start-up, 'cold' fluid will be produced, if the fluids pass through a choke at the wellhead then the temperature downstream of the choke may be very low since the initial temperature upstream was low. This may result in hydrate formation with the potential for forming a blockage.

It is therefore important to know how quickly the well would warm up in order to determine the length of time that methanol/hydrate inhibitor injection would be required.

6.4.2 WELLTEMP

WELLTEMP is an unsteady-state single or two-phase simulation program. It can model many unsteady-state temperature aspects of wells such as drilling and mud circulation. It's primary use within the Multiphase Group of BP Exploration's Technical Provision (XTP) is the modelling of the start up of wells.

The program requires fairly detailed data on the completion, such as tubing and casing diameters, annular fills, cement details and profile (measured depth and true vertical depth). Detailed composition cannot be input and the program uses basic oil data (such as density, viscosity, and yield point) with compositional data for the light end.

Capabilities of WELLTEMP include:

- Offshore wells
- Deviated wells
- Multiple geothermal gradients
- Production
- Injection

Direct comparisons between MULTIFLO and WELLTEMP have been favourable. It is possible to approximate steady-state with WELLTEMP by specifying lengthy flowing times. Usually WELLTEMP predicts a WHFT l-2°F cooler than MULTIFLO. This could be due to either the simulation of slightly different cases or that the WELLTEMP simulation has not come to complete steady-state.

Major differences in the WHTP prediction of MULTIFLO and WELLTEMP have occurred when there is a large section of the well tubing/casing surrounded by sea water. In this section the rate of heat loss is high due to the sea temperature being low and the high heat transfer coefficient between the outside of the casing and the water. Small difference in the specification of the problem may lead to the prediction of different heat transfer characteristics and therefore different WHFTs. This difference between the results of the two programs is more pronounced with greater riser heights.

It is therefore important that this section of the well is accurately specified.

6.4.3 Start-up

When production commences from a well the WHFT will rise quite quickly. Usually, the well head temperature will approach the steady-state value within two residence times of the well. Typically this is no more than a few hours. The relationship between the flowing wellhead temperature and the flowrate during the start-up as shown in Figure 2.

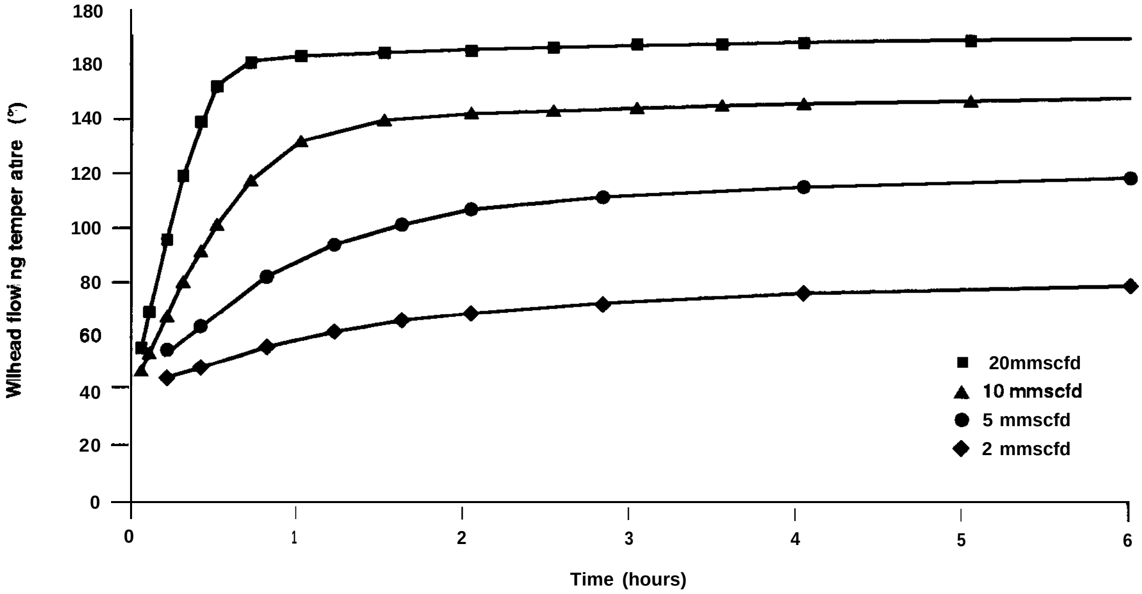


Figure 2. Well warm up rates, 4.5" tubing

6A Appendix: Setup of Well Using MULTIFLO

6A.1 Introduction

This appendix of the Wellhead Temperatures section of BP Multiphase Design Manual describes the method by which a well is modelled within Multiflow. It is not intended to be a exhaustive description of the method (please refer to the MULTIFLO Manual) but it does describe some problems involved in the calculation.

6A.2 Configuration

An oil or gas well can have many configurations (e.g. platform drilled, pre-drilled, subsea tiebacks etc). Most wells can be suitably described by MULTIFLO by combining the Tubing, Risers and/or Pipes segments.

For an offshore well, from the reservoir to the surface the well must be described as Tubing. From the seabed surface, the well can no longer be described as Tubing since the Tubing segments will not allow the external medium to be either air or water, therefore the well must be described as a Riser.

If the system is a subsea tieback, then Tubing should be specified from the reservoir to the wellhead, and then Pipes and Risers in the normal way back to the platform.

The production fluids enter the well through the section of the tubing that has been perforated. This section cannot be modelled using MULTIFLO, and so the 'inlet' or the beginning of the system is usually at the top of the perforated section or the bottom of the smallest casing.

6A.3 Tubing

MULTIFLO models the tubing as having only one casing. In practice, there can be a number of casings and/or conductors. MULTIFLO models the remaining casings, annuli and cement as equivalent thickness of cement (the calculation of the equivalent thickness of cement is described in Section A.7). The thickness of cement is entered in the Casing option of the Edit menu (shown below).

MULTIFLO Editor Dataset 53 (File 5) Forth Low Angle Well

Well Segment	Casing dimensions (in) Int diameter	Roughness	Thickness	Cement Thickness	Heat Xfer
1	8.68	0.0018	0.47	1.313	Y
2	8.68	0.0018	0.47	2.456	Y
3	8.68	0.0018	0.47	3.643	Y
4	8.68	0.0018 ... etc			

6A.4 Annulus Mediums

The annulus between the tubing and the casing can be filled with either Oil, Water, Cement, or Stagnant Gas.

Normally the annulus will be full of completion brine (i.e. Water). However, if gas lift is to be employed, then the annulus will be used to convey the gas to the gas injection point. The gas provides an insulating layer between the production tubing and the inner casing. The available options do not cover this explicitly, however the stagnant gas and brine options should roughly bracket the GLT case.

6A.5 Geothermal Temperature Profiles

It is unusual for a detailed geothermal gradient to be supplied with the completion data. Usually, the reservoir temperature and the sea temperature are supplied and a linear temperature profile is assumed. MULTIFLO will allow the 'top' and 'bottom' hole external medium temperature to be added and will calculate the intermediate temperatures.

In the example below (within the HEAT option), the inlet to the first section and the outlet of the last section have been entered.

			What?	Inlet	Outlet
1	TUBING	CALCULATED	ROCK	140	
2	TUBING	CALCULATED	ROCK		
3	TUBING	CALCULATED	ROCK		
4	TUBING	CALCULATED	ROCK		
5	TUBING	CALCULATED	ROCK		
6	TUBING	CALCULATED	ROCK		
7	TUBING	CALCULATED	ROCK		47.6

By exiting the screen (pressing F10 or Gold-E) MULTIFLO will fill in the 'missing' temperatures.

			What?	Inlet	Outlet
1	TUBING	CALCULATED	ROCK	140	136.841
2	TUBING	CALCULATED	ROCK	136.841	112.024
3	TUBING	CALCULATED	ROCK	112.024	110.502
4	TUBING	CALCULATED	ROCK	110.502	85.4298
5	TUBING	CALCULATED	ROCK	85.4298	-83.819
6	TUBING	CALCULATED	ROCK	-83.819	54.4032
7	TUBING	CALCULATED	ROCK	54.4032	-47.6

The temperatures now correspond to a linear geothermal gradient. The inlet and outlet temperatures are stored with the information for that segment, changing the co-ordinates of the system will not automatically change the inlet/outlet temperatures to the geothermal gradient. The temperatures have to be refreshed by repeating the procedure described above.

Rock thermal conductivities and diffusivities are given in Appendix G of the MULTIFLO Manual.

6A.6 Pressure Drop Correlations

6A.6.1 Crude Oil/Gas Wells

It is important when sizing well tubing to use a correlation which is appropriate for the particular system. Numerous validation exercises have been performed, but no one correlation has proved acceptable for all systems.

A recent extensive review of mechanistic models (performed by BP) has concluded that the Ansari method gave reasonable results over a wide range of conditions in oil and gas condensate wells, performing worst in high GLR conditions. In the absence of validation data for the field in question, Ansari is the recommended correlation for vertical pressure drop.

6A.6.2 Gas Condensate Wells

The Gray correlation is usually preferred. However, Gray will give inaccurate results if flow is other than single phase gas, annular or annular mist. Hence if modelling a system with a high condensate loading, or high produced water make, Gray should be avoided. The Ansari method generally gives good results over a range of possible liquid loadings.

6A.7 The TOOLKIT Option – Equivalent Thickness

6A.7. 1 Introduction

The TOOLKIT collection of utility programs includes COATINGS. This option allows the effects of multilayer coatings on heat transfer to be calculated as an equivalent single layer.

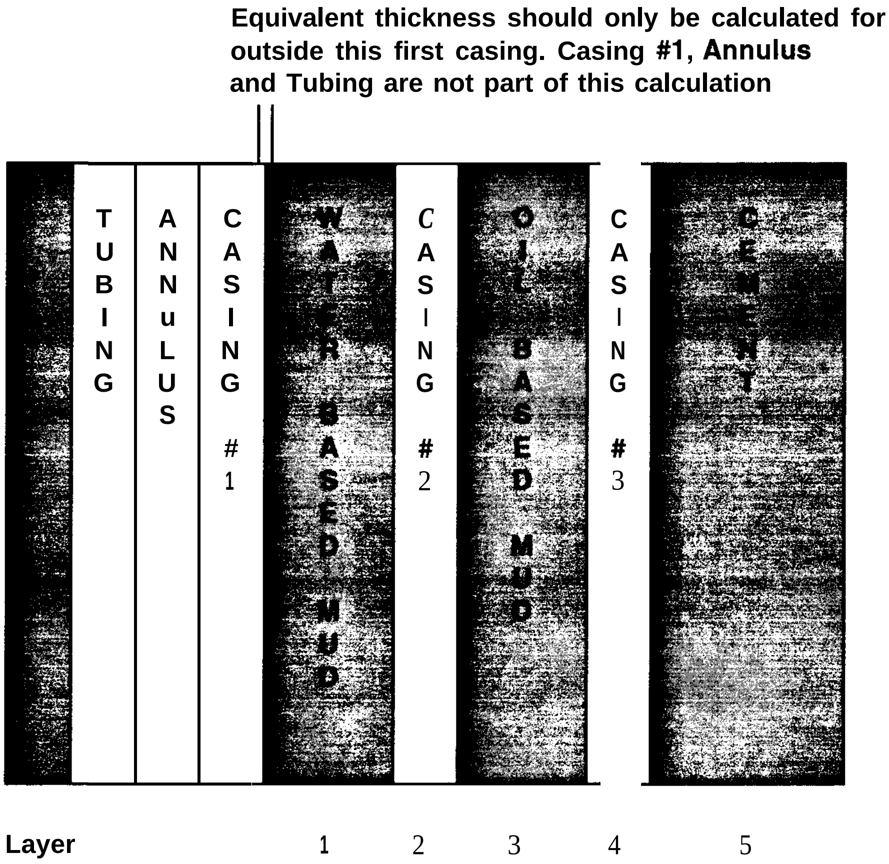
6A.7.2 Example Problem Set-up

The text below follows a worked example.

The diagram below shows a tubing with three casing strings. **The equivalent thickness should only be calculated for the layers outside the first casing.** There is a water based mud in one annulus and an oil based mud in the other, outside the third annulus there is a 2" thick layer of cement

Layer	i.d. (inches)	Conductivity (Btu/hr/ft/°F	Description
1	9.625	0.36	Water Based Mud
2	12.415	26	Casing #1
3	13.375	0.141	Oil Based Mud
4	16.5	26	Casing #2
5	18	0.5	Cement
o.d.	24		

Thermal conductivity of equivalent layer = 0.5 Btu/hr/ft/°F



A.7.3 Coatings

The coating calculation option can be accessed through the TOOLKIT utilities menu by simply typing COATINGS. The first screen appears as below:

```
                MULTI-LAYER COATING DATA

Number of layers of coating EXCLUDING the wall/casing [     5]
                     in the range [1:20]
```

In this case there are 5 layers outside the 9.625" casing. The first layer is the water based mud with a thermal conductivity of 0.36 Btu/hr/ft/°F. The data for the first layer can then be entered.

```
Collecting Data         for  Layer  [       1 ]
Inner Diameter          for  Layer  [   9.625 ]  in inches
Thermal Conductivity    for  Layer  [    0.36 ]  in Btu/hr/ft/°F
```

On pressing RETURN the inner diameter and the thermal conductivity of the first layer disappears and the program asks for the information for the second layer.

```
Collecting Data         for  Layer  [       2 ]
Inner Diameter          for  Layer  [  12.415 ]  in inches
Thermal Conductivity    for  Layer  [      26 ]  in Btu/hr/ft/°F
```

This is repeated until the inner diameters and thermal conductivities for all five layers have been entered. The program now needs information on the o.d. of the outer layer and the thermal conductivity of the layer (usually cement) to which the multilayer system is to be equivalent.

```
Outer Diameter of Outermost Layer [          24  1    in INCHES
Equivalent Conductivity of Single layer [   0.5  ]    in Btu/hr/ft/F
Inner Diameter of Flowline/Casing [      6.184  ]    in INCHES
```

The `'Inner diameter of Flowline/Casing'` is NOT part of the calculation but is reported in the results summary. All the information has now been entered and the program can now perform the calculation.

```
Flowline/Casing ID (in):                   6.184
Flowline/Casing OD (in):                   9.625
Equivalent OD (in):                       38.60
Equivalent thick (in):                    14.488

Flowline/Casing cond (Btu/hr/ft/°F):      26.00
Coating cond (Btu/hr/ft/°F):               0.50
```

To accurately model overall heat transfer, the equivalent OD should be reasonably accurate. Coating conductivity should be selected to provide correct OD.

The 5 layers have an equivalent thickness of 14.488" of a coating with a thermal conductivity of 0.5 Btu/hr/ft/°F. This can now be entered in the CASINGS option of the EDIT menu (see Section A.3).

Moving up the well, the annulus material changes and the number of casings increase. The calculation must be repeated whenever the annulus medium changes and/or another layer is added. A typical well may require 7-10 of these calculations to be performed. (It is advisable that calculations begin at the bottom of the well since the program stores information from the previous calculation and much of the information will for the subsequent calculation will be the same or similar).

A.7.4 Riser Section

As mentioned previously, the TUBING option in MULTIFLO will not allow air or water to be the external medium. Therefore these sections are described as RISERS.

Usually, if the well is a pre drilled well then the annuli will be filled with completion brine. The equivalent thickness of 'insulation' can be approximated to half of the difference between the o.d. of the outermost casing less the o.d. of the tubing. Small convection currents are set up within the various annuli containing the water, effectively increasing the thermal conductivity of the water. This has been estimated to increase the thermal conductivity by a factor of five to 1.8 Btu/hr/ft/°F. (0.36 x 5).

The increase in the thermal conductivity due to the convection currents **only applies to the completion brine in the riser section** where the temperature difference is the highest. The oil and the water based mud are more viscous and convection currents are not thought to be generated. Hence it is recommended that the thermal conductivities for the oil and water based muds quoted in Section A.8 are used.

The conductivity of the steel is high and so it's resistance to heat transfer can effectively be ignored. The calculated equivalent thickness of cement must be entered as a SPECIFIED layer in the sub menu of the HEAT transfer section shown below:

```
Segment  Coating  Characteristics

Standard  Coatings    USER SPECIFIED     KP-ENTER  to  change

Thickness             7.25   (ins)
K  Actual             1.8    (Btu/ft/hr/°F)
K  Effective          N/A    [??? means pipe diam not set]
```

If the annuli of the riser section consists only of either of these two types of mud then the thickness of the 'insulation' can be calculated as above with the thermal conductivity of the pure mud.

If the annuli are a combination of oil based mud, water based mud and/or completion brine, then a similar procedure will have to be adopted to calculate the equivalent thickness described in Section A.7.3 above. The section from the seabed to the platform has to be modelled as a RISER to allow specification of air and sea as the eternal medium. The calculation of the equivalent thickness of cement must now include the annulus and the first casing i.e. **the problem must be specified from the outside of the tubing.**

6A.8 Typical Thermal Conductivity Values

6A8.1 casings

The thermal conductivity of the casings and the tubing is high compared with the other mediums. Therefore the calculation of the equivalent thickness of cement is insensitive to this value. For the range of temperatures normally experienced with typical wells the thermal conductivity of steel is 26 Btu/hr/ft/°F.

6~8.2 Cement

The thermal conductivity of a typical cement is 0.5 Btu/hr/ft/°F.

6~8.3 Oil Based Mud

The thermal conductivity of the mud varies with density, different sources quote different values. a typical range is given below:

Density (s.g.)	Thermal Conductivity (Btu/hr/ft/°F)
1.2	0.200
2.0	0.343
1.08	0.220
1.26	0.245
1.51	0.285

6~8.4 H_2O Based Mud

As with oil based muds, the thermal conductivity varies with density. Typical values are given below:

Density (s.g.)	Thermal Conductivity (Btu/hr/ft/°F)
1 .0	0.36
1.2	0.393
2.0	0.489

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Nomenclature

A	Area available for heat transfer (ft^2)
C_{pm}	Fluid mixture specific heat (Btu/lb/F)
D_i	Inside diameter (ft)
D_o	Outside diameter (ft)
D_{ci}	Diameter to inside of concrete/coating (ft)
D_{co}	Diameter to outside of concrete (ft)
h_{comp}	Completion heat transfer coefficient ($Btu/hr^2/ft/°F$)
h_{earth}	External medium transfer coefficient ($Btu/hr^2/ft/°F$)
h_i	Internal film heat transfer coefficient ($Btu/hr^2/ft/°F$)
f(t)	Function of time, Equation 8 (-)
k_c	Cement thermal conductivity (Btu/hr/ft/°F)
k_e	Soil thermal conductivity (Btu/hr/ft/°F)
k_j	Layer thermal conductivity (Btu/hr/ft/°F)
k_m	Fluid mixture thermal conductivity (Btu/hr/ft/°F)
M	Mass flowrate (lb/hr)
T_{in}	Inlet temperature (°F)
T_{out}	Outlet temperature (°F)
Q	Heat duty (Btu/hr)
z	Dimensionless term in Equations 8 and 9 (-)
α	Thermal diffusivity (ft^2/hr)
ΔT	Temperature difference (“F)
ΔT_{lm}	Log mean Temperature difference (“F)
λ_L	Non-slip hold-up (-)
μ_g	Vapour viscosity ([ft^2/hr)
μ_m	Fluid mixture viscosity (ft^2/hr)
μ_l	Liquid viscosity (ft^2/hr)
ρ_m	Fluid mixture density (lb/ft^3)
Re	Reynolds Number
Pr	Prandtl Number
U	Overall heat transfer coefficient ($Btu/hr^2/ft/°F$)

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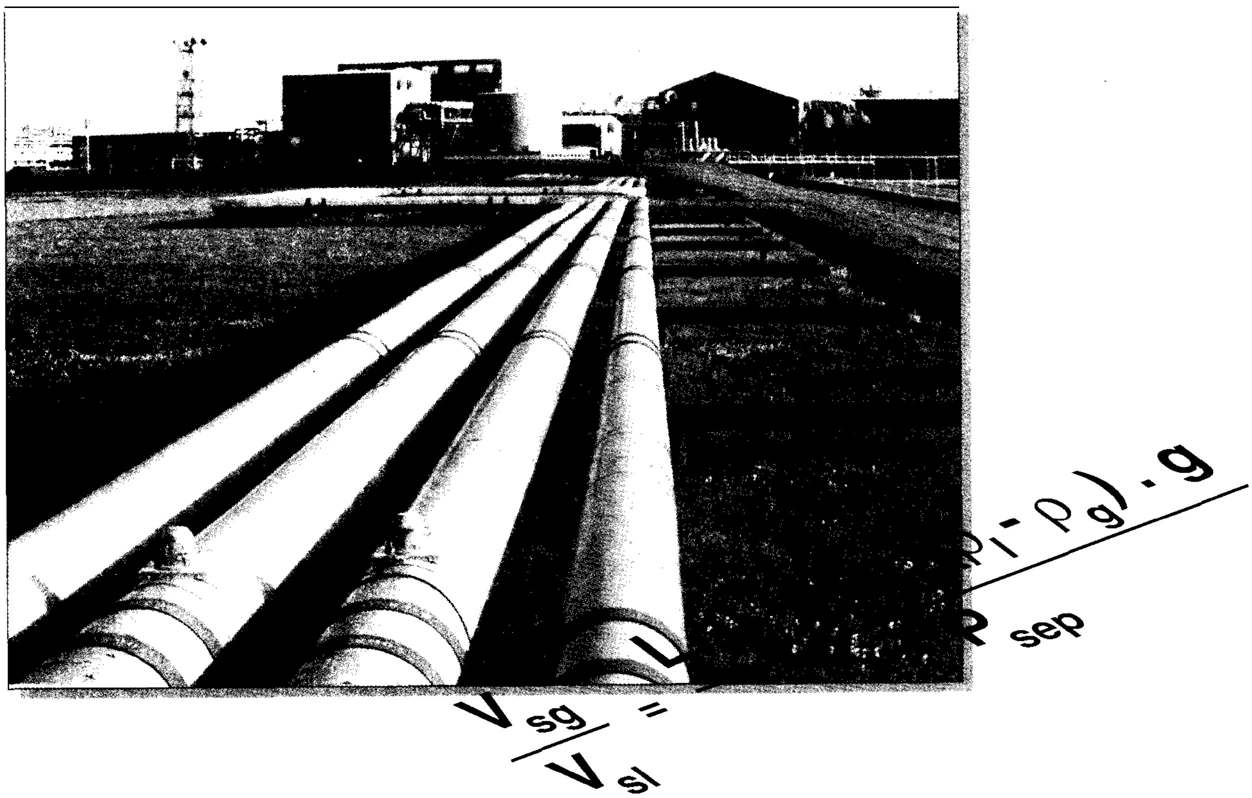
Section 7 Transient Flow



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7.1 Introduction

Steady state operation is the exception rather than the rule in multiphase systems due to a number of factors such as the nature of the flow regimes, interactions with the flowline profile, and changes in supply or demand. As a result several types and speed of transient need to be considered in order to size separation equipment, to determine operational procedures to achieve maximum cumulative production, to ensure safe operation, and to assess the impact of pipeline ruptures, for example.

This section of the Multiphase Design Manual aims to describe where transient multiphase flows occur and what design tools are available to the engineer. A number of sections are devoted to the different types of transients which have been categorised as follows:

- **Transients caused by flowrate changes**
- **Transients resulting from pressure changes**
- **Transients created by topographical effects**
- **Pigging transients**
- **Pressure surge transients**

Transients caused by pipeline ruptures are covered separately in Section 9 of this Design Manual.

Over the past few years as dynamic simulators have been developed, their use has become widespread, and central to the determination of efficient operating guidelines. This has been driven by costly lost production due to unforeseen transient flows causing plant shutdowns (such problems may not have been expected from steady state analysis), and the increased concern for the safety and environmental impacts of pipeline failures, which often results in transient two-phase flow, even though the normal pipeline operation may be single-phase. As transient simulations have become available they have also found use in other aspects of pipeline operation including corrosion assessment and pipeline leak detection.

In parallel with computer code developments has been the increasing complexity of multiphase production systems. Such schemes have generated the need for greater sophistication in multiphase simulation and have been part of the driving force behind the development of the codes available today. In addition the rapid advance in computer technology has meant that expensive and time consuming calculations that were previously run on large company main frame machines are now efficiently run on desk top machines and even portable personal computers. This has greatly improved the access to transient simulators and significantly reduced the cost of running the codes.

This chapter of the manual is designed to show the reader the types of transient two-phase flow simulations that can be carried out by current technology. The aim is not to swamp the reader with the mathematics of the codes but to show by examples some of the work undertaken in the past. Simple hand methods are outlined where possible, however the complexity of transient two-phase flow analysis usually requires computer solutions provided by specialist codes.

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7.2 Transient Codes in Use

Although there are a large number of engineers and scientists working in the area of transient two-phase flow modelling, the number of commercial software packages is limited to just a few. The most widely used is the OLGA code developed by IFE and SINTEF in Norway. OLGA was the first commercial general transient two-phase flow package, which has enjoyed wide use by the 6 members of the consortium that sponsored the development. The early versions of the code required considerable expertise to create input data and to interpret the results. As a result many of the oil companies using the code found it difficult to use and very costly. Some of them have developed their own versions of OLGA for their specific needs.

In competition to OLGA is the PLAC code developed by Harwell laboratories which was only commercialised in the autumn of 1992, but had been used for consultancy for more than two years previous. PLAC was developed with support from BP and is the transient simulator most widely used in-house by BPX.

There are several other transient two-phase flow codes presently under development. Total are working on a transient version of the mechanistic PEPITE program called TACITE, which will feature a slug tracking model. Shell have a code called TRAFLOW which is under development as an on-line simulator and analysis code.

There is a great deal of activity in the development of transient programs. However most of this is being carried out in-house by oil companies and research centres, hence little is known about them. For this reason the discussion in this Section will be limited to the codes that BPX have had direct experience with. It is however expected that competition in the market place will increase when some of the other codes are commercialised.

A description of the BP developed BPREAKL code for estimating release rates from volatile oil pipelines is given in Section 9 of the Multiphase Design Manual.

OLGA

OLGA is a dynamic, one dimensional modified two fluid model for transient two-phase hydrocarbon flow in pipelines and networks, in which processing equipment can be included. The code has been developed by the SINTEF/IFE two-phase flow project which commenced in 1984 and was based on the computer program OLGA 83. This was originally developed for Statoil by IFE in 1983.

The two-phase flow project was aimed at improving OLGA by expanding an experimental database from a high pressure 8" large scale test facility run by SINTEF at Tiller in Norway. Extensive testing was carried out by IFE and by the projects member oil companies which included Conoco Norway, Esso Norge, Mobil Exploration Norway, Norsk Hydro, Petro-Canada, Saga Petroleum, Statoil, and Texaco Exploration Norway.

The basic models in OLGA contain three separate mass conservation equations for the gas, the continuous liquid and the liquid droplets, these are coupled with the interphase mass transfer terms. Momentum conservation is applied to the gas-droplet field and to the continuous liquid, hence giving two-equations. The mixture energy equation is written in conservation form, accounting for total energy balance in the system. OLGA predicts as a function of time the pressure, temperature, mass flow of gas and liquid, the holdup and the flow pattern. Closure laws are required for the friction factor and the wetted perimeters of the phases and these are flow regime dependant. The droplet field also requires a entrainment and deposition model. The two basic flow regimes adopted are distributed and separated flow. The former contains bubble and slug flow and the latter stratified and annular flow. The transition between the two regime classes are determined according to a minimum slip concept.

Due to the numerical solution scheme the original versions of OLGA are particularly well suited to simulate slow mass flow transients. The implicit time integration applied allows for long time steps which is important for the simulation of very long transport lines, where typical simulation times are in the range of hours to days.

The necessary fluid properties (gas/liquid mass fraction, densities, viscosities, enthalpies etc.) are assumed to be functions of temperature and pressure only, and have to be supplied by the user as tables in a specific input file. Thus, the total composition of the two-phase mixture is assumed to be constant both in time and along the pipeline for a given branch. The user may specify a different fluid property table for each branch, but has to ensure realistic fluid compositions if several branches merge into one.

In 1989 Scandpower A/S acquired marketing rights to the OLGA program and commercial use of the code increased considerably. The early versions of OLGA had a simple data input and output file format which made the setting-up of problems very time consuming. Later versions have user friendly interfaces which greatly enhances the use of the code, PreOLGA is used for input data generation and PostOLGA is used for post processing. Fluid properties are generated using a package called PVTOL.

In 1993 Statoil acquired the Tiller facility for 5 years and embarked on a substantial R and D effort to improve OLGA, areas under development include:

- **Improved mechanistic modelling and closure laws**
- **Slug flow modelling**
- **Three phase flows**
- **Compositional tracking**

These areas also indicate where there are deficiencies in the present code.

In addition, OLGA has been interfaced with the Fantoft D-SPICE multi-compositional dynamic process plant simulator to improve the modelling of equipment, however this interface only provides a dynamic exchange of data between the codes and is hence not a fully integrated model.

Some of the other original two-phase flow project sponsors have also developed OLGA for their own use, notably Conoco who have developed improved thermal modelling and pigging simulation in their version called CONOLGA.

PLAC

The PipeLine Analysis Code (PLAC) was originally developed from the nuclear reactor safety code TRAC (1986) under a four year programme sponsored by BP Exploration and the UK Offshore Supplies Office. The major parts of that work were to remove the redundant reactor specific components and to convert from steam/water physical properties to multi-component hydrocarbon properties. More suitable models for interfacial friction and flow regimes were also introduced. The first release of PLAC was available for use by AEA Petroleum Services and BP Engineering in mid 1990. Commercialisation of the code was not realised until the autumn of 1992 however BP and AEA conducted numerous validation tests on the code with good and bad experiences. PLAC is the transient simulator used most widely in-house by XFE to model transient two-phase pipeline flows. Like OLGA the original versions of the code were cumbersome to use and the ease of use has been greatly improved by the development of interactive pre and post processors.

PLAC solves mass, momentum and energy equations for each phase using a one-dimensional finite difference scheme. Six equations are solved, gas and liquid mass and momentum conservation, total energy conservation, and gas energy conservation. Unlike OLGA, PLAC does not have a separate equation incorporating a droplet field in the gas stream. Appropriate flow pattern maps and constitutive relationships are provided for wall and inter-facial friction and heat transfer, and a model for multi-component phase change is included. PLAC has flow regime maps for vertical and horizontal pipes and switches to vertical flow if the angle of inclination is above 10 degrees. The horizontal flow pattern map is based on the method of Taitel and Dukler (1976).

The fluid physical properties are calculated from a user supplied mixture composition using an internal PVT package, however this generates a table of properties similar to OLGA, and hence still relies on simple equilibrium phase behaviour predictions. PLAC can only use one composition in a network. The current version of the code has the capability to handle PIPES, TEES and VALVES and is able to predict a range of phenomena including flowrate and/or thermal transients, severe slugging, shutdown/restart problems and pipeline depressurisation. The boundary conditions are specified by FILL and BREAK components.

TRANFLO

TRANFLO is a three phase flow transient program developed by the mathematical sciences group at BP Research, Sunbury. The program is based on the one-dimensional averaged equations of motion for the conservation of mass and momentum for each phase and does not include the effects of gas entrainment. The equations for each phase are transformed into evolution and algebraic equations. The resulting system of equations is analogous to that for motion of liquid down a open, inclined channel. The program can generate waves on the interface due to physical instabilities. These evolve non-linearly into a train of periodic waves whose fronts propagate as shocks or hydraulic jumps. The waves can touch the top of the pipe leading to the formation of slug flow.

TRANFLO was under development at Sunbury on the VAX computer and showed some promise in the simulation of sloshing in dips, slug tracking, and water puddling in pipelines. The program development ceased in 1993 and was handed over to AEA Petroleum Services, Harwell, where its future is unclear, however it may be incorporated into PLAC or ported onto a PC. The version at BP is limited to isothermal cases without phase change, and is also poor for vertical flow. It is an R&D tool and is not suitable in its present form for modelling complete pipeline systems, but may be useful for investigating three phase effects in simple specific geometries. The three phase version of PLAC is expected to make TRANFLO obsolete in the future.

SPEEDUP

SPEEDUP is a general purpose simulation package capable of solving coupled, ordinary differential equations. It is used within BP to examine transient conditions in process facilities such as advanced control of reactors and distillation columns, the interaction of slugging oil and gas production lines with facilities, compressor control and surge, stream fuel gas and relief systems etc..

SPEEDUP is also capable of addressing a wider range of problems. It can perform steady state calculations, dynamic simulation, optimisation, parameter estimation and data reconciliation. It is also a flowsheeting package and data is organised in terms of streams of connecting units. The units in turn are models taken from commercial or user generated libraries.

The program is generally used by the dynamic simulation group to investigate the response of process plant to multiphase pipeline transients such as severe slugging and normal slugging. This is handled by a Dynamic Slugging Model that has been incorporated as a module within SPEEDUP and allows and integrated model of the pipeline and facilities to be used to assess the effect of slugging on the entire process plant. The model simulates the deceleration of the slug as it enters the riser (due to the increase in the hydrostatic head) and the subsequent acceleration of the slug as it exits the riser and enters the topsides facilities. The slug production phase is followed by a period of gas blowdown as the pressured gas bubble behind the slug depressures into the facilities. The Dynamic Slugging Model assumes a two-phase system (liquid and gas) and requires the size and frequency of the slugs to be input. It does not allow slugs to break-up or to amalgamate.

WELLTEMP

WELLTEMP was designed to examine temperature transients caused by flow in wells. It can also be used to examine temperatures in pipelines. It handles heat flow through tubing, casings, annuli, cement and formation. various operations can be modelled such as injection, production, forward and reverse circulation, shut-in and cementing. It assumes steady state flowrates; although different flow periods can be entered (steady within each period). It can handle dual or even triple completions (and co-bundled flowlines). In dual completions it would allow injection down one string and production back up the other.

It can be used to examine surface and downhole tubing temperatures during drilling or simulated treatments. This is useful when considering fluid properties, cement characteristics etc. It can also help when considering corrosion, wax deposition or hydrate formation. It can be used to examine cool down times in pipelines. It could also be used to refine the temperature profile in the annulus and wellbore during gas lift operations.

WELLTEMP is marketed by Enertech Engineering and Research Co of Houston and has been primarily used by XFE to investigate the warm-up of wells and flowlines. WELLTEMP can usually run much quicker than PLAC and has been used to investigate sensitivities before performing a detailed PLAC simulation. The simple composition treatment in WELLTEMP is a limitation to detailed transient two-phase flow analysis.

University of Tulsa Codes

The University of Tulsa through the TUFFP JIP have developed a number of transient codes over the years which have been made available to BP. One of the earliest codes was produced by M W Scoggins in 1977 and was based on a one-dimensional two-phase flow model with slip. The model is based on the existing steady state Eaton method for holdup and the Dukler two-phase friction model. The treatment of phase behaviour and physical properties is based on hard wired black oil correlations and the simulation is isothermal only. This program has been ported onto a PC by XFE and is called SCOGGINS, however the code suffers from numerical problems and its inherent limitations mean that it is little used.

The latest transient analysis code from TUFFP was developed by Minami in 1991 who implemented a simplified model proposed by Taitel. He also developed a new pigging model that was coupled to the transient model. The sets of equations are discretized using a lagrangean grid system for the pigging model. The resulting set of equations are solved with a semi-implicit scheme and an approach based on the stability of the slug flow structure is used to predict the flow pattern transition boundaries.

The version of this code released to BP is only capable of simulating conditions in the TUFFP 1400 feet long 3" test facility and is hence of little use. Comparison of the predictions with recent data from transient tests on the Tulsa facility indicate that the code is inferior to PLAC.

FLOWMASTER

Flowmaster is a computer program designed primarily for single phase fluid flow analysis and pipe network design, and has various analysis options including steady state hydraulic analysis, flow balancing, and pressure surge analysis. The network is constructed from a pallet of components and the design data input for each item. XFE primarily use FLOWMASTER for single phase pressure surge analysis, however there are a number of features which can facilitate a simple surge analysis for two-phase flows including a bubbly pipe component, which allows the speed of sound to be modified to account for a homogeneous gas fraction, and a primer component, which can simulate liquid filling a gas space.

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7.3 Transients due to flowrate changes

The operation of oil and gas pipelines often involves changing the inlet or outlet flowrates for a number of reasons, namely:

- Start-up and shut-down.
- Fluctuating supply and demand, i.e gas delivery changes.
- Switching wells on and off for maintenance or testing.

Such flowrate changes can have a considerable effect on the downstream processing plant as flowrate increases can give rise to high transient liquid production rates and potential gas surges, and reductions in flow can result in periods of low flow. This is illustrated by the following example.

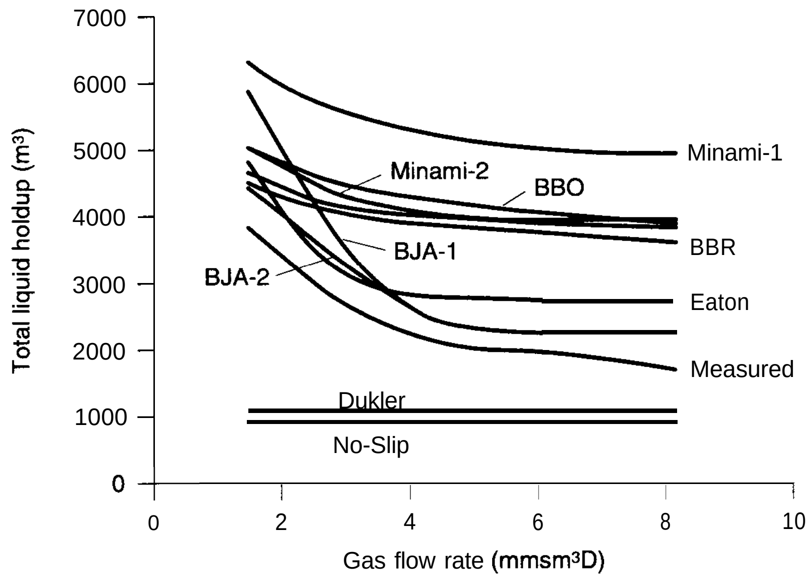


Figure 7.1 Marlin pipeline holdup profile

Figure 7.1 shows the predicted liquid content of the Marlin gas condensate pipeline as a function of the gas flowrate. This is a 67 mile long, 20" diameter wet gas line operating with a liquid loading of around 65 bbls/mmscf. It is seen that the general trend is that the liquid content reduces as the gas flowrate increases. Hence, if a gas flowrate increase is made, liquid will be removed from the pipeline as the new equilibrium liquid content is established. If the change in the gas flowrate is carried-out too fast, the excess liquid can be swept out as a large 'slug' which may overfill the downstream plant, whereas if the changes are made slowly, the liquid can be gradually swept out within the slug catcher capacity. It is also interesting to note in this Figure the wide variation in the predicted holdup obtained from using the various methods. This may not be too much of a problem when calculating the amount of liquid swept out during a transient because it is the difference in the holdup that it most important. In this example the Eaton holdup correlation is expected to give reasonable answers. However, one should be wary of using the correlation approach rather than mechanistic models since they do not usually predict the steep rise in holdup as the velocity and inter-facial friction reduces – hence a gross underestimate can result when starting from low flowrates.

Figure 7.2 shows the liquid flowrate at the outlet of the Marlin gas condensate pipeline during an increase in the flowrate from 155 mmscfd to 258 mmscfd. During the test the gas rate was held constant at 155 mmscfd for 52 hours in order to reach equilibrium conditions. The rate was then increased to 258 mmscfd in a period of one hour and held constant for a further 26 hours to obtain equilibrium conditions again. It is seen that during the transient the outlet liquid flowrate is considerably higher than the final equilibrium value.

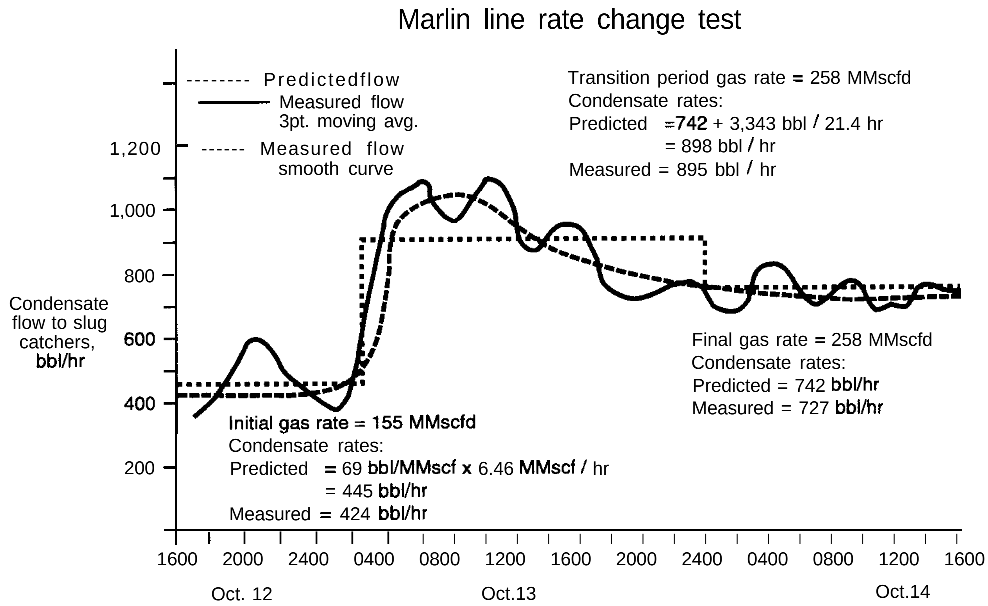


Figure 7.2 Marlin rate change test

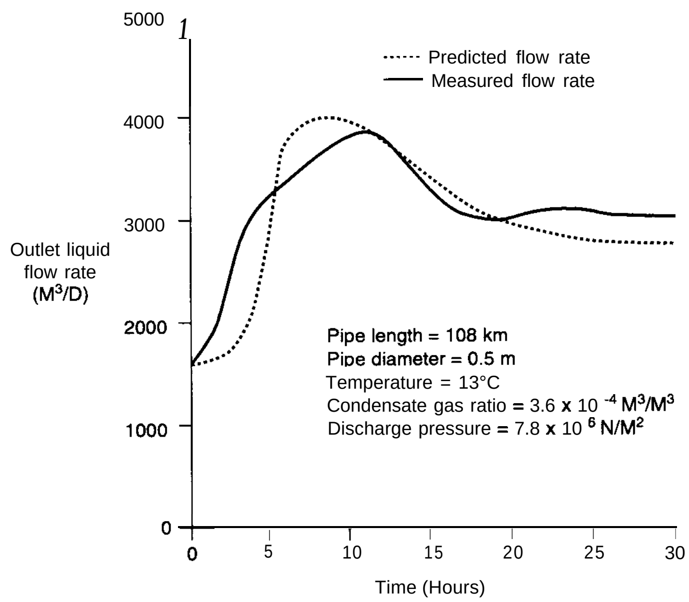


Figure 7.3 Marlin comparison with SCOGGINS

In the next section we shall outline a simple method for estimating the slug catcher size required to handle the liquid removed during flowrate increases. Such a method was used to determine the predicted flowrate profile shown in Figure 7.2. In later sections the use of transient computer codes will be outlined. However, Figures 7.3 to 7.5 are provided here to illustrate the potential accuracy of transient codes on the Marlin rate change data. This data has often been used as a test case as it is one of the few transient field data sources available in the the open literature.

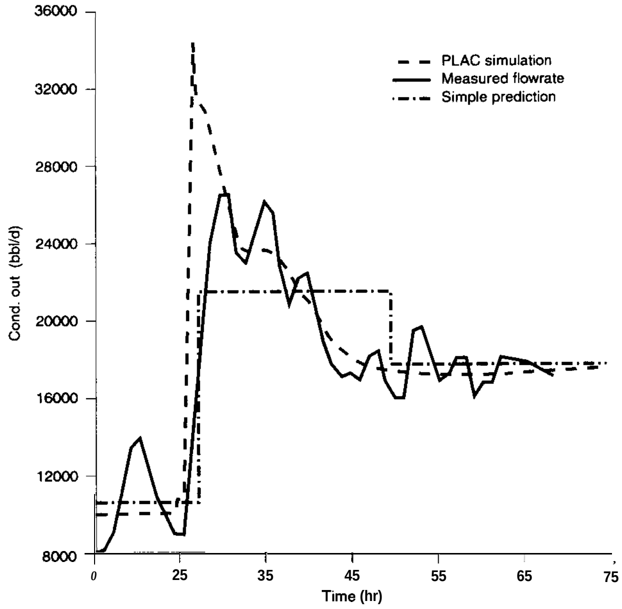


Figure 7.4 Marlin comparison with PLAC

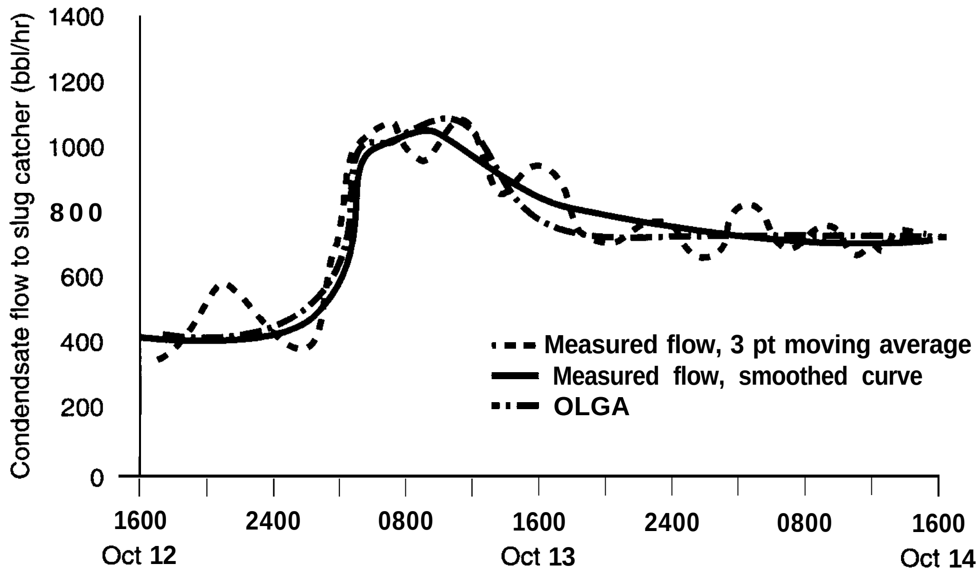


Figure 7.5 Marlin comparison with OLGA

Pigging gas condensate pipelines can also result in large slugs. However, in many cases it is not feasible to design slug catchers of a sufficient size to handle the equilibrium slug produced by running pigs at low throughputs. In this case pigs must be run frequently to prevent the liquid holdup reaching equilibrium, which can incur high operating costs. Sometimes gas rates can be reduced during the pig arrival to allow the produced liquids to be processed, but in others the gas flowrate may be determined by the consumer, and the pigging operation carried out at the prevailing gas rate. In some situations pigs are only required on an infrequent basis for corrosion control or inspection, in which case the pipeline liquid content may be high and a procedure must be put in place to handle the liquid swept-out by the pig. One way of doing this is to stop the pig offshore and 'walk it' into the slug catcher at a rate compatible with the liquid processing capacity. This approach was successfully carried out at the restart of pigging operations on the Amethyst pipeline where a liquid volume of over four times the slug catcher capacity was allowed to accumulate in the sealine when the offshore pig launcher failed. In some instances it is possible to reduce the liquid content of the pipeline prior to pigging by controlled rate increases to remove liquid. In other cases pigging is not possible, and flowrate changes must be controlled to prevent overfilling the downstream plant.

With these factors in mind it is seen that for gas condensate systems at least, it is often the transient 'slug' that determines the slug catcher volume. For oil and gas pipelines pigging may be required frequently for wax control etc, where the lines have reached equilibrium, hence this may determine the size of the slugcatcher. However for some developments, particularly subsea, pigging is required less frequently and can be accomplished with some operating ingenuity. Here it is often the longest normal slug or the transient rate change liquid sweep out that determines the required slug catcher surge volume. The next section outlines a simple calculation method for estimating the liquid outflow profile due to rate change transients.

7.3.1 Simple analysis methods for flowrate increases and reductions

The first step in considering the required surge volume for transient flowrate increases can be made by considering the equilibrium holdup/flowrate profile.

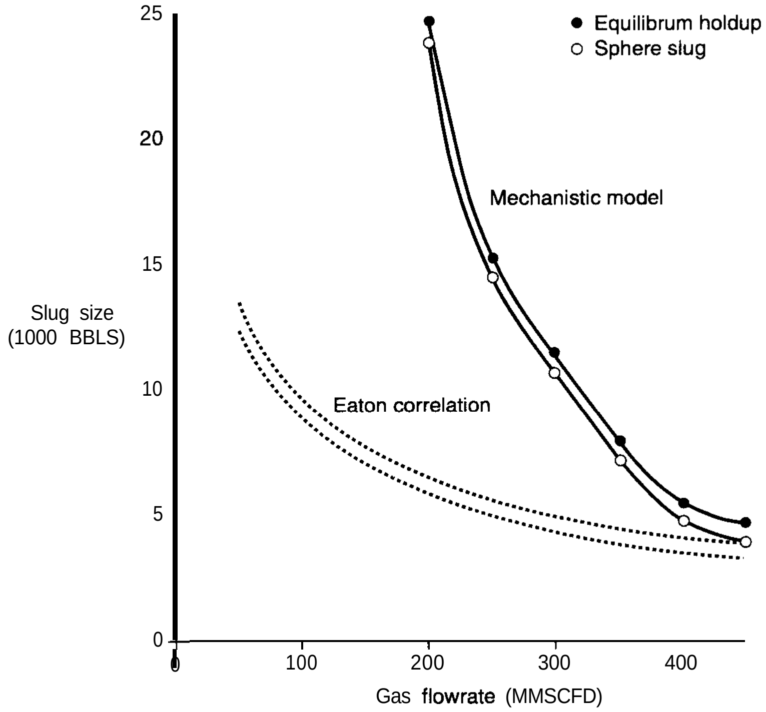


Figure 7.6 Amethyst pipeline holdup profile

Figure 7.6 shows the equilibrium holdup for the Amethyst pipeline in the Southern North Sea. The line is 30" diameter and connects two unmanned production platforms with the Easington gas processing terminal 30 miles away. The working volume of the slug catcher is 3800 bbls, with a maximum liquids pumpout rate of 6000 bbl/d. The liquid loading is typically 6.5 bbl/mmscf of condensate and 0.54 bbl/mmscf of water and methanol. It can be seen that for normal **gas** flowrates in the region of 250–350 mmscfd the equilibrium liquid holdup is well in excess of the slugcatcher capacity, hence pigging at equilibrium conditions could overfill the process facilities. Higher gas rates than those available would be required to give equilibrium holdups below the surge volume. In fact this pipeline was designed to be pigged daily, and hence the liquid expected to be received was the order of 1760 bbls at 250 mmscf/d. This design allows for a missed pig, ie two days between sphere launches, without causing liquid handling problems.

The holdup profile shows a rapid rise in the liquid content at gas rates below 200 mmscfd, hence operating in this region could cause problems even without pigging, as relatively small gas rate perturbations can cause large quantities of liquid to be swept from the pipeline. For example a 10% increase in the gas rate from 200 to 220 mmscfd could remove 4750 bbls of liquid and, possibly, swamp the slug catcher.

Shortly after the start-up of the Amethyst pipeline system the pig launcher failed and it was decided to investigate the consequences of continuing production, and allowing the pipeline liquids to build up to equilibrium values. A decision was taken to limit the minimum gas flowrate to 250 mmscfd, and hence to avoid possible uncontrolled sweep-out due to inherent flowrate fluctuations. The pipeline was operated without pigging over the winter, where it took several months to obtain an equilibrium holdup, which was estimated to be in the region of 18000 bbls from a mass balance. This is within 17% of the value predicted by the old segregated flow mechanistic model in MULTIFLO, but is 213% higher than the value predicted by the Eaton correlation. Some of the simple transient analysis outlined below was used to investigate how to resume pigging operations.

A simple approach to sizing a vessel to handle the liquid produced by a gas rate increase would be to consider the change in the equilibrium holdup and ignore the effect of the liquid pump-out rate. For example if the gas rate were increased from 250 mmscfd to 350 mmscfd the equilibrium liquid removed would be 15373 - 8018 = 7355 bbls. Hence the gas rate could be increased in two steps from 250 mmscfd to 300 mmscfd which would remove 3800 bbls and then 300 mmscfd to 350 mmscfd which would remove 3555 bbls. A 7355 bbl surge volume would be required if the increase were made in one step without taking account of the liquid pump-out rate. The maximum gas flowrate available is 350 mmscfd which gives a pipeline inventory of 8018 bbls, and hence it is still necessary to 'walk in' the first pig.

We will use the Amethyst case to illustrate a simple way of determining the effect of the pump-out rate on the required surge volume. Consider the case of an increase in the gas rate from 250 to 350 mmscfd.

At 250 mmscfd the equilibrium holdup is 15373 bbls

At 350 mmscfd the equilibrium holdup is 8018 bbls

Hence the difference in holdup is 15373 - 8018 = 7355 bbls.

Next calculate the initial and final liquid production rates from:

Liquid rate = liquid loading x gas flowrate

hence:

Initial liquid flowrate = 250 mmscfd x 7.04 bbl/mmscf = 1760 bbl/d

Final liquid flowrate = 350 mmscfd x 7.04 bbl/mmscf = 2464 bbl/d

Calculate the duration of the transition time for the transient, which is the length of time over which the high flowrate occurs. If it is assumed that all the liquid in the line accelerates to the equilibrium liquid velocity corresponding to the final gas rate, then the transition time is the same as the residence time at the final rate, ie:

Transition time = (final holdup / final flowrate) = (8018 / 2464) = 3.25 days

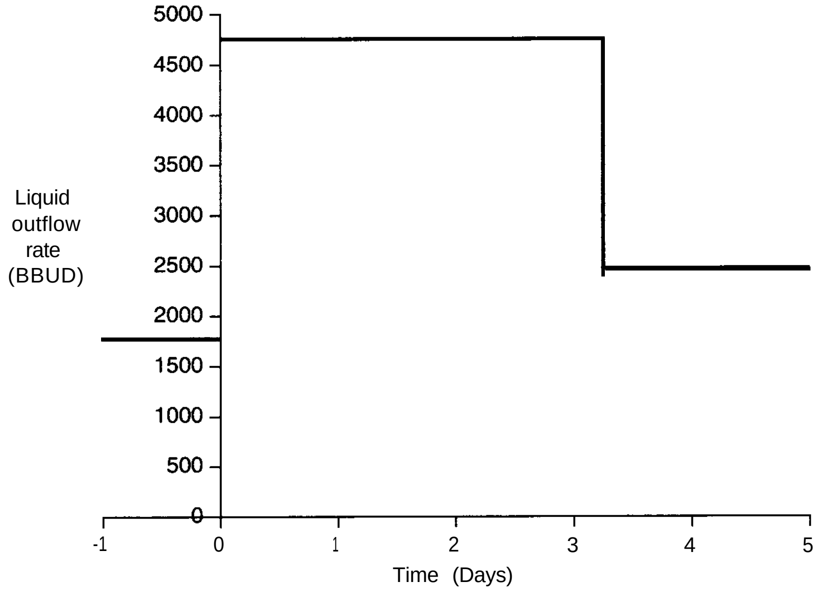
The transition flowrate is the sum of the final flowrate and the increase due to the rate change and is given by:

Transition flowrate = Final flowrate + (holdup change /transition time)

= 2464 + 7355/3.25 = 4727 bbl/d

It is seen that based on this method the surge can easily be handled by using the 6000 bbl/d pump-out capacity of the Easington terminal. Figure 7.7 shows the predicted liquid outflow profile.

Figure 7.7. Simplified liquid outflow profile during ramp-up



This approach can be extended to investigate the trade-off between the required slug catcher surge volume and the pump-out rate using the relation below:

Surge volume = transition time x (flowrate in - flowrate out)

$$= Tt \times (Q_{in} - Q_{out})$$

If the pump-out rate is fixed at the final equilibrium value the required surge volume is:

$$V_s = 3.25\ (\ 4727 - 2464\ ) = 7355 \text{ bbls}$$

ie. the change in equilibrium holdup.

If the pump-out rate is 4727 bbl/d then this method shows that no surge volume is required. The solution to the equation is the linear relationship shown in Figure 7.8. It can be seen that for a surge volume of 3800 bbls a minimum pump-out rate of 3550 bbl/d is required.

The relationship shown in Figure 7.8 can be subject to large inaccuracies at the extremes as the flowrate tends to zero and at the final transition flowrate, the reason being as follows; at zero pump-out rate the surge volume is equal to the transition flowrate multiplied by the transition time, whereas in practice liquid continues to flow into the vessel at the final equilibrium flowrate, hence the required surge volume becomes infinite as the pump-out rate goes to zero. When the pump-out rate is equal to the transition flowrate the above method indicates that no surge volume is required. However in practice the flowrate during the transient is not usually constant, and typically peaks at the start. Hence the solution for a surge volume of zero is a pumpout rate equal to the peak flowrate during the transient.

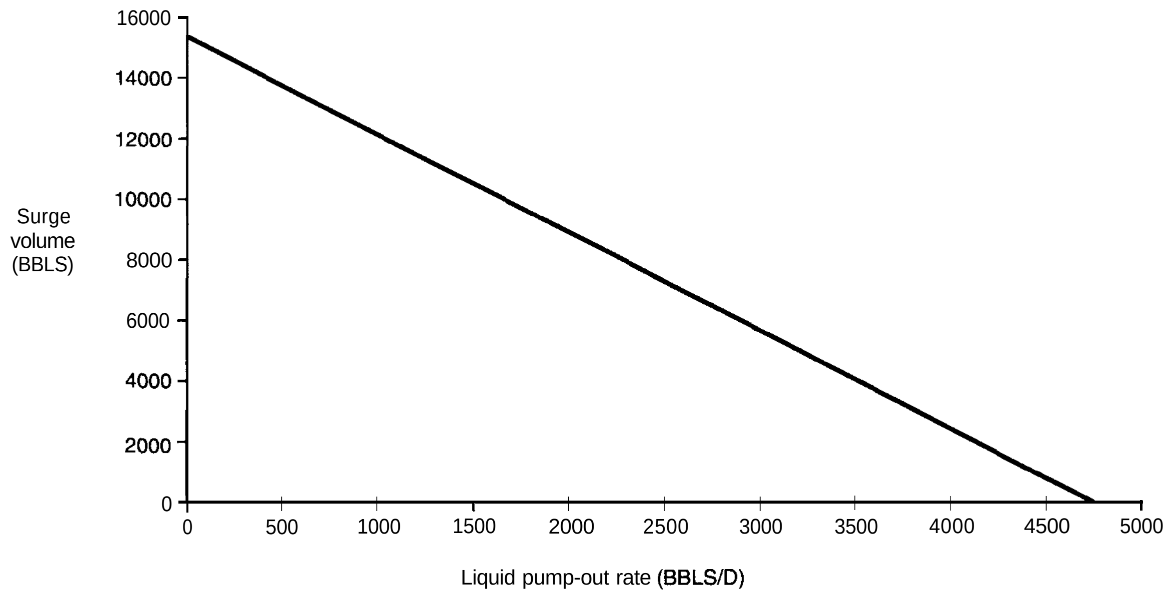


Figure 7.8 Surge volume as a function of processing rate

The same simple method can also be applied to the estimation of the outlet flowrate profile during a flowrate decrease, as follows: Consider a reduction in the gas flowrate of the Amethyst pipeline from 350 mmscfd to 250 mmscfd.

First calculate the residence time at the final flowrate, this is also assumed to be the transition time:

Transition time = (final holdup / final flowrate) = (15373 / 1760) = 8.73 days

then:

Transition flowrate = Final flowrate – (Holdup change / Transition time)

= 1760 – (7355 / 8.73) = 918 bbl/d

Hence the hand calculation method predicts an initial fiowrate of 2464 bbl/d falling to 918 bbl/d over a 8.73 day transition period after which the rate increases to 1760 bbl/d. This is illustrated in Figure 7.9.

Example 7.1 illustrates a PLAC simulation of the shutdown and restart of Pompano subsea wells and Example 7.2 gives a comparison with the simple hand calculation method.

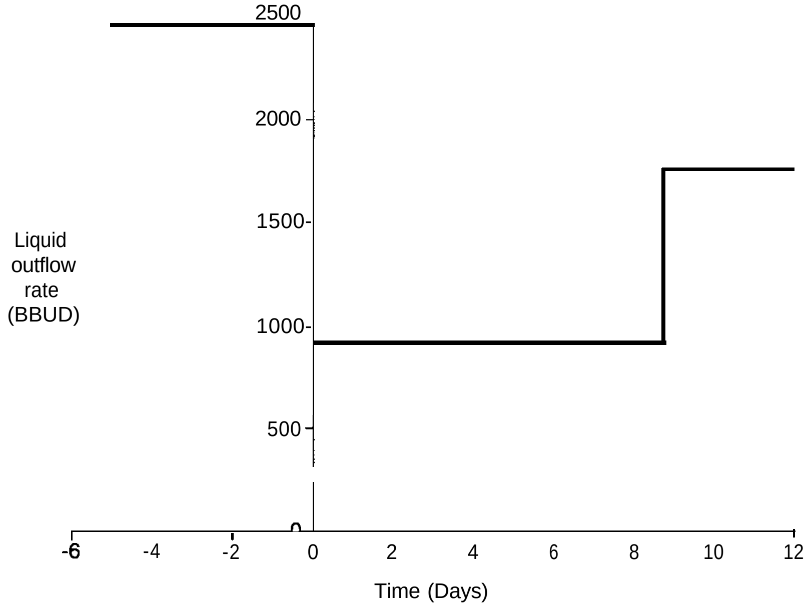


Figure 7.9 Simplified liquid outflow profile during ramp-down

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7.3.2 Use of transient computer codes for other boundary condition changes

There are other changes in the boundary conditions of a pipeline apart from flowrate changes, for example outlet pressure changes, which are often required to be simulated by transient analysis. With general transient simulation programs such as PLAC there is a wide variety of boundary conditions that can be changed. For example PLAC uses the FILL component to specify a flowrate boundary and the BREAK component to specify a pressure boundary. When running PLAC the user can specify signal variables which are used to vary the conditions at the inlet or outlet and to control the action of valves. One of the simplest ways of changing a boundary condition is to select time as the signal variable, and to the define the flowrates at the fill as a function of time. In this way rate changes can be simply entered as a table of values. PLAC assumes a linear change of the variable with time. There are 25 signal variables available in PLAC at present, allowing a wide range of boundary condition changes such as pressure, temperature, flowrate. heat transfer and flow area. It is beyond the scope of this chapter to consider all combinations, and hence reader is directed to Example 7.3 as an illustration of a boundary condition problem.

Thermal transients such as pipeline cool down and warm up are important to the estimation of the amounts of inhibitor that is required to prevent hydrate formation during periods of zero or low flow. Here WELLTEMP can sometimes be used to screen cases for detailed analysis by more sophisticated transient two-phase flow codes.

Pipeline pack and draw is also often modelled to investigate the survival time of a given system and the ability to meet the gas sales contract.

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7.3.3 Hydrate problems due to shut-down and start-up

In normal operation hydrates can be controlled by maintaining the flowing temperature above the hydrate formation point and/or injecting sufficient quantities of inhibitor. The required dosage is usually determined from running GENESIS, HYSIM or using the GPSA charts for the prevailing conditions. However, during shut-down and start-up, conditions can occur which increase the likelihood of hydrate formation, and hence require additional quantities of inhibitor. Figure 7.10 shows a typical hydrate dis-association curve. A pipeline pressure/temperature profile is superimposed as line 'A' representing conditions in a pipeline from a subsea template to the slug catcher on a host platform, where hydrate inhibition may not be required under normal operating circumstances. In some cases the pipeline may be long and the fluids may cool to seabed conditions where some inhibition may be required. This could also occur as a result of lower flowrates, and is shown as curve 'B'. Hence, in normal operation steady state analysis can be used to determine the amount of inhibitor required.

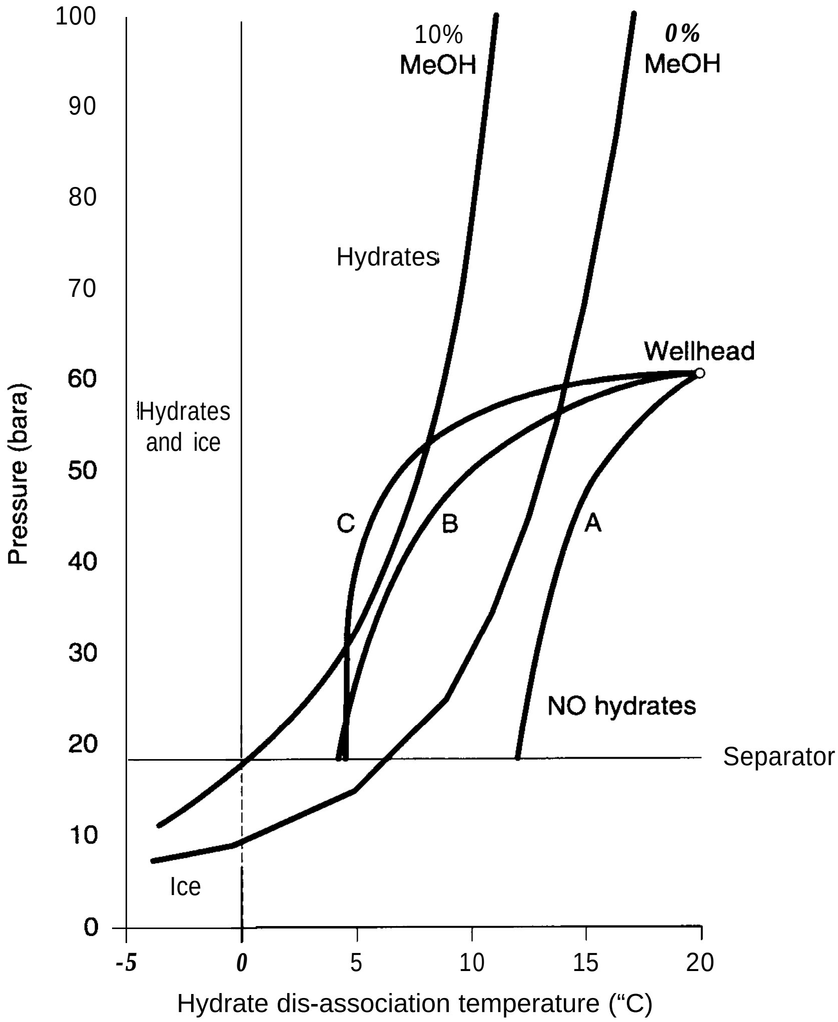


Figure **7.10 Hydrate formation conditions**

During start-up or shut-down transient conditions may arise which can lead to operation in the hydrate region. For example, as the line is started-up the pressure can rise quicker than the temperature giving the pipeline profile drawn in line 'C'. It can be difficult to ensure that the required inhibitor dosage is present during such start-up transients, and hence it is common for inhibitor to be pumped into a pipeline before shut-down to protect against controlled start-up and shut-downs. During unforeseen transients special care may be required.

Another aspect which can complicate the determination of hydrate formation conditions is the time required to form hydrates, ie. the kinetics of the process. If the conditions during start-up are in the hydrate formation region for only a short period of time, then extra inhibition may not be required.

7.4 Transients Due to Topographical Effects

7.4.1 Hilly terrain slug flow and dip slug generation

In all the classical steady state programs used to model multiphase flows the flow regime is assumed to change abruptly as the topography changes. However, in practice the previous flow regime may persist for sometime in the next section. An example of this is flow over a hill where slug flow may be predicted in the uphill section and stratified in the downhill part. In a steady state simulation the slugs would abruptly disappear at the apex. In practice this may not happen, and the liquid slugs may persist for sometime in the downward sloping section which can give rise to a much higher pressure recovery than in stratified flow. This phenomena has been addressed in part by the spreadsheet program HillyFlo, which has been developed by XFE to model slug flow in hilly terrain pipelines. HillyFlo is a steady state program that determines the slug size in uphill inclined sections and models the slug decay in downward sections, hence allowing for the pressure recovery.

The interactions with the terrain may cause additional effects. For example, the liquid sloshing in a dip may cause perturbations in the gas and liquid flowrates downstream, giving rise to flow regimes different to those expected from analysis with steady flowrates. Flowrate fluctuations caused by slugs dissipated in a downward sloping section may be sufficient to generate slugs in a following downstream section normally operating in stratified flow.

By taking account of the hysteresis in the flow regimes and by tracking the local phase flowrates it is easy to see how in practice flows may differ greatly from those predicted by classical steady state methods. It is here that transient two phase flow simulators offer considerable promise if interface tracking methods can be improved. However, one must be cautious about the application to real systems, It can be envisaged that in a long hilly terrain pipeline a section may exist close to the slug/stratified boundary. A small perturbation in the pressure or flowrate may be all that is required to trigger a slug and change the characteristics of the whole pipeline.

In some cases it has been observed that the exit of a slug in a pipeline apparently triggers the formation of another slug. Slugs occur regularly with one, two, or three in the line at one time. This is due in part to the time taken for the liquid level to re-establish, and partly due to the depacking effect as the slug exits, removing the high pressure drop over the slug body. This causes an increase in the gas velocity which can trigger another slug. This is illustrated in Figure 7.11 which shows the passage of slugs throughout a 400 m long, 8” diameter, horizontal air/water test facility. The Y-axis shows the identification number for 30 Light Emitting Diode (LED) slug detector probes, which are located along the top of the pipeline, and give a binary ‘on’ output when water wet and a ‘off’ signal when in air. As the number of slugs in the line increases the influence of each slug exiting diminishes and slug formation becomes less orderly. The topography of the pipeline can exaggerate this effect if uphill sections are present at the outlet. This can increase the pressure fluctuations due to hydrostatic effects, and can lead to severe flowrate fluctuations such as demonstrated by severe slugging. The random type of slugging with considerable decay and coalescence is demonstrated in Figure 7.12.

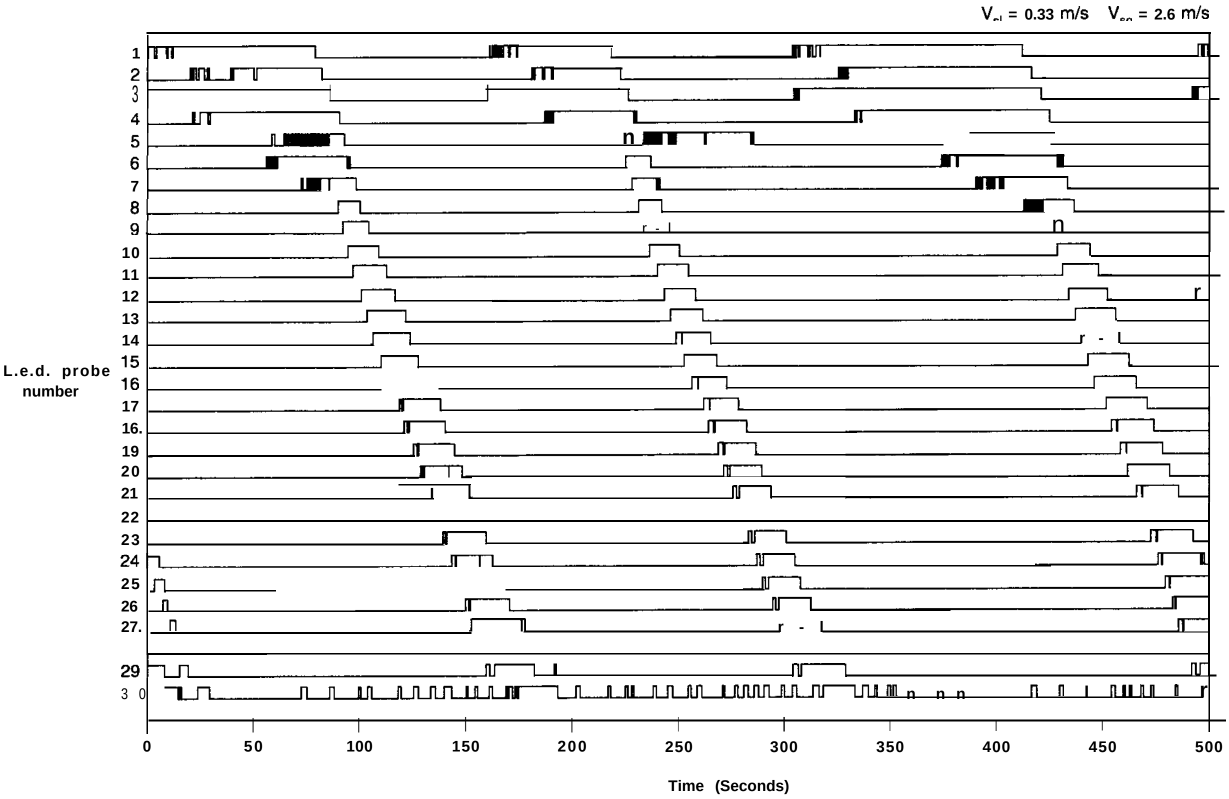


Figure 7.11 Periodic single slug generation

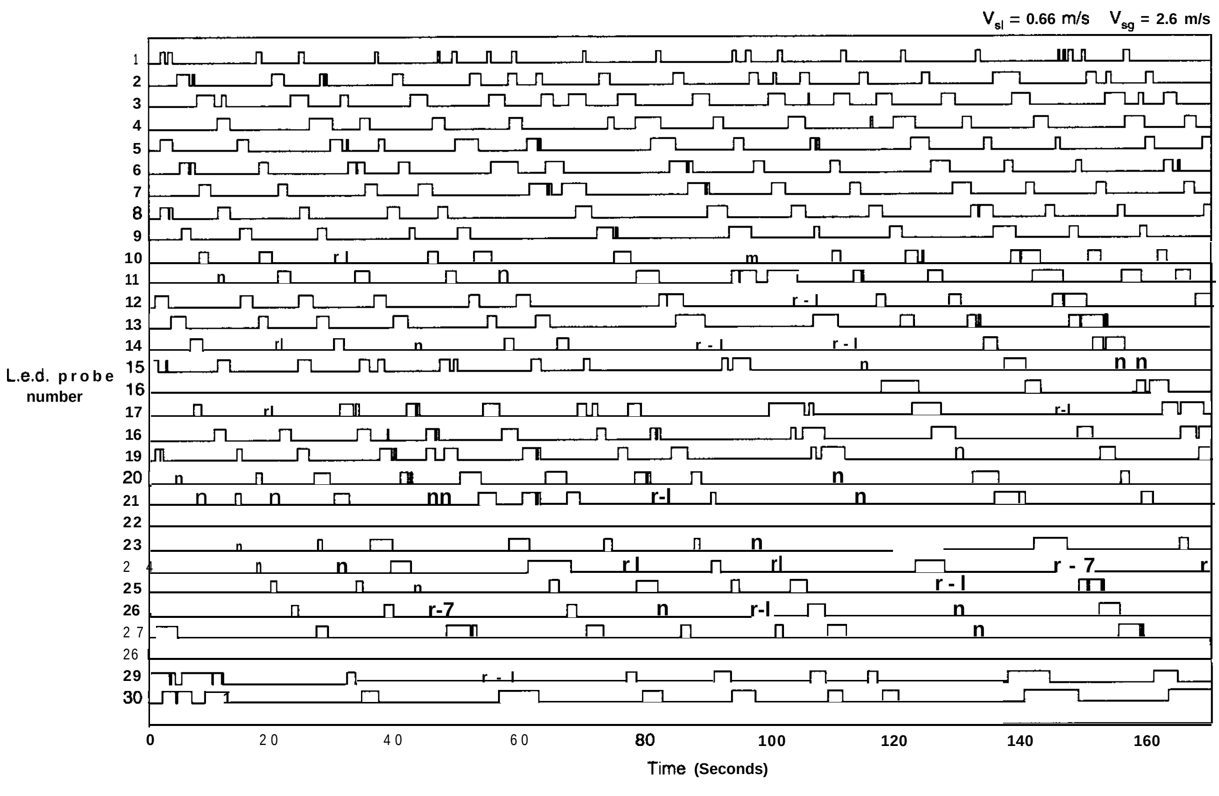


Figure 7.12 Erratic high frequency slug generation

In some cases the bubble length is greater than the pipeline length. Here it can take some time for the liquid level to be re-established before the next slug is generated. Slugs generated in this way can often have the longest lengths, as the equilibrium liquid content in the pipeline can be high. The slug grows as it sweeps up the liquid in front of it, leaving behind a thin film which may take a considerable time to re-establish. Such phenomena may be well predicted by PLAC when the slug tracking model has been implemented. In the meantime it is intended to implement a simple maximum slug length model in MULTIFLO. This has been shown to give good agreement with the slugs generated in the Endicott pipeline under these conditions.

PLAC has been shown to be capable of accurately predicting severe slugging, and can simulate slug generation caused by slug exits (see Section 7.4.2). However, the models in PLAC require improvements in order to properly track the slugs and to simulate slug growth. The code will in general give reliable results when the flow is dominated by hydrostatic effects, such as severe slugging and flows in flexible risers. More work is required in order to predict slug flows in horizontal lines. Here the existing design methods should be used.

Some initial studies using PLAC to predict slug distributions have shown promise. For example Figure 7.13 shows PLAC predictions of the slug length distributions for the Sunbury tests from the 6" air/water horizontal pipeline, the corresponding measured slug length distribution is shown for comparison in Figure 7.14 and gives a mean slug length of 4.4 m and a maximum length of 12 m. It is seen that the agreement between the simulation and the experiments is good. Under these conditions of high frequency slug flow PLAC predicts tall waves rather than slugs. A tracking method is being developed to increase the velocity of these waves to the slug velocity in order to model the slug growth from liquid sweep-up in front. Work is continuing in this area.

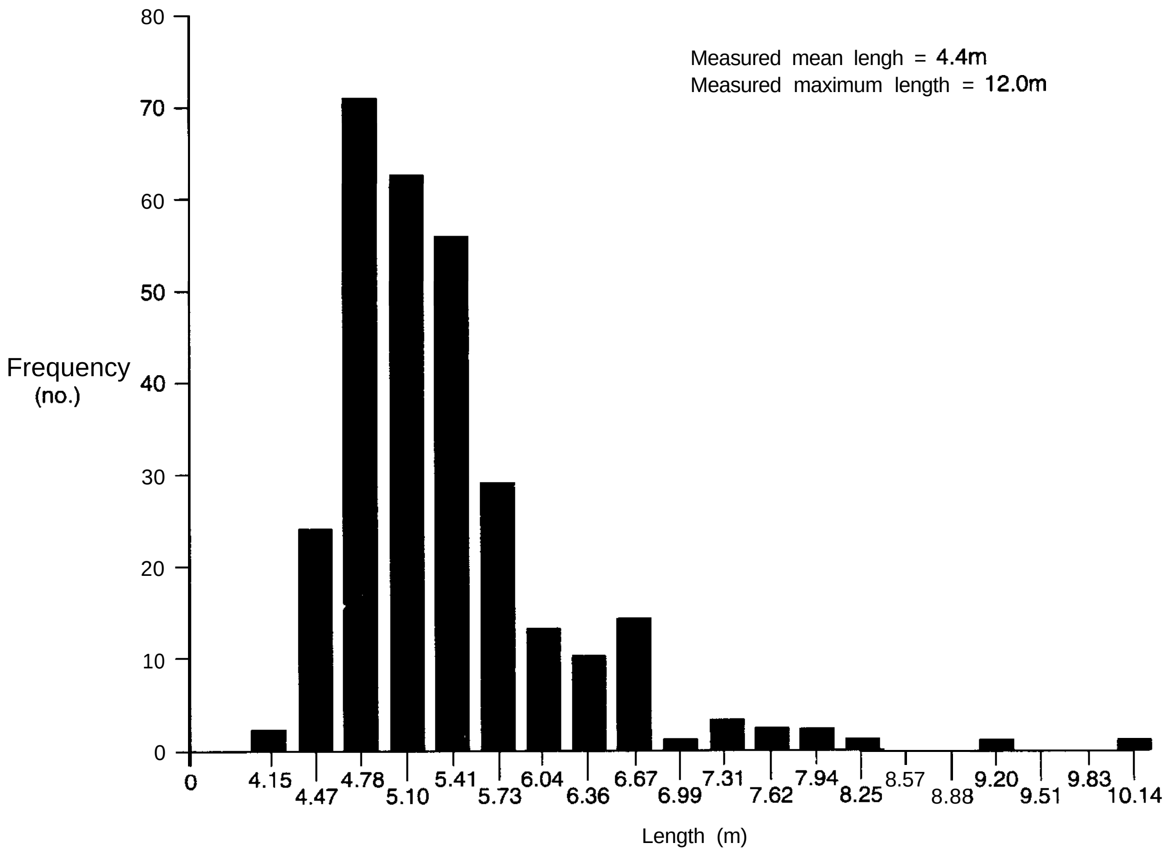


Figure 7.13 PLAC comparison with RCS slug distribution

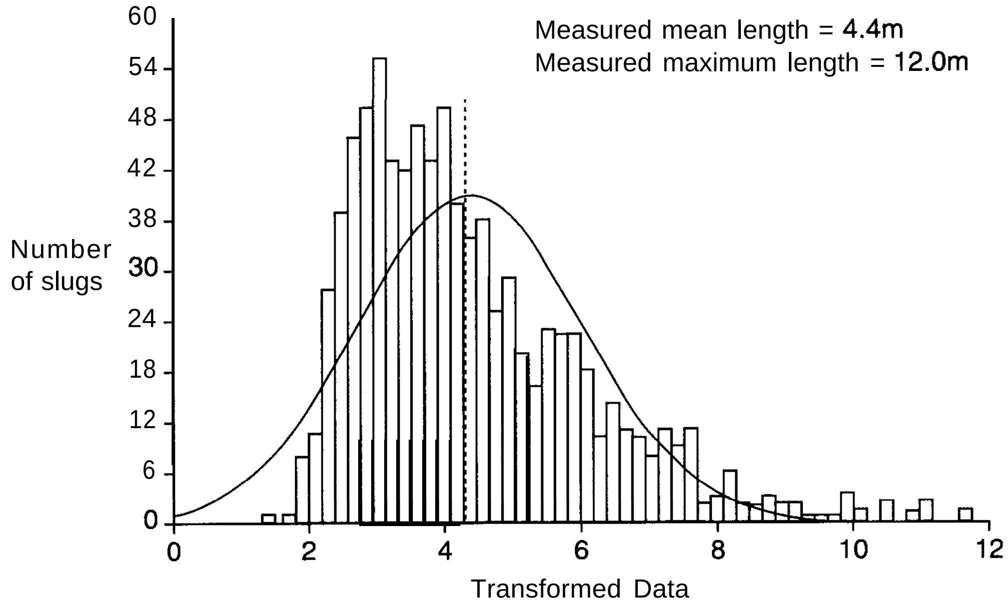
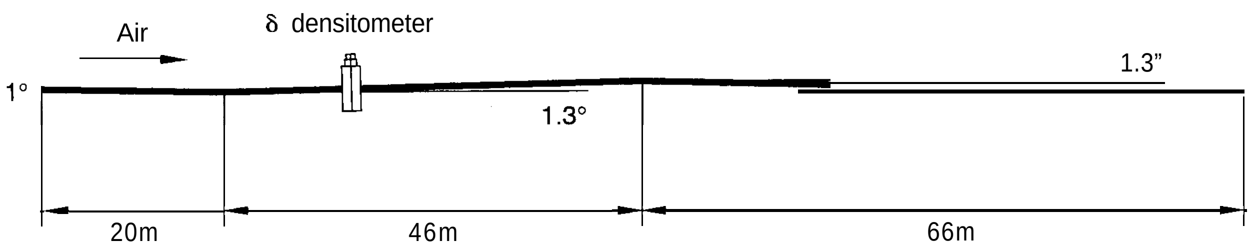


Figure 7.14 Experimental slug length distribution

The extrapolation of existing design methods to predict slug flow in hilly terrain pipelines may be risky, and it is here that dynamic simulation may be of use. This is an area where validation work is underway. PLAC has been compared with other Sunbury experiments performed with a pipeline dip configuration on the 6" air/water rig, the rig set-up is shown in Figure 7.15. In these tests a known quantity of water was poured into the dip and allowed to settle. The air flow was then turned on at a constant rate and the flow behaviour observed. The tests were repeated for a range of initial liquid quantities and air flowrates. At low gas flowrates the slugs are generated at the dip where the liquid depth is the greatest and leads to a higher local gas velocity which is sufficient to produce waves that grow into slugs. This is illustrated in Figure 7.16. At higher gas velocities the slug generation point moves downstream of the dip and the slug generation is more random. This is seen from the densitometer traces illustrated in Figure 7.17. The flow pattern map generated by the matrix of liquid volumes and gas velocities is shown in Figure 7.18, where the solid line shows the limiting gas velocity at which liquid is removed from the dip. At this point the liquid is drained and measured and gives an indication of the equilibrium liquid holdup. This is compared with the terrain induced slug flow model described in Section 4, where the agreement is seen to be generally good.



Rig topography

Figure 7.15 Experimental set-up

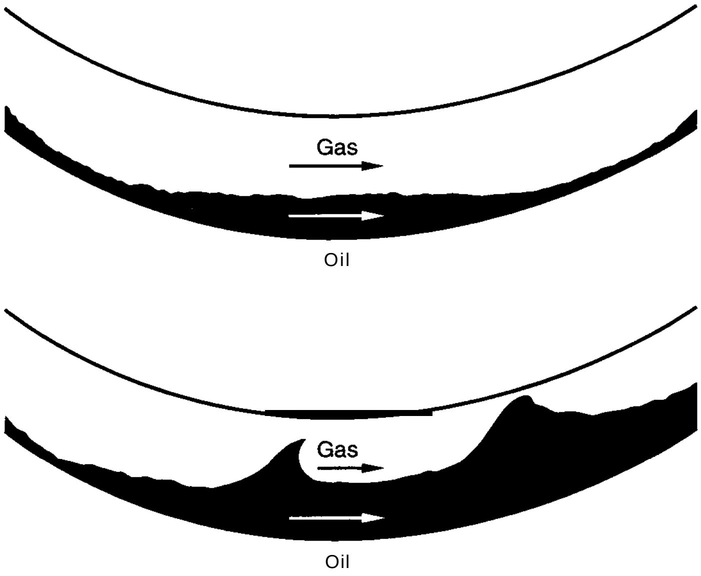


Figure 7.16 Dip slug generation

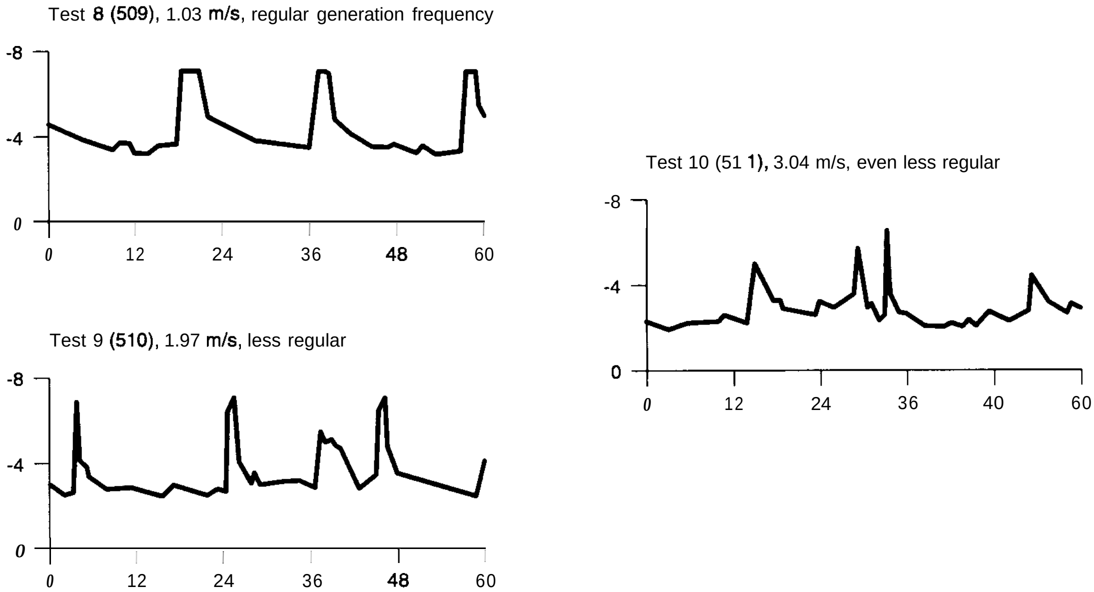


Figure 7.17 Densitometer traces for various gas velocities

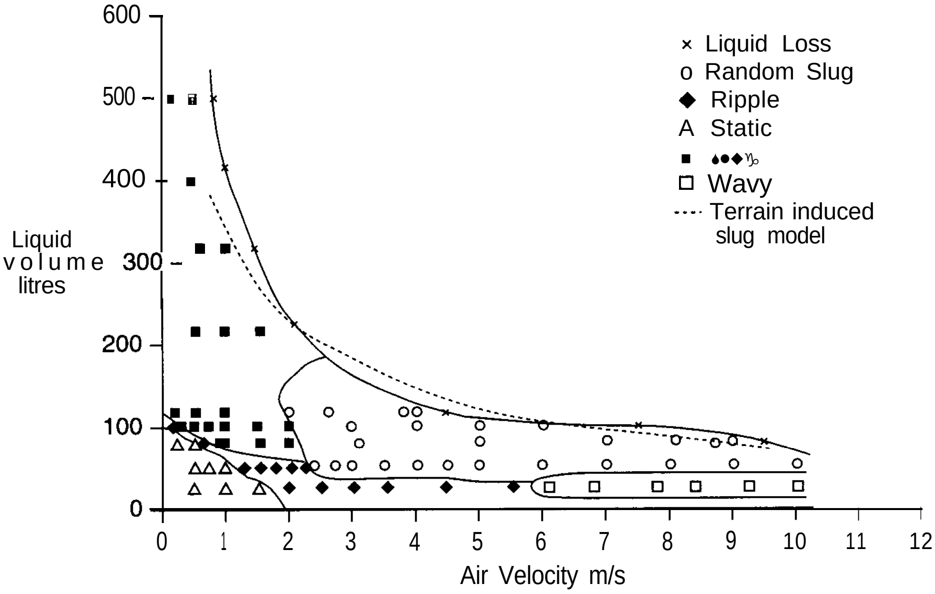


Figure 7.18 Pipeline dip experiments-flow pattern map and equilibrium holdup

Initial comparisons with PLAC have been encouraging as illustrated in Figure 7.19, which shows that PLAC generally predicts the gas velocity at which liquid is removed from the system and the equilibrium holdup, however PLAC tends to underestimate the holdup at the higher gas velocities. Figure 7.20 shows that the predicted slug frequency is generally lower than that measured. However, the results are still in reasonable agreement. This is because PLAC simulates waves rather than slugs, Figure 7.21 shows the oscillations produced by PLAC for case of 317 litres of water with an air velocity of 0.5 m/s.

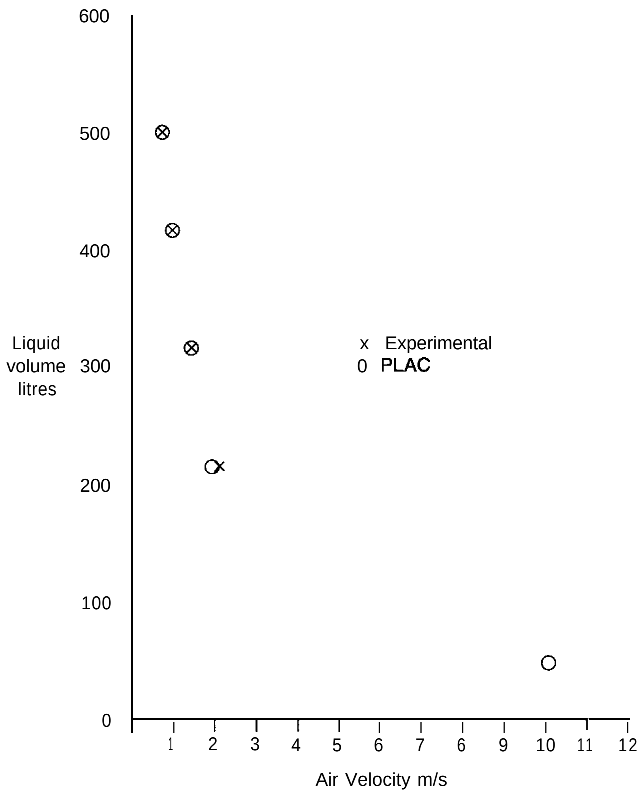


Figure 7.19 Comparison between PLAC and measured liquid removal limit

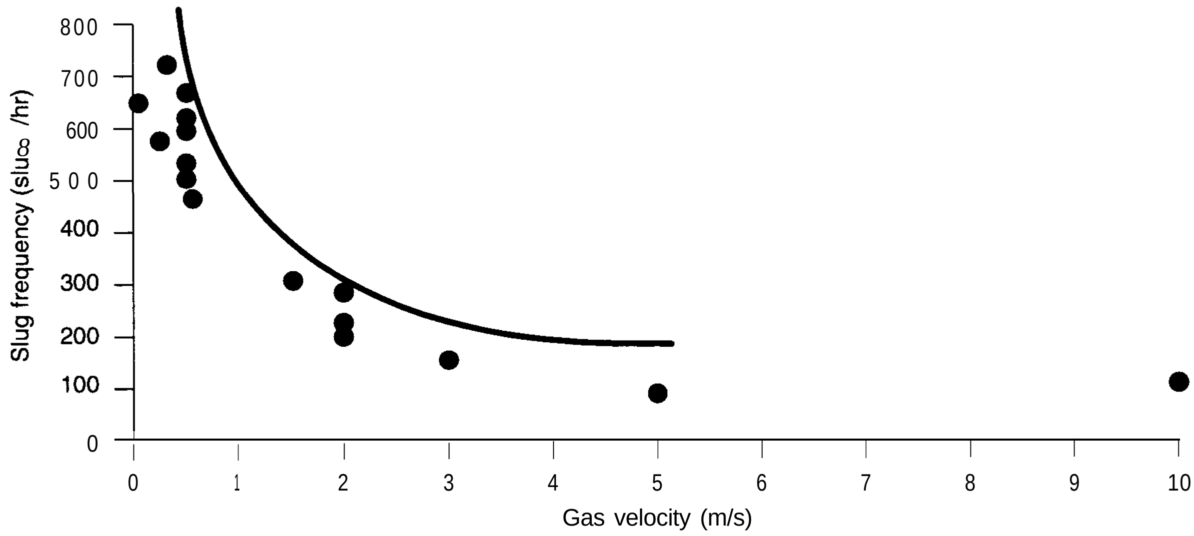


Figure 7.20 Comparison between PLAC and measured slug frequency

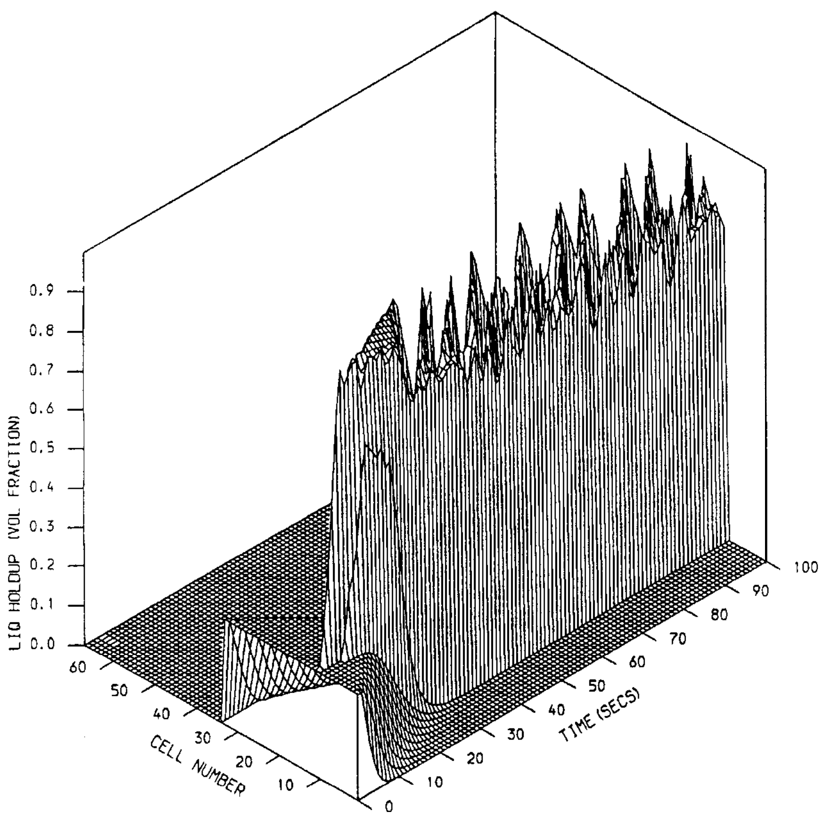


Figure 7.21 PLAC holdup oscillations for 317 l of water and 0.5 m/s air velocity

TRANFLO can also be used to simulate slugging in dips, and was successfully employed to investigate the water motion in the dips of the Amethyst gas pipeline following a failure of the pig launcher. In this case MULTIFLO was used to predict flowing in-situ gas velocity and TRANFLO was used to assess whether there was sufficient motion to provide adequate corrosion inhibitor transport to the dip. Figure 7.22 shows three-phase TRANFLO predictions of the slugging in a dip with gas, water, and condensate present simultaneously, the contours represent the water holdup fraction. The oscillatory motion in the dip (distance = 1 00m) was thought to be sufficient to effectively distribute the corrosion inhibitor.

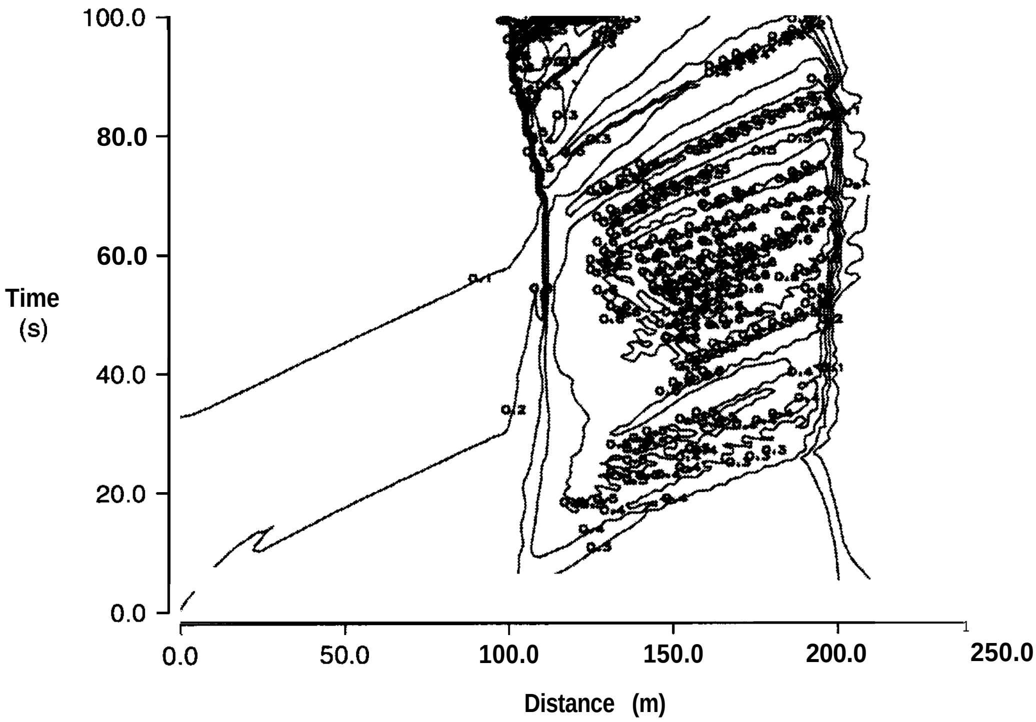


Figure 7.22 TRANFLO simulation of three phase flow in a dip.

In practice oil and gas pipelines will usually reach or attain equilibrium first at the start of the pipeline, hence slugs can be initiated near the pipeline inlet. It is possible for inlet flowrate fluctuations to initiate slug formation. Such inlet flowrate oscillations may result from slugging wells, gas-lift operations, and test well changes for example, and should be suspected as possible causes for flowrate perturbations which generate slugs in otherwise stratified flowing conditions. When using transient flow simulators it is worthwhile to consider the effects of practical flowrate perturbations on the result. This is facilitated in the transient code TRANFLO which allows the user to enter noise parameters at the inlet. The amplitude and the frequency of the inlet perturbations have been shown to have a large influence on the simulation, and have been useful to 'tune' the code. Some good results have been obtained using this technique. However, it is not possible at present to determine what inlet perturbation is required to give the right answer apriori.

7.4.2 Effects of topography on process plant

For gas/condensate systems the point of slug formation usually occurs further down the pipeline, depending on the liquid condensation rate and topography. Slugs are not usually formed at the start since the pipeline conditions can be above the dew point. For typical off-shore gas condensate pipelines terminating on land, the greatest uphill gradients can be at the beach, making this a possible location for slug formation. This phenomena was observed in dynamic simulations of the Bruce/Frigg pipeline system where liquid drop-out occurred when Bruce was shutdown. This resulted in a pressure decrease along the pipeline which gave rise to an increase in the liquid drop-out rate due to retrograde condensation. Predictions using PLAC showed that following a 12 hour shutdown it required around 48 hours to re-pack the pipeline (Figure 7.23). During the shutdown liquid drop-out occurred in the pipeline, which was partly swept out when Bruce gas was re-introduced into the system. PLAC shows that during the start-up phase some of the liquid evaporates as the pressure increases. That liquid which is swept out arrives as a train of 700 bbl slugs which are easily handled by the process plant (Figure 7.24). Another result from the PLAC simulations was the fact that the 12 hour shut-down was not long enough for equilibrium conditions to be obtained, hence the liquid content is less than the equilibrium value. This is contrary to the results obtained using steady state analysis of the holdup change for the cases with and without Bruce gas which indicated that around 4000 bbls of liquid could be removed from the pipeline.

The user should take care with this type of analysis to make sure that the rate of condensation and evaporation is realistic, as it is possible that the simple nature of the phase change model in PLAC may give rise to large errors. For example, the use of a single composition for the fluid may mean that the condensate is re-evaporated at a much lower pressure than would be expected in practice because the condensed liquid would have a heavier composition than the feed stream.

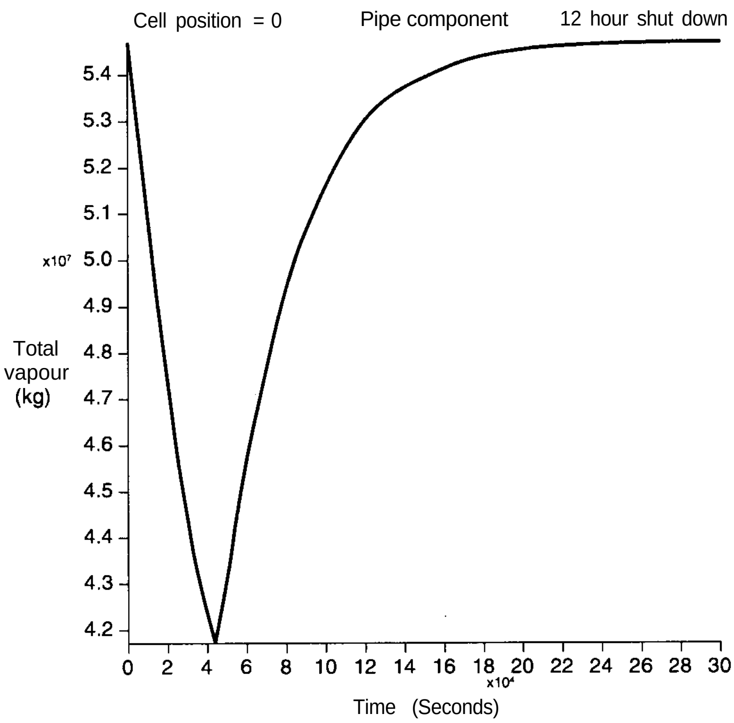


Figure 7.23 Pipeline vapour content

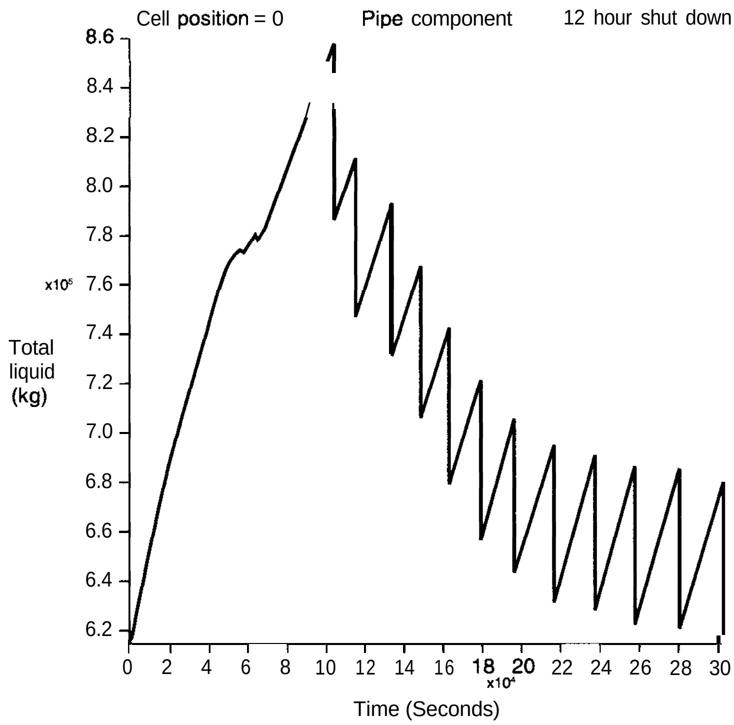


Figure 7.24 Total liquid content showing slugging during re-start

It is seen that there can be numerous transient effects that are caused by the topography. These can be difficult to predict using steady state analysis techniques. The lack of an inclination term in the present slug sizing methods and the inability to allow for hysteresis effects are both examples and HillyFlo goes part way to addressing these problems. Transient simulators can in theory account for both phenomena. However, in slug flow improved models are required to track the slugs. The effects of slugs discharging should be well suited to transient analysis provided that slug generating mechanisms can be clearly defined. However, if a system has numerous slug initiation sites, and is susceptible to small perturbations, then one may expect that the long term accuracy of the simulations may be suspect.

Codes such as PLAC and TRANFLO have produced good results with pipe emptying problems and slug generation with topographical changes. However the slugs produced by PLAC are not real slugs. PLAC can generate slugs produced by gravity effects but is limited in the way that the slug density and velocity are modelled, hence – more work is required.

Dynamic flows produced by topographical changes are important to the design of process plants, as slug characteristics are changed while negotiating the platform riser and pipework upstream of the receiving vessel. If the slug is long, the additional hydrostatic head requires an increase in the pipeline pressure to force the slug up the riser. As the slug is produced into the separator it may accelerate due to the reducing gravitational resistance and the reduced frictional length (Figure 7.25). This can cause large velocity increases which can impact on the process plant control, and can also give rise to large loads on the vessel internals. This is discussed more fully in Section 10 of the Multiphase Design Manual.

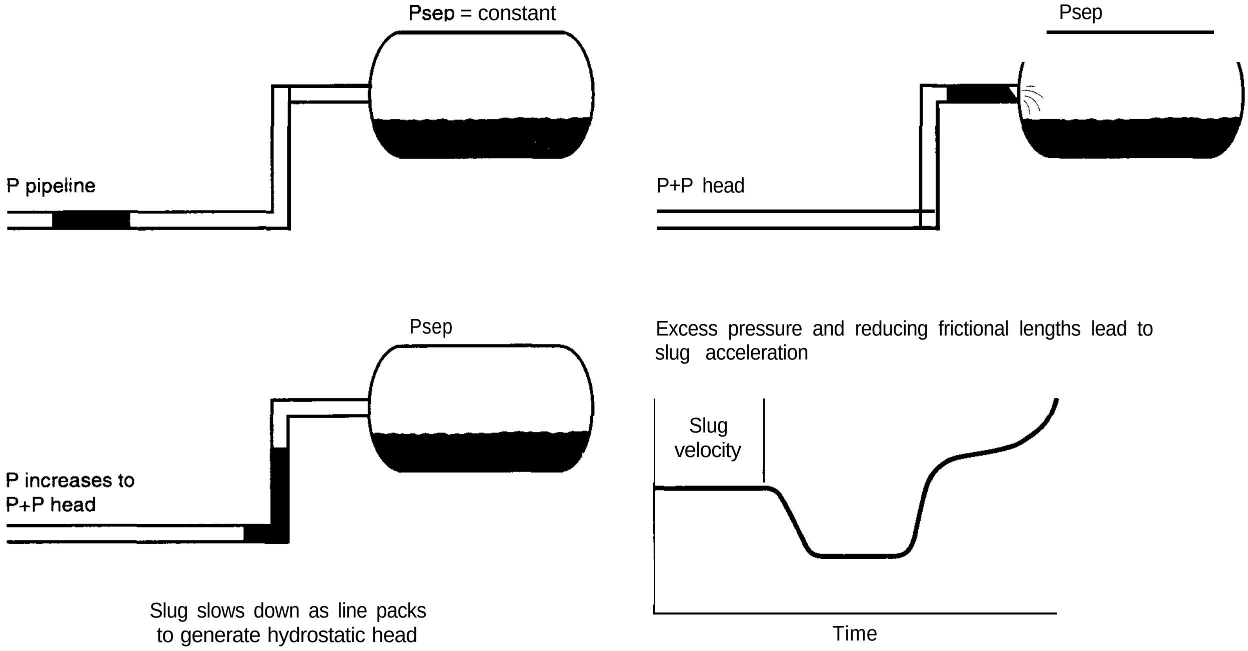


Figure 7.25 Velocity increase during slug reception

Process plant dynamics can be simulated by a number of codes. BPX Sunbury presently uses the SPEEDUP code which has also been interfaced with a simple slug hydraulic model to enable the simulation of slug effects on process plant to be investigated. The slug size is an input to the simulation. However, it is possible to see the effect of the slug catcher pressure on the passage of the slug. Classical control schemes generally reduce the gas compressor speed if the separator gas flow is reduced, hence when the slug is produced the compressor is decelerating. A better solution can be to increase the compressor speed when the gas flow drops off, hence reducing the separator pressure and sucking the slug up the riser. This has the effect of reducing the gas starvation period and also means that the compressor is accelerating when the gas surge occurs.

A recent study using SPEEDUP to investigate control strategies to limit the impact of slug dynamics on the process plant has shown that gas outlet flow control can reduce the peak in the gas surge. As the slug is received the gas flowrate reduces, by opening the gas outlet control valve the pressure in the slug catcher is reduced, hence sucking the slug in and allowing some scope for increase in the slugcatcher pressure to absorb some of the gas flowrate surge.

PLAC can also model a simple separator with controls, including valves. However, it is not capable of simulating the compressor in detail. In the future is is hoped that the transient pipeline and process dynamic simulators will be coupled to facilitate integrated system analysis. A simple interface already exists between OLGA and D-SPICE.

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7.4.3 Hilly terrain pipeline shutdown and restart

Experience with the operation of the Cusiana Phase 1 system in Colombia has shown that maximum flowline pressures may be generated by flowline start-up. The longest multiphase flowline in the Phase 1 gathering system is the 20" diameter, 20 km long flowline that transports the production from wellpads T and Q. The mountainous topography from the remote well T to the Cusiana Central Processing Facilities is illustrated in Figure 7.26. When the production from T was stopped the shut-in pressure at T pad was around 1300 psia compared to the arrival pressure of 640 psia. The high back pressure resulted from segregated liquids draining and filling the pipeline dips, however the liquid levels were suspected to be different, hence causing a non-equaiised hydrostatic equilibrium condition. A simple 2" air/water model of three pipeline dips was constructed at Sunbury to demonstrate what happens during the shutdown and restart and it was confirmed that it is possible for the liquid to drain into the dips with unequal levels, hence producing a back pressure due to the hydrostatic difference in the levels. Figure 7.27 shows the topography of the test rig as modelled by PLAC whereas Figures 7.28 and 7.29 show the PLAC simulations of the liquid draining and inlet pressure variation.

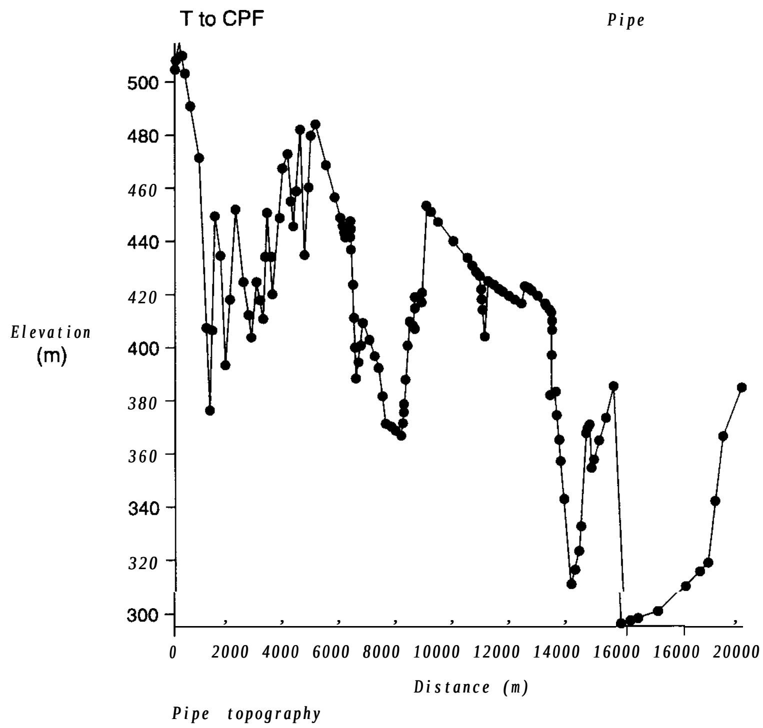

Figure 7.26 Cusiana T pad to CPF topography used in PLAC

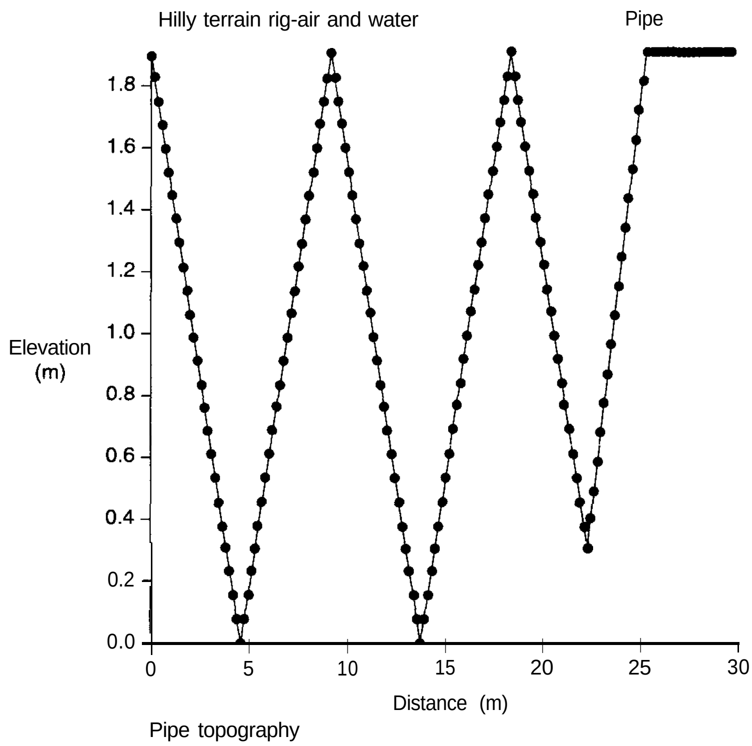

Figure 7.27 Topography of of hilly terrain test rig

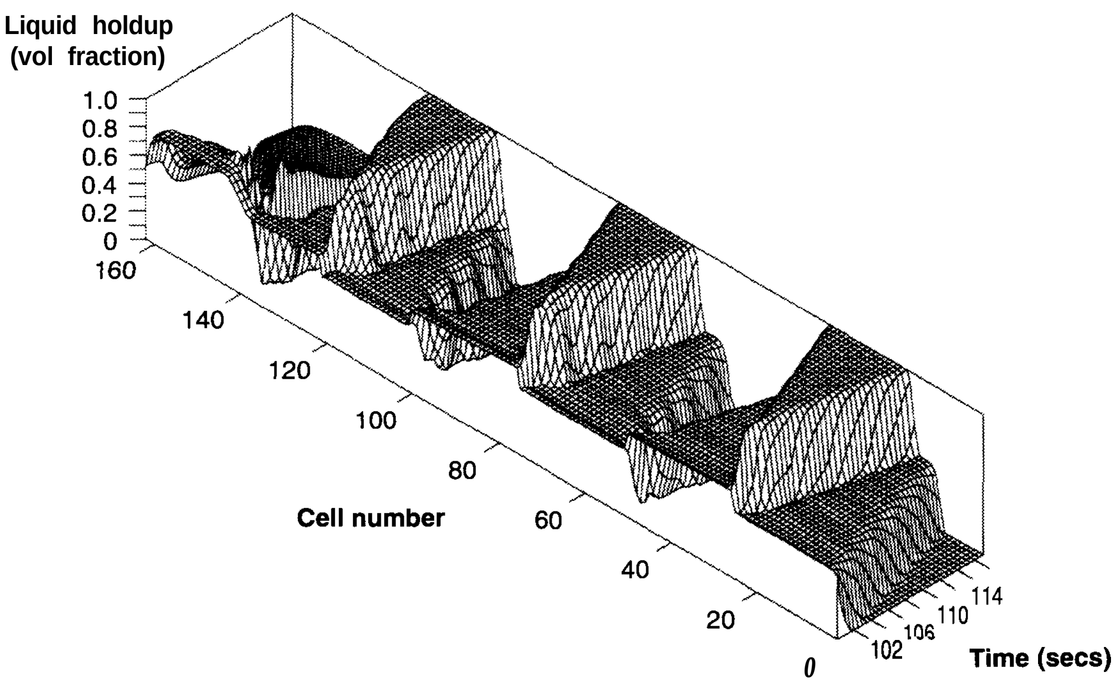

Figure 7.28 PLAC simulation of liquid draining into dips

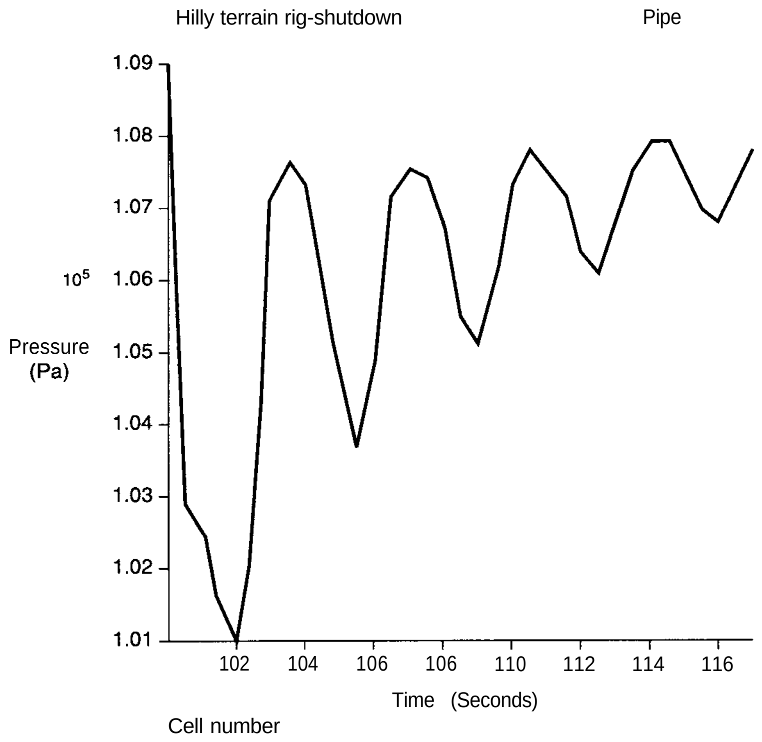

Figure 7.29 PLAC simulation of back pressure due to liquid settling in dips

While it is seen that PLAC can simulate the liquid settling behaviour, the comparison with the experimental results is not always good. This is because PLAC does not model slugs properly and hence the holdup in the uphill sections is sometimes underestimated, hence the initial condition is not always correct.

The highest back pressures can be generated when the flowline is restarted as it is possible for the liquid in the dips to be displaced downstream before the gas is able to penetrate the dip, and this increases the difference in the hydrostatic levels. Once the gas penetrates the pressure is reduced as the gas lifts the liquid out of the dip. One can imagine that the start-up pressure can be large if all the liquid in the dips is displaced simultaneously.

A simple method has been established to estimate the pressure rise during the start-up of hilly terrain flowlines. For long hilly flowlines the total uphill elevation changes may be very large and hence the total theoretical hydrostatic head may lead to very conservative start-up pressures. It is hence recommended that the flowlines are started up by unloading the sections closest to the facilities first. The pressure can be estimated by summing the uphill elevation changes and calculating the hydrostatic head for the section being started-up. In practice experience with the start-up of Cusiana T pad indicates that the actual hydrostatic pressure generated was around 2/3 of the maximum theoretical value, however this is very dependant on the particular system.

If the analysis of start-up pressures is required to check the design pressure of the flowline, then the topography should be investigated to determine if there are low points where the pressure can be higher than at the well pad. This is the case with the T-Q flowline where the maximum pressures are generated in the valley just downstream of the wellpad.

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7.4.4 Pipeline-riser interactions

A special topographical effect is the interaction between a pipeline and a vertical platform riser. This can give rise to a phenomenon called severe slugging if the pipeline inclination is downhill at the base of the riser and if flowrates are low enough. Severe slugging has been discussed in Section 3.3.3 and will not be described again here. However, it is useful to demonstrate that dynamic simulation can be employed to accurately predict the size of severe slugs and the cycle time, enabling topside process plant to be sized. One of the validation cases for the PLAC code was a comparison with severe slugging experiments carried-out by BP at Sunbury. The test rig consisted of a 45m flowline inclined downhill at 2" followed by a 15m vertical riser. The rig operated at near atmospheric conditions and the test fluids were air and water. A schematic of the rig is shown in Figure 7.30 where the pipeline inner diameter is 2". Figures 7.31 and 7.32 show the measured slug size and frequency for a range of fluid superficial velocities.

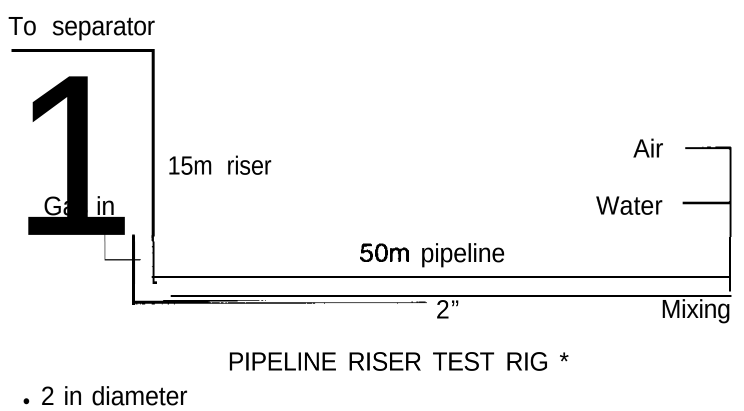
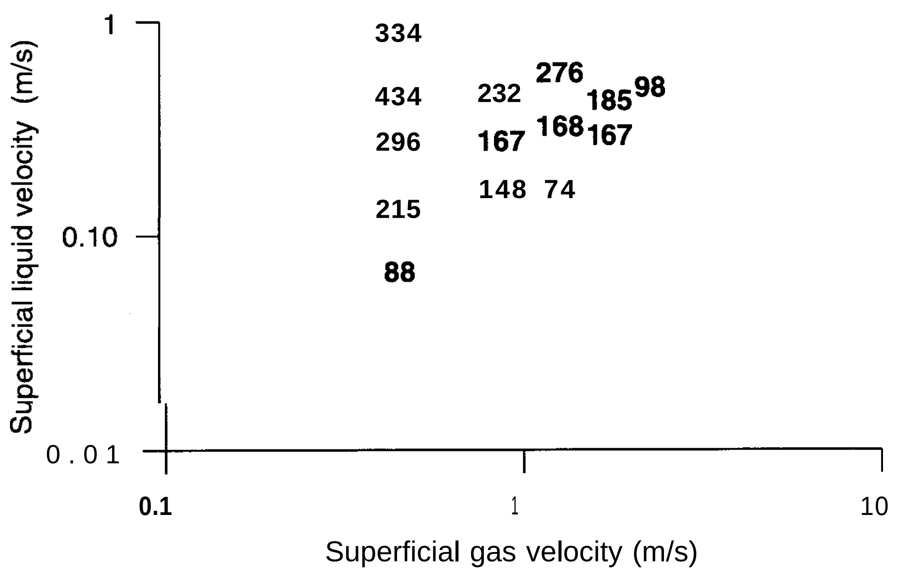
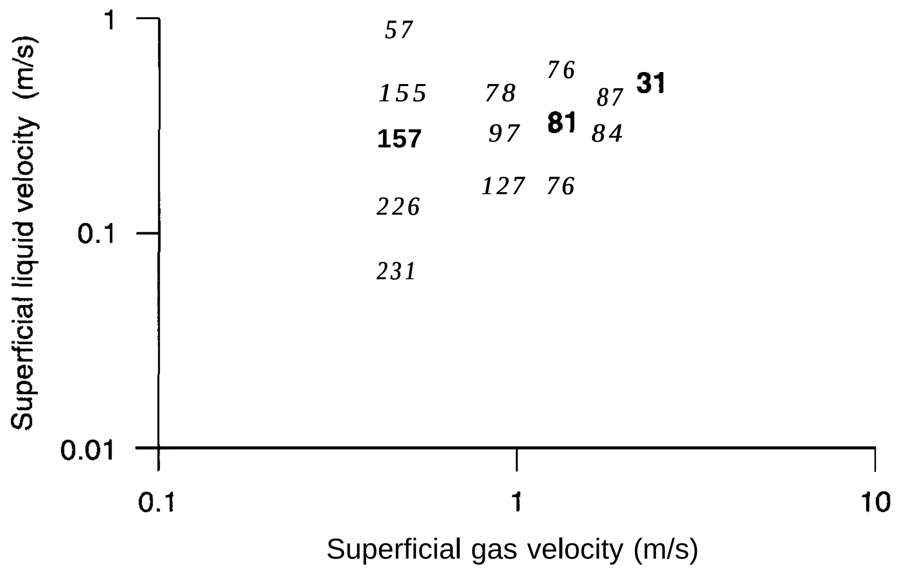


Figure 7.30 Schematic of severe slug test rig.



Figures 7.31 Experimental slug volume (% riser)



Figures 7.32 Experimental cycle time (s)

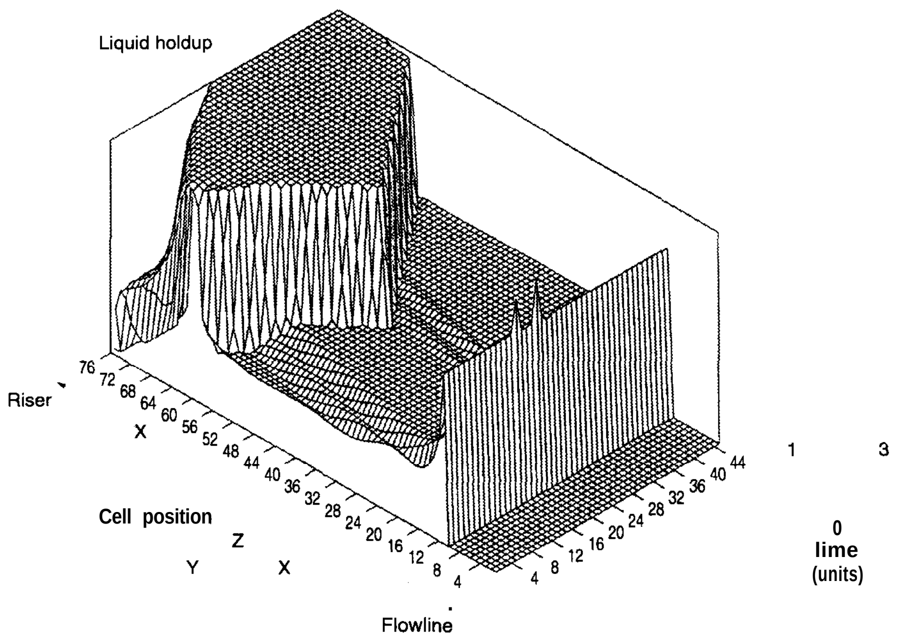


Figure 7.33 PLAC holdup profile prediction.

Figure 7.33 shows the liquid holdup profiles predicted by PLAC for gas and liquid superficial velocities of 0.44 m/s and 0.43 m/s respectively. The Figure shows the liquid build-up and blowdown. Figure 7.34 illustrates the intermittent nature of the outlet liquid mass flowrate, showing the rapid increase in the flowrate as the tail of the slug accelerates up the riser. For this case the experiment gives a severe slug size of 434% of the riser volume and a severe slug cycle time of 155 seconds. The PLAC simulation gives 430% and 150 seconds respectively. It should be noted that it is important to correctly model the volume of the system in order to match severe slugging cycle times. In the case of the Sunbury experiments the model in PLAC included a section of air filled pipe at the inlet which was used to model the line supplying air to the mixing section.

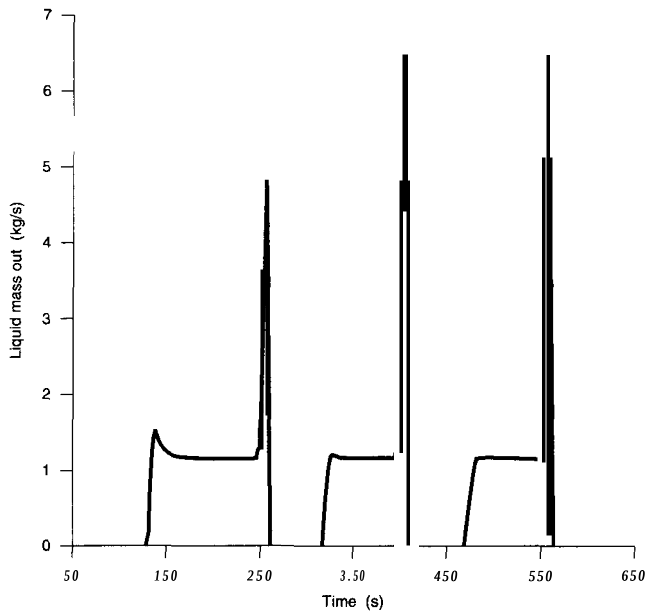


Figure 7.34 PLAC prediction of outlet liquid mass flowrate.

A similar type of severe slugging can also be observed when flexible risers are employed, particularly if the lazy 's' configuration is adopted. The analysis of the flows in these curved configurations is complicated by the possible bi-directional motion of the flow, and the interaction with the complex geometry. Here transient simulators show promise in being able to predict the flows, whereas the scale-up of laboratory simulations to practical situations is uncertain. Example 7.4 shows a comparison between PLAC and some laboratory experiments of severe slugging in a catenary riser.

The relationship below can be used to estimate the critical liquid velocity at which severe slugging occurs in a pipeline-riser system. If the liquid superficial velocity is above this value severe slugging is unlikely to occur.

$$V_{sl} = V_{sg} P_{sep} / [L(\rho_l - \rho_g) g (1 - H_l) \sin \beta]$$

where:

P_{sep} = pressure at flowline outlet (N/m*)
L = line length upstream of riser (m)
ρ_l = liquid density (kg/m^3)
ρ_g = gas density (kg/m^3)
H_L = average liquid holdup in line
g = acceleration due to gravity (9.81 m/s*)
ß = inclination of riser from horizontal

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7.4.5 Transients Caused by Looped Pipeline Systems

Pipeline looping is a common practice to increase the capacity of systems transporting single phase oil or gas. The parallel loop may be installed along the whole or only part of the pipeline to increase the effective flow area, and hence to remove a bottleneck in the system or to facilitate an increase in the flowrate. In single phase flow the analysis of the hydraulic performance of the loop is usually determined using a single pipe with an equivalent diameter. This method is also often applied to the looping of two-phase pipelines using simple tee junctions, where only one leg with half the flowrate would be analysed if both pipes in the loop have equal diameters, for example. Inherent in such an analysis is the assumption that the gas and liquid splits in equal proportions in both of the legs. However, in practice this is not usually the case.

The two-phase flow split at pipeline tee junctions is complex, and at present there is no universal method for determining the flows to each leg, due to the dependence on the geometry of the tee, and the various gas and liquid flowrates and physical properties. However, it is apparent that in most cases the flow split at a side arm tee junction is usually very different from being equal. See Section 11 of this manual.

At very low liquid fractions in stratified flow typical of gas transmission systems with a small amount of liquid dropout or carryover, it is possible for the liquid in the stream to follow the gas into the side arm of the junction, and hence the run remains virtually dry. Instances have been recorded where liquid has consistently taken one path in a complex gas transmission network, resulting in overfilling pipeline drips, and causing an interruption in the supply of gas, although the overall liquid content of the liquid transmission system is very low. At the other extreme, in slug flow for example, the inertia of the liquid causes the majority of it to flow forward into the run, whereas all the gas may flow through the side arm. In fact some simple tee junctions are used as in-line separators.

It is hence seen that the flow split at a simple tee may be very different from equal, and hence the flowrates in the individual legs will be unequal. However, since in a loop the lines rejoin downstream, the number of possible solutions is reduced to those flow splits that give rise to the same overall pressure drop in both pipelines.

Transient flows can result from the operation of looped pipelines since it is possible for a number of solutions to exist for the flow split. For example, high flow in one leg causing a high frictional pressure drop may be balanced by low flow in the other with a high hydrostatic component and high holdup. Small inlet flowrate perturbations at the tee, or changes in operating conditions, may cause the flowing conditions in each leg to change, with the possible removal of excess liquid in the form of a slug. In the extreme it is possible in pipelines with a large uphill elevation for a manometer effect to exist. Here the flowing pressure in one leg is balanced by a static column in the other. A change in operating conditions can cause the static liquid to be swept-out. This type of phenomena has been known to give rise to operational difficulties with gas transmission pipelines laid over hilly terrain.

Problems with parallel loops are well documented in the power generation industry, where instabilities can occur in a bank of boiler tubes, for example. The total flow into and out of the headers is normally constant, as is the pressure drop. However, flows in the individual tubes can vary considerably. In such situations the analysis is complicated by the addition of heat transfer effects and steam generation, which can give rise to multiple solutions for the location of the boiling front. So called ‘parallel channel’ instability in the boiler tubes can cause corrosion problems or dry-out leading to premature failure. Experience with such situations has shown that adding flow restrictors to the inlet header has a stabilising effect, whereas throttling the outlet can exacerbate the problem.

The method proposed to analyse possible transient problems caused by operating looped pipelines is along the lines outlined by Gregory and Fogarasi (Ref 7.1). This involves calculating the two-phase pressure drop in each leg for a range of liquid and gas flow splits. A graphical solution is employed to determine the conditions where the pressure drops in each leg are equal, and hence the possible operating regimes of the loop. Allowance should then be made for flow excursions between the parallel lines and the possible displacement of liquid.

As an example consider the schematic shown in Figure 7.35

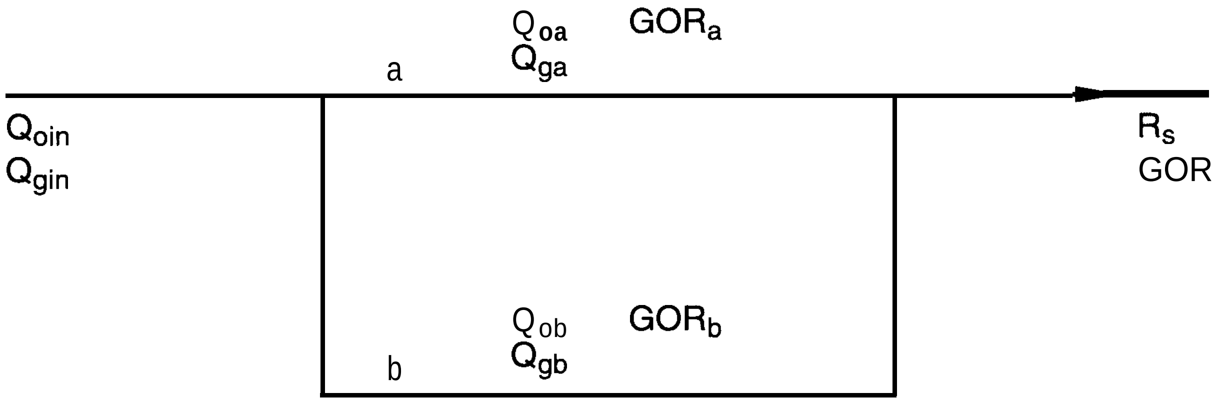

Figure 7.35 Schematic of a pipeline loop

Let α equal the fraction of the inlet liquid flowrate in line A and β equal the fraction of the free inlet gas in line A, Hence:

$$Qoa = \alpha Qoin \text{ and } Qga = \beta Qgin$$

The free gas at the inlet to the loop is given by the difference between the producing gas-oil ratio and the solution gas-oil ratio at the inlet conditions.

hence:

$$Qga = \beta(GOR - RSin)\ Qoin$$

These expressions can be used to determine the effective producing gas-oil ratio to use in the two-phase pressure drop analysis as follows:

$$GORa = [Qoa\ RSin + \beta(GOR - RSin)Qoin]\ /\ Qoa$$

hence:

$$GORa = RSin + \beta/\alpha\ (GOR - RSin)$$

Similarly for line B:

$$GORb = RSin + (1 - \beta / 1 - \alpha)(GOR - RSin)$$

Steady state pressure drop calculations can be performed for a range of gas and liquid splits using the effective GOR calculated above. The value of the solution gas-oil ratio at the inlet conditions can be determined using the PVTTAB facility in MULTIFLO based on the inlet temperature and pressure.

Figure 7.36 shows the locus of the possible solutions for a equal pipeline loop and indicates that the 50/50 solution (α = 0.5) gives rise to the maximum downstream pressure, and hence minimum pressure drop, however this is not always the case. For other pressure drops there are two flow split solutions and hence it could be possible for the flows to oscillate between the two solutions.

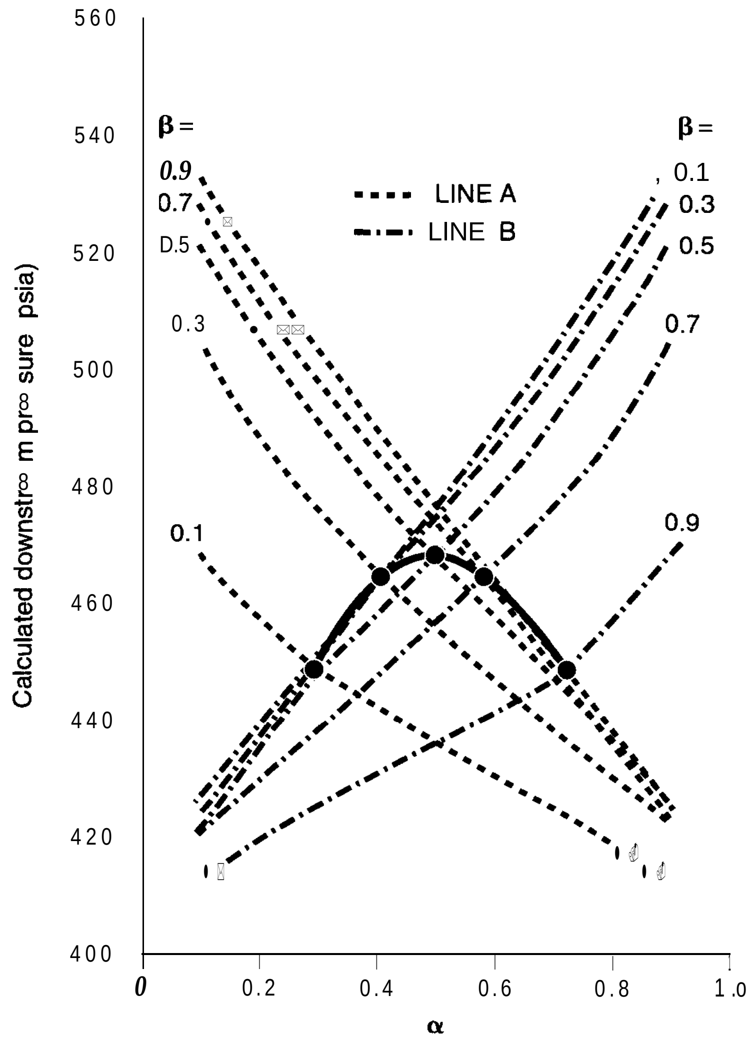

Figure 7.36 Locus of solutions for equal pipeline sizes

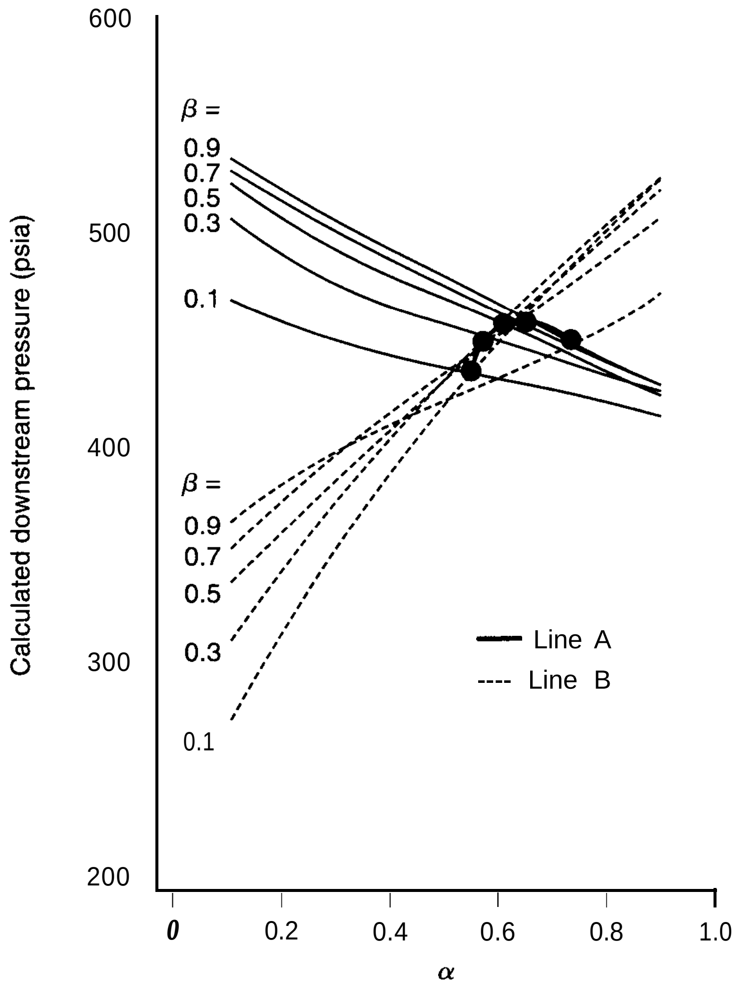

Figure 7.37 Locus of solutions for unequal pipelines

Gregory et al (Reference 7.1) suggests that the existence of a manometer leg effect can be checked if the net elevation increase over the loop is such that the maximum possible hydrostatic head that could result from the liquid phase static in the pipeline is greater than the pressure drop that will result from the total throughput of gas and liquid flowing in one of the parallel lines. This may not be wholly true as a symmetrical hill would have a net elevation change of zero. However the potential exists for the uphill section to contain liquid and provide a manometer effect. It is hence recommended that the manometer effect check be carried out using the sum of the uphill elevation changes. If one of the legs contains static liquid it is obvious that the loop has in fact not increased the capacity of the system as the flow only occurs in one leg.

In the above example the pipes in the loop are identical and hence the locus of possible solutions is symmetrical. This is not the case when the pipes are different as illustrated in Figure 7.37 for a 10" and 12" loop from Gregory's paper. Here the locus of the possible solutions is not symmetrical and the minimum pressure drop occurs with a liquid split of 65/35.

Example 7.5 shows the analysis technique applied to a proposed multiphase pipeline loop over mountainous terrain in Colombia.

The main points to remember about the hydraulics of looping two-phase pipelines are as follows:

1. If the pipes in the loop are identical, the assumption of equal split may lead to an under prediction or over prediction of the pressure drop in practice, as equal splits are unlikely. For some conditions equal split can give a minima in the locus and for others a maxima.

2. A range of flowing solutions are often possible which give equal pressure drops. While this may give rise to steady outlet flowrates, the natural perturbation produced by multiphase flows, and the complex characteristics of flow splitting, can lead to oscillations between the legs, and consequent transient liquid sweep-out.

3. Although the outlet flowrate from the loop may be steady at one condition, a small change in the operating conditions can lead to transient liquid sweep-out as the new operating point stabilises.

4. A check should be made for manometer effects based on the sum of the uphill elevation changes, although this may not give the maximum holdup change as a pipe full flowing case may also exist. If manometer legs occur the loop will not increase the flowing capacity.

5. Static liquid legs may have implications for corrosion.

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7.5 Transients due to pigging pipelines

7.5.1 Introduction

A method for estimating the slug sizes resulting from pigging pipelines is presented in section 4. This approach assumes that the pig travels at the in-situ mixture velocity, which is not always the case in practice. The motion of the pig may vary as a result of the compressible nature of the fluid in the pipeline, which allows pressure and velocity excursions to arise from the variations in the pipeline inclination and the friction between the pig/pig train and the pipeline. For example, the pig may require additional driving pressure in order to push the slug over a hill or to compensate for an increasing frictional resistance as the slug grows in length. If the system is 'spongy' the slug velocity may decrease or even stop as the fluid packs behind the pig to provide the additional driving pressure. In downward sections the head recovery can cause the pig to accelerate and transfer the excess pressure into friction loss.

If the pig stops, then it is possible to observe a 'stick-slip' phenomena caused by static and dynamic friction effects, where more pressure is required to get the pig moving than is required to keep it moving. The additional pressure generated after the restart can cause the pig to accelerate, de-packing the line, and hence stopping again.

These phenomena can have a large effect on the time required to pig a pipeline, on the dynamic pressures generated, and on the fluids produced. This may affect the design and operability of the system. In addition, it is sometimes important to predict how long it takes to re-establish the liquid holdup behind the pig, and hence to re-distribute corrosion inhibitors.

Simple pigging transients can be investigated using the OLGA and PLAC transient two-phase flow codes, both of which allow for a variable leakage past the pig during the pigging process. The PLAC model is a recent addition and has yet to be properly validated, but is used here to demonstrate the transient effects caused by pigging pipelines.

7.5.2 Transient pigging example-pigging a hilly terrain pipeline

Figure 7.38 shows the topography of a hypothetical hilly terrain pipeline which is 200 mm in diameter, 10 km long, and includes a 50 m high hill followed by a 50 m deep dip. A high GOR composition was used and the simulation run to a steady state with an inlet mass flow of 2 kg/s and a delivery pressure of 10 bara. A pig was launched into the pipeline at time 20100 s and it was estimated that the pig transit time would be around 2900 s based on the average mixture velocity in the pipe. Figure 7.39 shows the pig position with time, and indicates that the pig motion was not steady and displayed three regions of steady motion and two regions of slower motion, and in some cases reversal. The actual pig transit time is 5900 s, which is almost twice as long as estimated from the average mixture velocity. This discrepancy results from the fact that the pig slows down in the uphill section as the pressure builds-up.

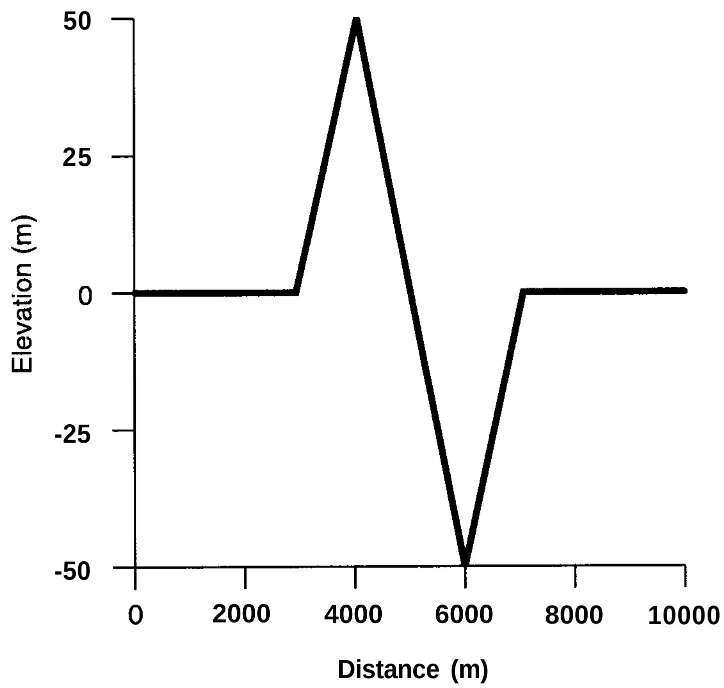

Figure 7.38 Topography of hilly terrain pipeline

Figure 7.40 shows the holdup profile throughout the pipeline for various times and indicates that before the pig is launched the holdup in the upward inclined sections is around 35 %, whereas in the horizontal and downward sloping sections it is around 2 %. The pig slug is shown just after the pig has passed the top of the first hill and indicates that the holdup is around 95 %, and that the slug is less than 500 m in length. The holdup profile at 40,000 s indicates that the liquid content of the pipeline is still building-up 4 hours after the pig has been received. Figure 7.41 shows the pressure at the pipeline inlet and indicates that a pressure rise of 1.3 bara is required for the pig to negotiate the first 50 m hill. The liquid full hydrostatic head would be equivalent to 2.8 bara, hence the generated pressure is consistent with the observation that the pig slug half fills the upward inclined section.

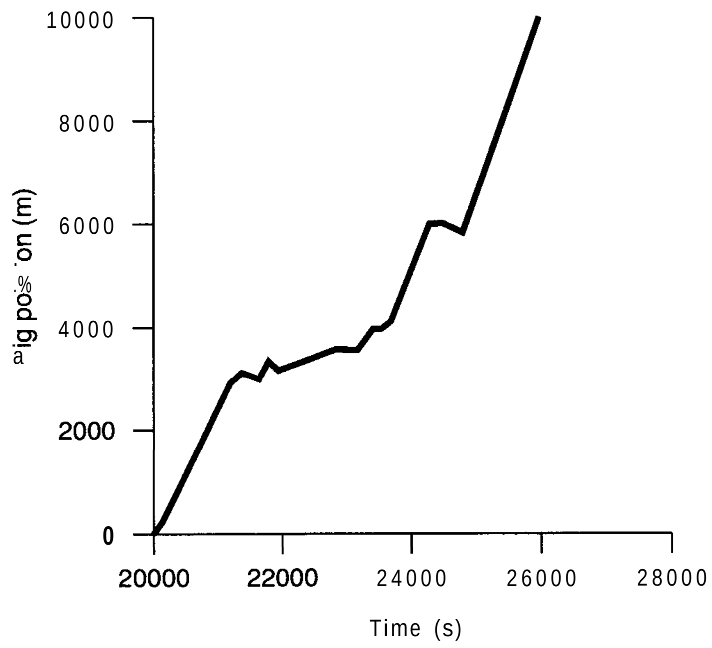

Figure 7.39 Pig position with time

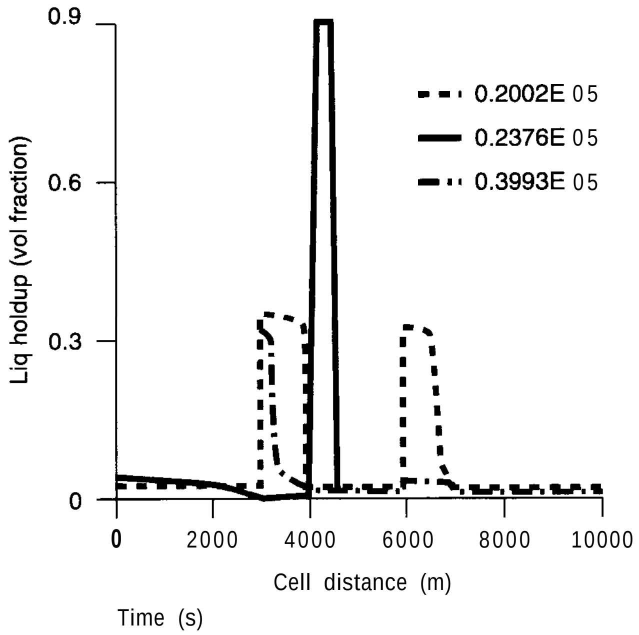

Figure 7.40 Pipeline holdup profiles

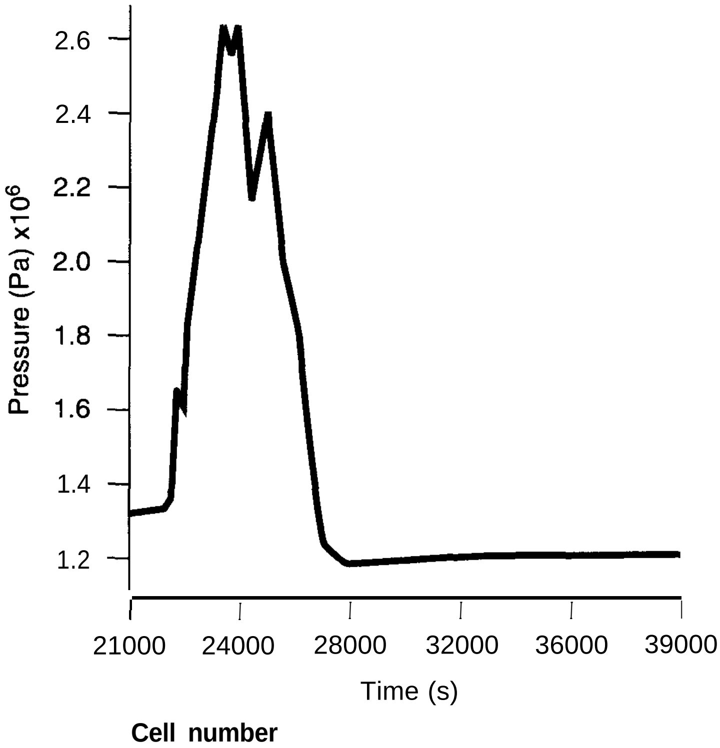

Figure 7.41 Pressure at pipeline inlet

Although the new pigging model in PLAC is simple in nature, it is seen that the general motion of the pig can be very different from that expected by simple steady state analysis. Future modifications to the model are required to improve the treatment of the pig/pipewall friction and code stability.

7.6 Pressure surge transients in two-phase flows

Pressure surge transients in single phase flows are well understood, with computer packages available to predict the surge characteristics of complex pipeline networks with components such as pumps, valves, and surge suppression devices. A computer program called FLOW-MASTER is extensively used by BP Exploration for this purpose. When a two-phase flow is introduced into the system the analysis is made far more complicated since the speed at which the pressure waves travel is modified by the gas fraction. In addition, the discontinuous fluid interfaces in two-phase flow provide additional locations for pressure waves to be reflected and attenuated. Such is the complication that no general purpose pressure surge code exists for two-phase systems. There are however a number of ways that two-phase surge pressures may be analysed with FLOWMASTER and these will be outlined below. We will begin with a brief introduction to single phase pressure surge phenomena, followed by a discussion of the effects of two-phase flow, supported by some examples.

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7.5.1 Single phase hydraulic surge analysis

When a velocity change occurs in a flowing fluid some of the kinetic energy is converted into pressure energy giving an effect called hydraulic surge or water hammer. In most single-phase pipelines this can be caused by delivery valve closure and pump start-up or shutdown. There is normally the potential to create pressures significantly above or below those occurring during steady conditions giving a potential for overpressuring and subsequent rupture or the generation of low internal pressures leading to pipe collapse. In the case of a closing valve the effect of the increased restriction is to reduce the flowrate. At the instant of the increased restriction the flow further upstream of the valve continues at its previous rate which leads to the conversion of kinetic energy into potential energy. The slowing or stopping of the fluid travels as a pressure wave against the previous flow direction. In simple systems the highest pressures are generated when the flow is stopped instantaneously, however in practice the magnitude of the surge pressure is influenced by pressure waves reflected from the other end of the system. These waves require a finite time to return to the location of the flow stoppage, which is called the pipeline period. If the flow has completely stopped within the pipeline period, then an 'instantaneous' surge pressure will be experienced.

The pressure wave time period will be modified by the formation of vapour cavities that occur when the pressure falls below the vapour pressure. The pressure waves also experience a sudden change in magnitude when passing a change in the pipeline diameter. A pressure surge can also induce a flow in another line connected to the system, which can lead to amplification of the original surge if the effect of stoppage of this induced flow is added to the original pressure surge.

The interaction of flows and pressures with pumps, line friction and other modifying factors make it impractical to calculate anything but the most simple systems without computer programs such as FLOWMASTER. However, hand calculations can be useful to check if a system has the potential to produce surge pressures in excess of those the line can withstand using the equation below:

$$P = -\rho a \Delta V$$

where:

P = pressure rise
ΔV = reduction in flow velocity
a = speed of pressure wave
ρ = fluid density

If the fluid is stopped instantaneously the maximum head or so called 'Jacowski' pressure rise results. In a practical system the maximum head may be generated if the fluid is brought to rest within the pipeline period. For example, consider a pipeline discharging through an outlet valve which closes fully. The full Jacowski pressure rise occurs if the valve closes within the time taken for the pressure wave to travel to the other end of the pipeline and back. This is the pipeline period and is given by:

$$\tau = 2L / a$$

where:

τ = period
L = length of the pipeline

This method may be used to check the potential for overpressuring the pipeline, but may not give the maximum pressure if networks are involved due to amplification at the dead ends.

There are several operating changes that can cause flow changes and hence pressure transients. The most common are valve opening/closing and pump start/stop.

In the valve closure example the shape of the pressure/time variation will depend on the flow/time change which caused it and thus on the characteristics of the valve. Typical flow characteristics of ball, globe, and gate valves are shown in Figure 7.42. It is important to note that the last 10% of the valve closure has the most significant effect.

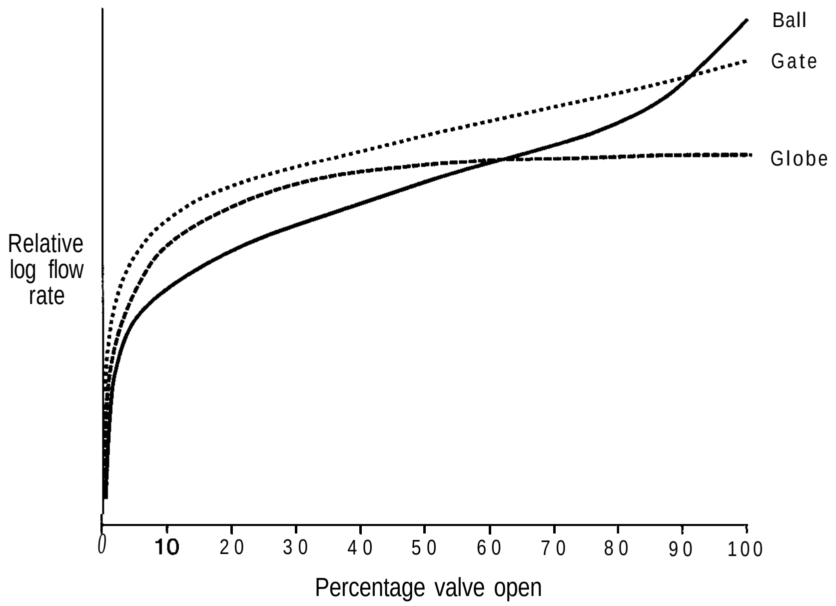

Figure 7.42 Relative flowrate against percentage valve opening for typical gate, ball, and globe valves

If the pipeline period is greater than the valve movement time then the type of valve characteristic will not influence the total pressure rise, although the rate of pressure rise will be important to any control devices and the mechanical load on supports. If the period is less than the valve movement time then the shape of the characteristic will influence the peak pressure. Special care should be exercised in specifying valves whose effective closure time is only a fraction of the total movement, eg. gate valves.

Pressure falls can be as damaging as pressure rises. Often the negative head changes lead to theoretical pressures less than the fluid vapour pressure. In such cases a vapour cavity forms. Large diameter lines are more susceptible to damage by reduced pressure conditions. More commonly vapour cavities create damage by their collapse. This occurs if there is a situation which can re-pressure the line. This could be caused by refilling from an elevated section of line or tanks or by re-starting a pump to continue operation. Often this refilling occurs at a high rate because the pump discharge pressure is lower than normal. When the cavity volume reaches zero the effect is of a truly instantaneous valve closure. These positive pressures can be higher than with normal valve closure and have step pressure changes. Therefore, they should be avoided if at all possible.

Pump start/stop has the capability of producing similar positive and negative pressure waves with the attendant damage potential.

From the above, it becomes apparent that the significant factors in creating high surge pressures are the wave speed and the rate of change of flow. The wave speed is fixed by the fluids used, although it is sometimes possible to modify the rate of change of velocity in a transient. This is particularly true of the effective valve closure time. Even valves with the most suitable closure characteristics create most of the flow change over the last quarter of their movement. Thus, two part closures, where the first part is fast and the last part slow can be beneficial.

Where surge is a problem, it is particularly important to avoid smaller bore area ends. When the surge pressure reaches a pipe junction it propogates along both lines. The pressure reduces in ratio to the new area divided by the old area and induces a flow by the pressure difference. At the instant of the wave arriving at the dead end the forward velocity is stopped creating another surge effect on top of the pressure wave. This can lead to a potential doubling of the pressure.

One solution to reduce surge pressures is to introduce surge accumulators into the system. This is particularly effective with pump generated surges. These accumulators are vessels which contain a gas pocket connected as closely as possible upstream of the surge generation device so that when the surge occurs, the excess fluid flows into the vessel compressing the gas. By removing excess fluid (which is nearly incompressible) the line pressure is reduced. The volumes and costs of such suppression equipment can be quite small when the transients are rapid.

7.5.2. Wave speed in single phase flow

Obviously the calculation of the speed of pressure waves in the fluid medium is crucial to the determination of the pressure surge. For a single phase system the pressure wave velocity is given by:

$$\alpha = (K/\rho)^{1/2}$$

where:

K = bulk modulus of elasticity
ρ = density of the fluid

Hence for water at 15 °C, K = 2.15 GN/m^2 and p = 1000 kg/m^3 the speed of pressure waves is 1466 m/s. For an ideal gas the bulk modulus is the ratio of the change in the pressure to the fractional change in volume and hence depends on whether the compression is adiabatic or isothermal. For an adiabatic process PV = γ constant and hence K = γP. Since P/p = RT the expression for adiabatic compression of an ideal gas is:

$$a = (K/\rho)^{1/2} = (\gamma RT)^{1/2}$$

For air at 15 °C, $\gamma = 1.4$, R = 287 J/kg/K hence the wave speed is 340 m/s. The equivalent bulk modulus of elasticity is 141.7 KN/m^2 and the density is 1.226 kg/m^2.

The pressure wave velocity is modified when boundaries are present such as pipe walls, free surfaces, gas bubbles and solid particles. The elasticity of the walls of the pipe through which the fluid is travelling has the effect of reducing the pressure wave speed by a factor dependent upon the size, cross-section shape, and pipe material. In addition the method of pipe restraint may also affect the result. The general equation for the wave speed in a fluid contained within a thin wailed pipe of circular cross-section is:

$$a = 1 / [p (l/K + D\Phi/tE)]^{1/2}$$

where:

D = internal diameter
t = wall thickness
Φ = pipe restraint factor

For materials with a high elastic modulus such as steel or concrete, or for pipelines with frequent expansion joints, Φ can be taken as unity. Hence for water at 15°C in a steel pipeline ($E=207\ GN/m^2$) of 200 mm diameter and 15 mm wall thickness, the wave speed is reduced from 1466 m/s for rigid pipewalls to 1374 m/s.

7.5.3. Wavespeedintwo-phase flow

The wave speed in two-phase flow can be determined by simply replacing the single phase bulk modulus and density with the two-phase equivalents, ie:

$$a = (K_{tp} / \rho_{tp})^{1/2} \text{ for rigid pipewalls}$$

and:

$$a = 1 / [\rho_{tp} (1/K_{tp} + D\Phi tE)]^{1/2}$$

where:

$$\frac{1}{K_{tp}} = \frac{(1-a)}{K_L} + \frac{\alpha}{K_g}$$

and:

$$\rho_{tp} = (1 - \alpha)\, \rho_L + \alpha \rho_g$$

α = in-situ gas fraction

Using air and water as an example it is seen that the wave speed is rapidly reduced by the introduction of only small quantities of air. For example, a void fraction of only 1% is required to reduce the wave speed from 1446 m/s to 119 m/s. The table below shows the wave speed for various gas fractions, also shown is the relative Jacowski head rise possible compared with the liquid only case. It is hence seen that in two-phase flow the wave speed and potential surge is much reduced by the combined high compressibility and high density of the two-phase mixture.

Void fraction (-)	Two-phase bulk modulus (N/m²)	Two-phase density (kg/m²)	Wave speed (m/s)	% potential head rise (-)
0.000	2.19e+9	1000.0	1466.0	100.0
0.005	2.80e+7	995.0	168.0	11.4
0.01	1.41 e+7	990.0	119.0	8.0
0.1	1.42e+6	900.1	37.7	2.4
0.2	7.08e+5	800.2	29.7	1.62
0.4	3.54e+5	600.2	24.3	1 .oo
0.5	2.83e+5	500.6	23.8	0.81
0.6	2.36e+5	400.7	24.3	0.66
0.8	1.77e+5	201 .o	29.7	0.41
0.9	1.57e+5	101.0	39.4	0.27
0.99	1.43e+5	11.2	113.0	0.07
0.995	1.42e+5	6.2	151.3	0.06
1 .o	1.42e+5	1.2	340.0	0.03

The wave speed is plotted as a function of the void fraction in Figure 7.43.

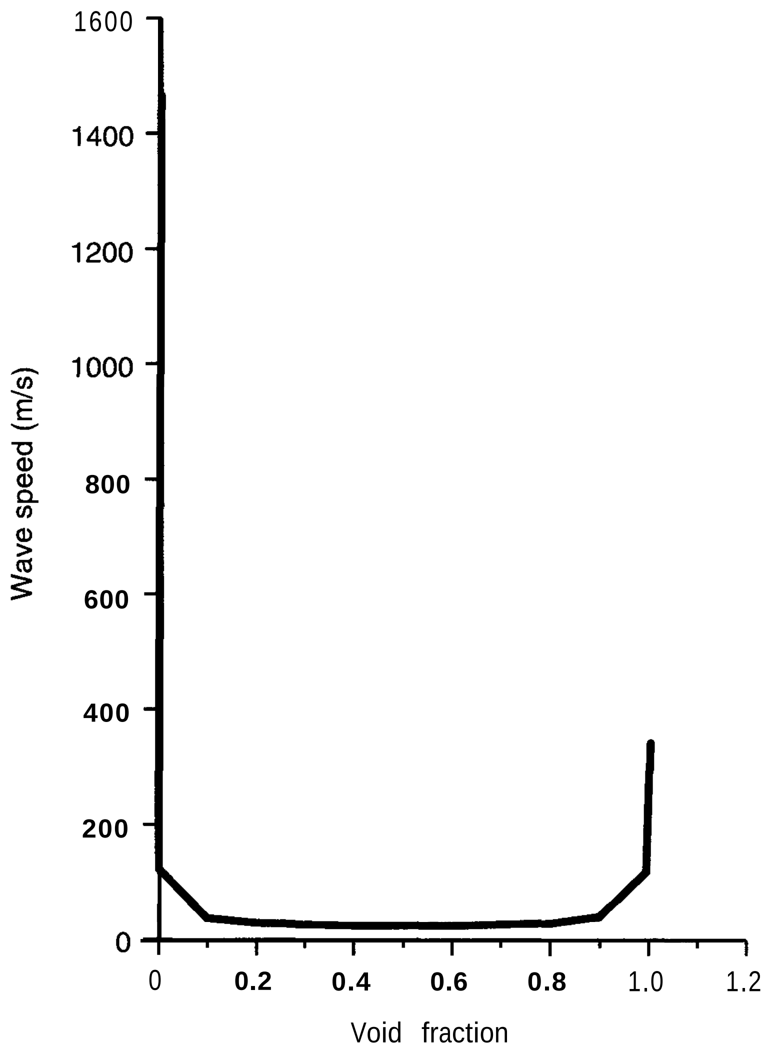

Figure 7.43 Pressure wave speed in two-phase flow

7.5.4. Two-phase pressure surge analysis using FLOWMASTER

Although FLOWMASTER is essentially a single phase hydraulic surge computer program, there are a number of features that may be used to investigate two-phase surge effects. Inherent in the code is the ability to form vapour pockets if the pressure falls below the vapour pressure specified for the fluid. If the pressure subsequently rises the collapse of the cavity can also be modelled. FLOWMASTER simulates a practical system as a combination of modules that are connected together to form a network. The modules or components represent pipes, valves, pumps etc and provide the equations for specific applications (for example valve loss coefficient as a function of opening). There are two components available in FLOWMASTER that are useful for two-phase flow surge calculations. These are the 'bubbly pipe' and 'primer' components.

Bubbly pipe component

The bubbly pipe component allows the user to model the wave speed in an elastic pipe with a small fraction of free gas present (the order of 1% void fraction) which results in a variable wavespeed throughout the pipe. The gauge vapour head and the wave speed in the pure liquid are specified. The rate at which gas is released from solution is also controlled by a gas release factor which is a function of the per unit difference between dissolved and equilibrium gas masses. This component is hence restricted to homogeneous bubbly type flows with small gas fractions, but may be useful to determine loads experienced as safety valves are closed on a volatile oil pipeline during a rupture, for example.

Primer component

The primer component is used to analyse the transient behaviour associated with the priming of a part-empty pipe, and was initially developed to study the impact loads and pressures created when a dry fire sprinkler system is initially primed with water. These loads are created when the liquid impacts on the sprinkler head restriction, and is similar to a slug impact. For two-phase slug flow analysis steady state or transient analysis should be carried out using other software to determine the length and velocity of slugs at the actual pipeline conditions, and the primer component in FLOWMASTER used separately to determine slug impact loads.

The primer is a length of pipe with a restriction at each end, and the wave speed in the liquid is specified as the initial fractional gas length in the pipe. The inlet pressure drives the liquid along the pipe like a piston, compressing the gas and forcing it through the outlet restriction until the liquid front reaches the end and the surge pressure is generated. By changing the speed of sound in the liquid and the density it is possible to consider the sensitivity of the resulting pressure rise to the gas entrainment fraction of the slug.

Reference

7.1 Gregory, G.A and Fogarasi, M "Estimation of pressure drop in two-phase oil-gas looped pipeline systems", Pipeline Technology, March-April 1982, pp75-81.

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Example 7.1 PLAC simulation of shutdown and restart of Pompano subsea wells

Pompano is a development in the Gulf of Mexico consisting of a fixed platform in 1300 ft of water, with a proposed remote 18 slot template located 4 miles away in a water depth of 1800 ft . The template was originally going to be connected to the host platform by two 6" pipelines allowing flexible operation as the flowrates change, and facilitating pigging from the platform using processed oil. The first phase of the development, now completed, involved the installation of the platform and start-up of the platform wells, which are expected to produce a maximum of around 18 mbd. The second phase of the development will involve installation of a subsea template and flowlines. However it was required to make provision for this during the installation of the first phase. The maximum oil production from the subsea wells was expected to be 12 mbd and slug catchers were required to be sized for the platform reception facilities.

The slugs produced by pigging the pipeline were to be handled by operational procedures and were not used for sizing the slug catcher. The predicted mean and maximum normal hydrodynamic slugs were 6 bbls and 16 bbls respectively, and were well within the feasible slug catcher size, which was in the region 50-I 00 bbls. It was hence necessary to consider the slugs produced by flowline rate changes which could be quite frequent as wells are switched to test and as turndowns are accommodated. A worst case was considered to be start-up from 1 well to full production, giving a flowrate increase from I-I 2 mbd. This was modelled with PLAC by running the simulation at 12 mbd for 13000s to give a steady state. The flowrate was then ramped down to 1 mbd over 60s, left at this rate for 5000s, then increased to 12 mbd over 60s. The predicted outlet liquid flowrate profile is shown in Figure 1 and indicates that the liquid flowrate drops off when the flowrate is reduced after 13000s. The plot shows a large overshoot when the rate is increased again at 18000s. The oscillations decay as the final equilibrium production rate of 12 mbd (equivalent to 17.7 kg/s) is approached. The peak production rate is 82 kg/s which is equivalent to 51 mbd, and is hence over four times the final equilibrium production rate.

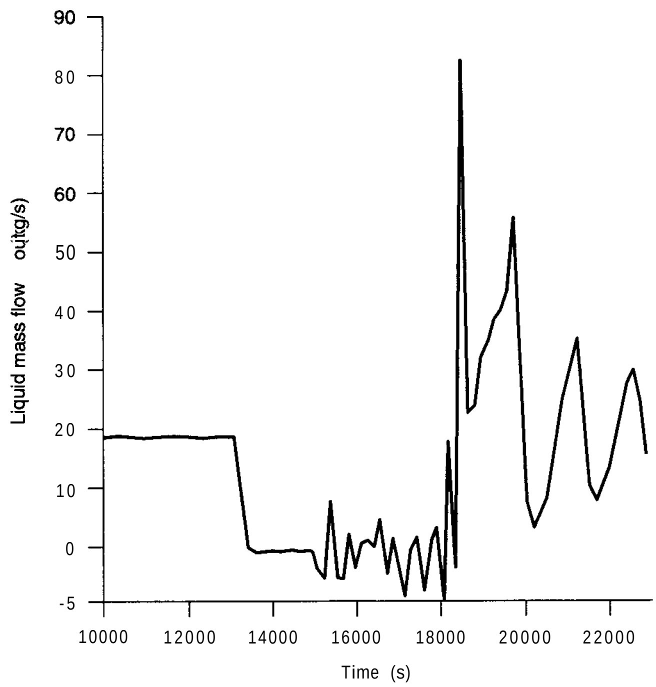


Figure 1. PLAC simulation of outlet liquid flowrate

The outlet gas rate is shown in Figure 2 and indicates that the main gas surge occurs after the second slug.

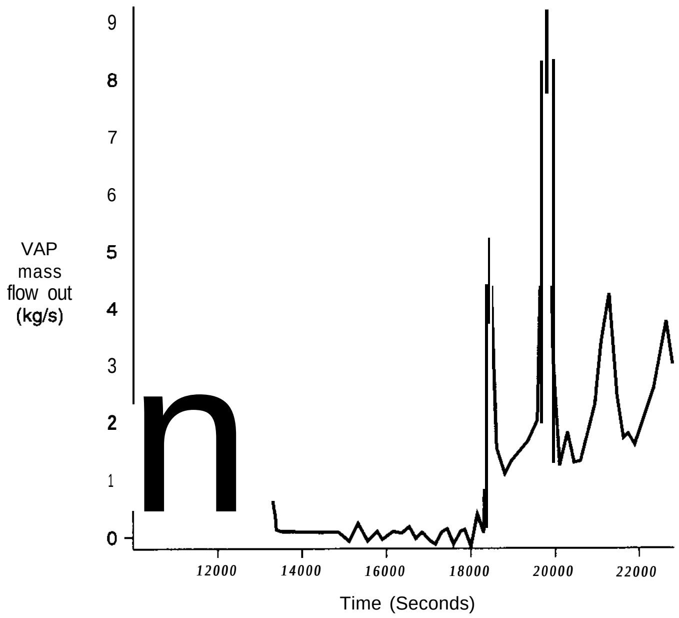


Figure 2. Predicted outlet gas flowrate

The liquid holdup plot in Figure 3, shows that the initial holdup is around 40% in the inclined flowline (cells l-l 8) and around 20% in the vertical riser (cells 19-39). During the low flow condition from 13000s the liquid drains from the riser into the flowline and also builds up at the template end of the flowline. During the rate increase the first slugs were due to the liquid drained from the riser. The liquid that drained to the template is smeared out along the flowline to re-establish the equilibrium holdup.

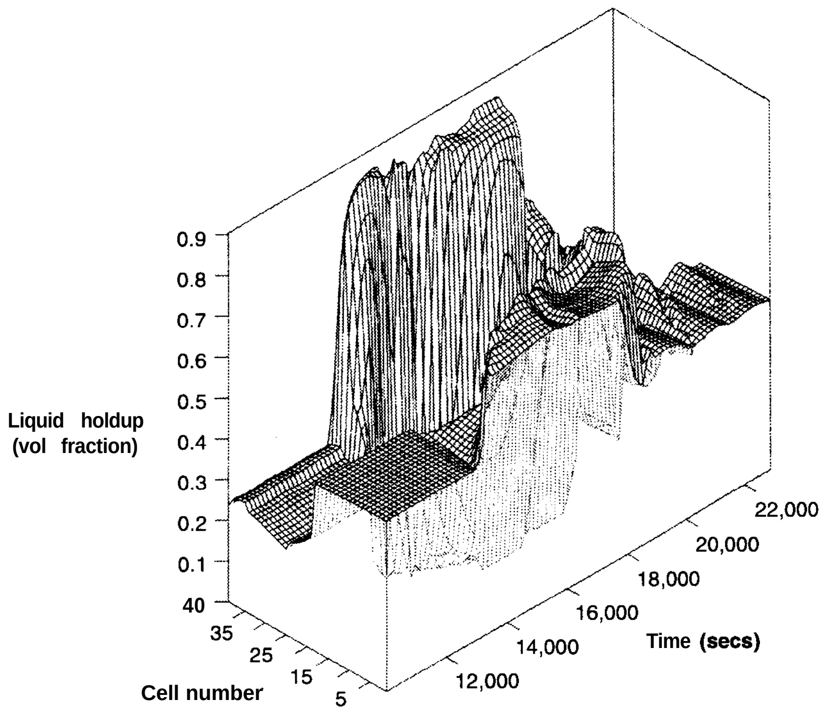


Figure 3. Holdup profile during shutdown and start-up

The outlet liquid flowrate predicted by PLAC was converted to a spreadsheet for further processing by a slug catcher simulation program called PLACSEP, which is available on the PC. This program converts the liquid volume flowrate into the level fluctuations in a cylindrical slug catcher vessel. Figure 4 shows the level for a 50 bbl vessel (5ft dia x 15ft long) where the liquid pump-out rate is fixed at 25 mbd. In one case there is no control on the minimum level in the vessel, and in the other a minimum level of 20% is imposed.

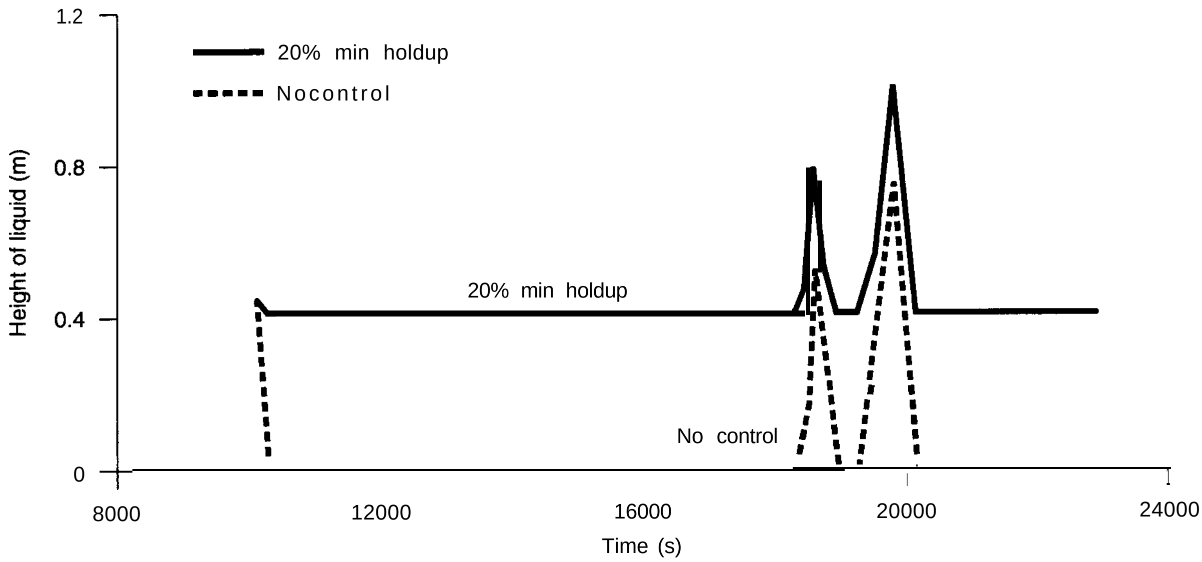


Figure 4. Slug catcher level fluctuations with and without level control

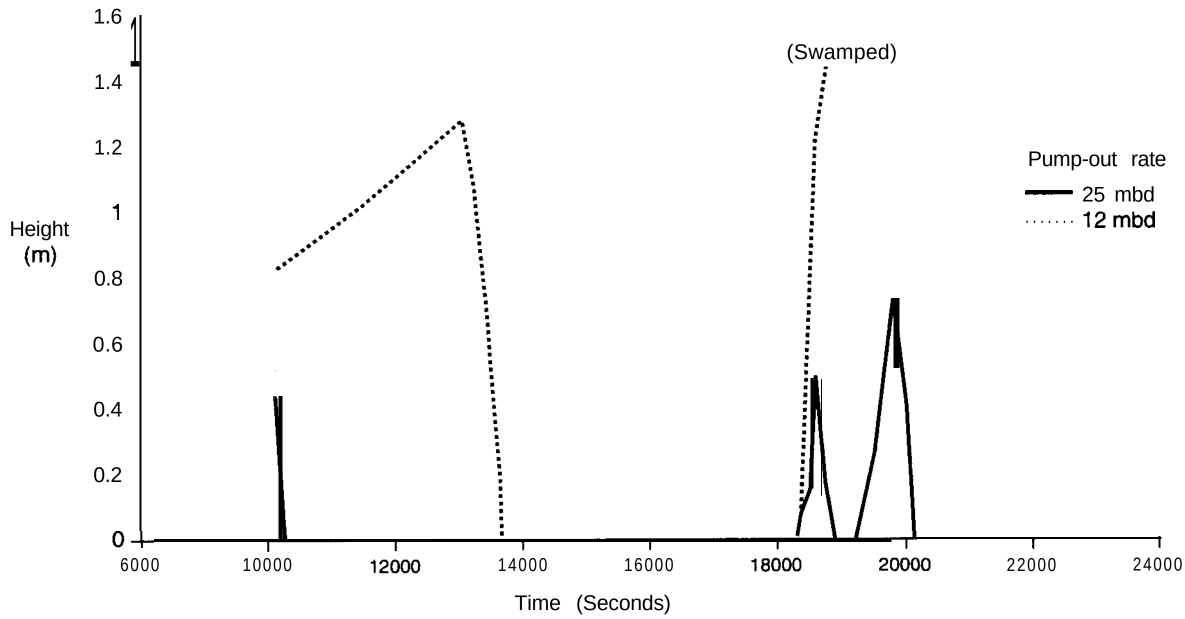
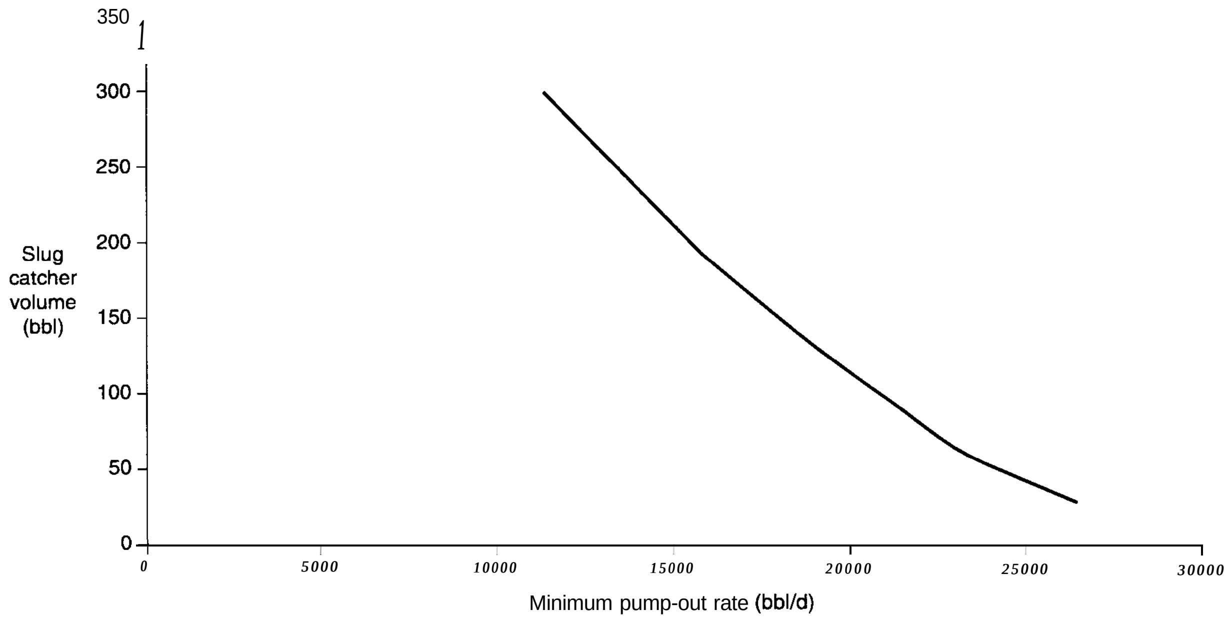


Figure 5. Effect of pump-out rate on slug catcher level

The result is also plotted in Figure 5 for a pump-out rate of 12 mbd and shows that the vessel would be swamped, even without low level control. Hence it is not possible to use a 50 bbl slug catcher if the liquid pumpout rate remains fixed at the maximum subsea production rate.

It is possible by iteration to determine the vessel size vs pump-out rate relationship that just handles the inlet flow transient. This is shown in Figure 6 and illustrates that if the pump-out rate is to remain at the normal subsea production rate of 12 mbd, then a 275 bbl slug catcher would be required, but if the spare capacity of the platform process is utilised to give a pump-out rate of 25 mbd, then the required slug catcher volume is around 40 bbls.



Ratio of L : D taken as 4 : 1

Figure 6. Predicted slug catcher volume vs pump-out rate relationship

PLAC is now capable of simulating the slug catcher as well as the pipeline and can include relatively complex control functions. However it would be very time consuming to use PLAC in this way to produce the above relationship. It is hence recommended that the PLACSEP program be used for this purpose. If it is expected that the slug catcher control set-up will influence the pipeline transient, then the PLACSEP program should be used to guess the initial vessel size and pumpout rate, and a full simulation used to finalise the design. In the future it is expected that PLAC will be interfaced with a process plant dynamic model so that the full pipeline and process dynamics may be integrated. This facility is already available via the OLGA/D-SPICE interface.

Example 7.2 – Simple hand calculation method applied to Pompano example

The simple hand calculation method outlined in Section 7.3.1 can also be used to estimate the required vessel size and pump-out rate for the Pompano restart transient that was simulated by PLAC in Example 7.1.

The starting point for the hand calculation requires the steady state holdup in the flowline at the initial and final production rates. Figure 1 shows the liquid holdup vs flowrate predicted by MULTIFLO for the 6", 4 mile long Pompano subsea flowline.

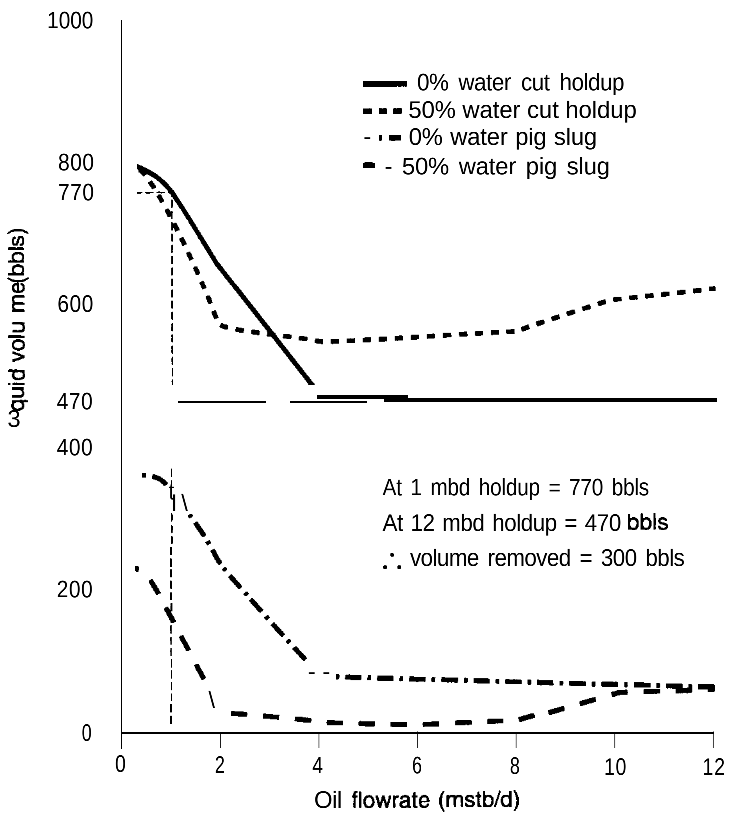


Figure 1. MULTIFLO holdup predictions for Pompano 6" subsea flowline

It is seen that for an initial rate of 1 mbd the equilibrium holdup is 770 bbls for 0% water cut, reducing to 470 bbls at a final flowrate of 12 mbd.

The residence time is approximated by the equilibrium holdup divided by the liquid flowrate, hence:

Residence time at 1 mbd = 770/1000 = 0.77 days
Residence time at 12 mbd = 470/1 2000 = 0.039 days

The transition flowrate is assumed to occur for a period equal to the residence time at the final condition and is equal to the final production rate plus the rate corresponding to the excess holdup swept-out during the period of the transition. Hence:

Transient Flowrate = 12000 + (770-470 / 0.039) = 19.692 mbd

The approximate outlet liquid flowrate history is illustrated in Figure 2 and is compared with the PLAC predictions.

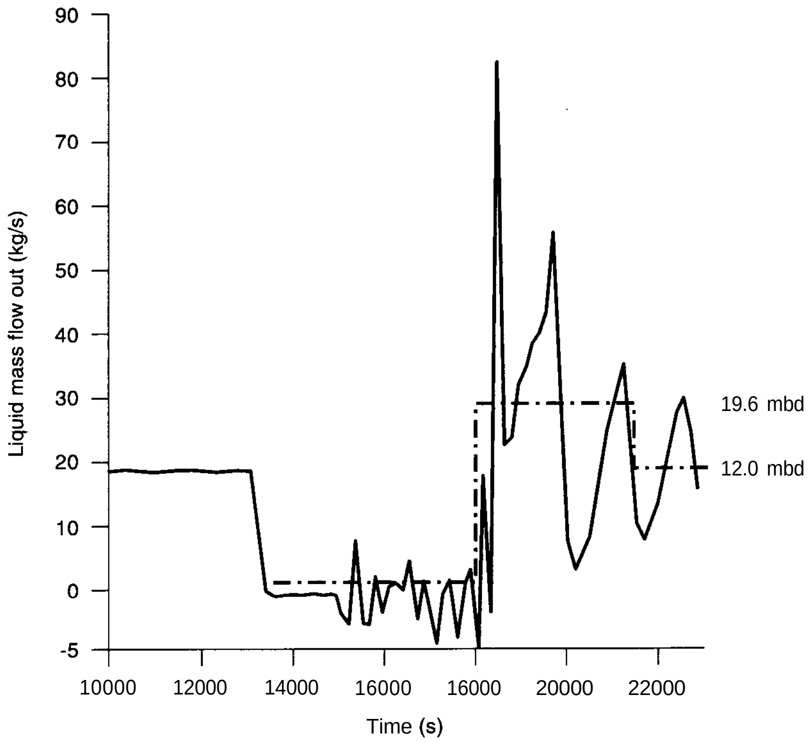


Figure 2. Approximate outlet liquid flowrate profile and PLAC prediction

The slug catcher volume vs pump-out relationship is hence given by a simple linear function since the flowrate during the transition is assumed to be constant, therefore:

Surge volume = 0.039 x (19692 - Qout)

hence:

at 0 mbd,	surge volume = 768 bbls
at 12 mbd,	surge volume = 300 bbls
at 19.692 mbd,	surge volume = 0 bbls

The comparison between the simple method and the rigorous PLAC simulation is shown in Figure 3 where it is seen that the agreement is close for pump-out rates similar to the final equilibrium value. However, there are large discrepancies at the extremes for the reasons outlined in Section 7.3.1. For zero surge volume the pump-out rate predicted by PLAC is 51 mbd compared to 19.7 mbd assuming a constant transition flowrate, hence the simple method fails to take account of the liquid distribution in the pipeline and consequently underestimates the peak outlet flowrates.

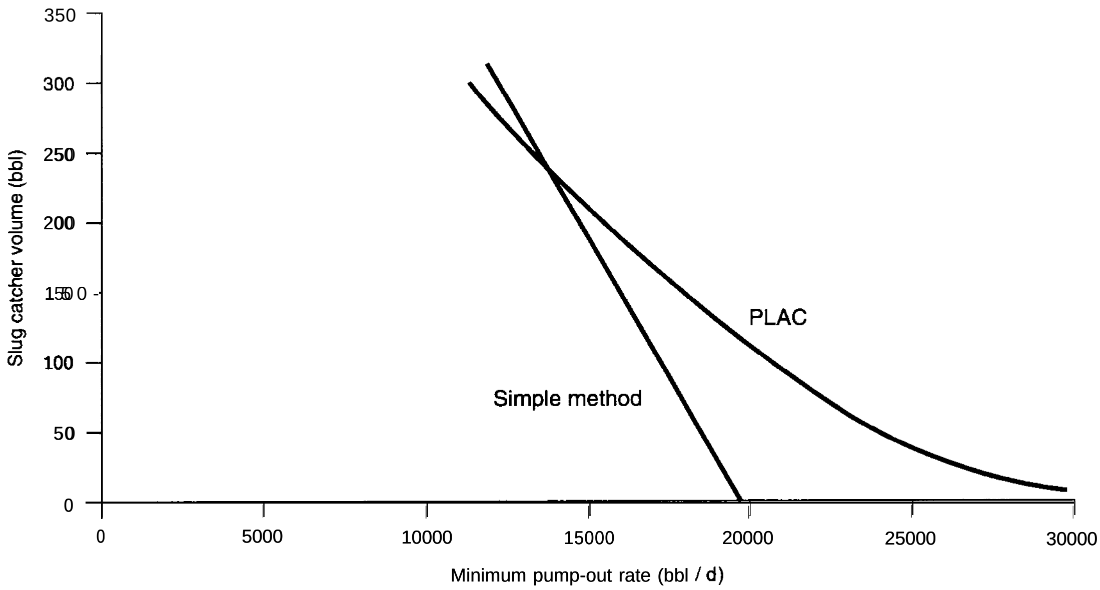


Figure 3. Comparison of slug catcher relationships derived from PLAC and hand calculations

Example 7.3 OLGA simulation of Miller landline shutdown and restart

The 30" Miller landline normally transports dense phase gas over 18 km from the St Fergus terminal to Peterhead power station. The gas undergoes dew point control and heating at St Fergus to remain single phase. Under normal operating conditions the gas flowrate is 215 mmscfd and leaves St Fergus at 37 deg C and 30 bara, arriving at Peterhead at 30 deg C and 25 bara. During shut-down and stat--up there is the possibility of liquid condensation and drop-out which may affect the size of the downstream plant and operating procedures necessary to control the liquid outflow. The combination of rate changes and associated cool-down gives rise to a more complicated transient analysis than previously discussed.

In such a transient the amount of liquid formed is a function of the richness of the input gas composition, the temperature gradient along the pipeline, Joule-Thompson effects, transient cool-down, the topography, and pressure drops. Some of the interactions between these effects can be complicated. For example, the pressure drop can cause retrograde condensation which can give rise to higher pressure drops due to interphase friction. This can lead to J-T cooling which produces more liquid dropout, and so on. Once liquid is produced it may take time to drain into dips, and may produce slugs. On start-up the the liquid can re-evaporate in the pipeline before reaching the outlet.

The OLGA code was used to simulate the effects caused by the shut-down and consequent cooling of the Miller landline. The start-up was also modelled in order to study the complex transient multiphase effects taking place throughout the pipeline system as a result of the introduction of warm gas into the cool environment. This study formed part of the design stage of the pipeline project and provided a valuable insight into the sizing requirements for slug-catching equipment at the power station delivery point.

The transient simulations were performed in three stages:

1. **Run to steady state at 215 mmscfd with an inlet pressure of 30 bara and an inlet temperature of 37°C.**

2. **Shut-down with the outlet closed and the line allowed to pack up to 35 bara through out. The line is subsequently allowed to cool for 48 hours.**

3. **Start-up the pipeline with inflow and outflow of 60 mmscfd and remain constant for 13 hours. The inlet flowing gas temperature was the order of 80°C. Finally, the flowrate is ramped to 215 mmscfd with an inlet temperature of 37°C and the simulation continued until the outlet temperature exceeds 30°C.**

The results of the simulation are best described with reference to the pipeline profile and the relevant cell numbers illustrated in Figure 1.

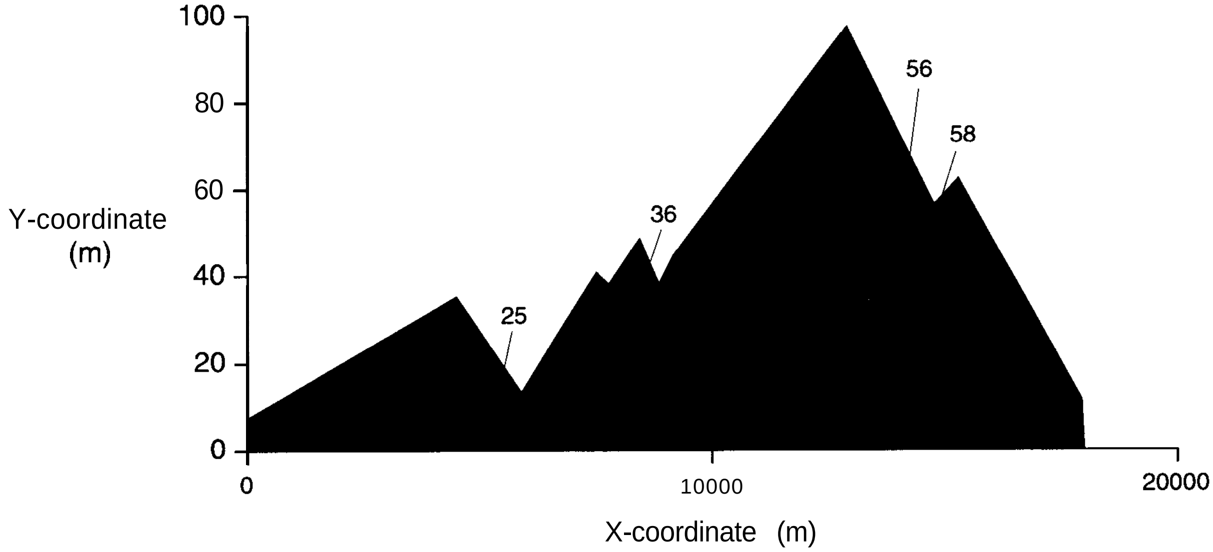


Figure 1. Miller landline profile

Steady state results gave pipeline outlet conditions of 27 bara and 30.4%. During the shut-down the pressure quickly packs up to 35 bara and the temperature gradually drops, taking 48 hours to drop to 6 Deg C at the outlet. This is illustrated in Figure 2. The ground temperature is -3°C. During the shut-down the liquid condenses and drains into the dips giving rise to an increase in the pipeline liquid content, which still rises after 48 hours as a result of the cooldown (Figure 3).

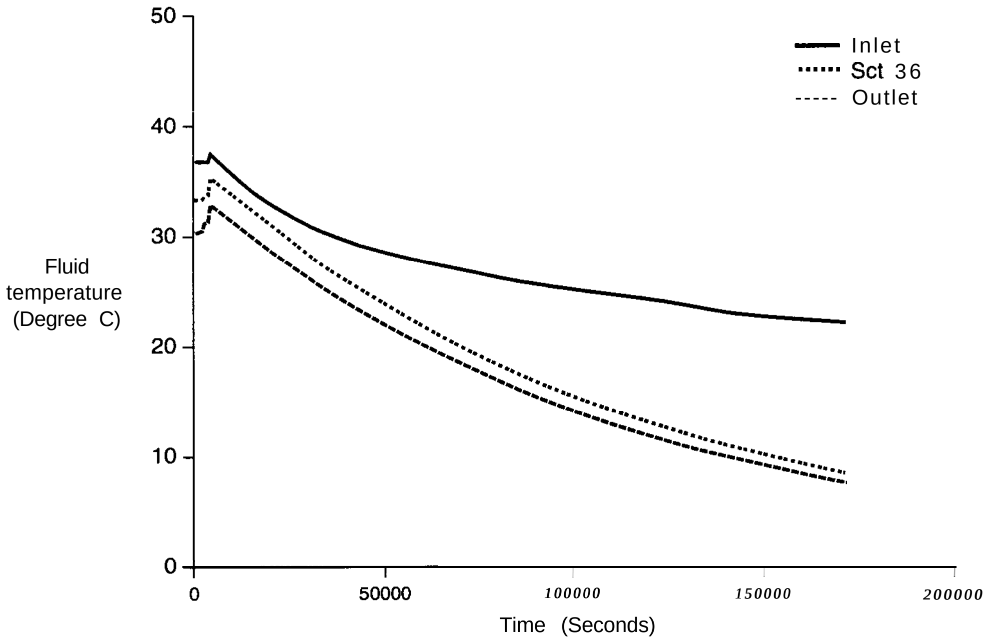


Figure 2. Variation of shut-in temperature with time

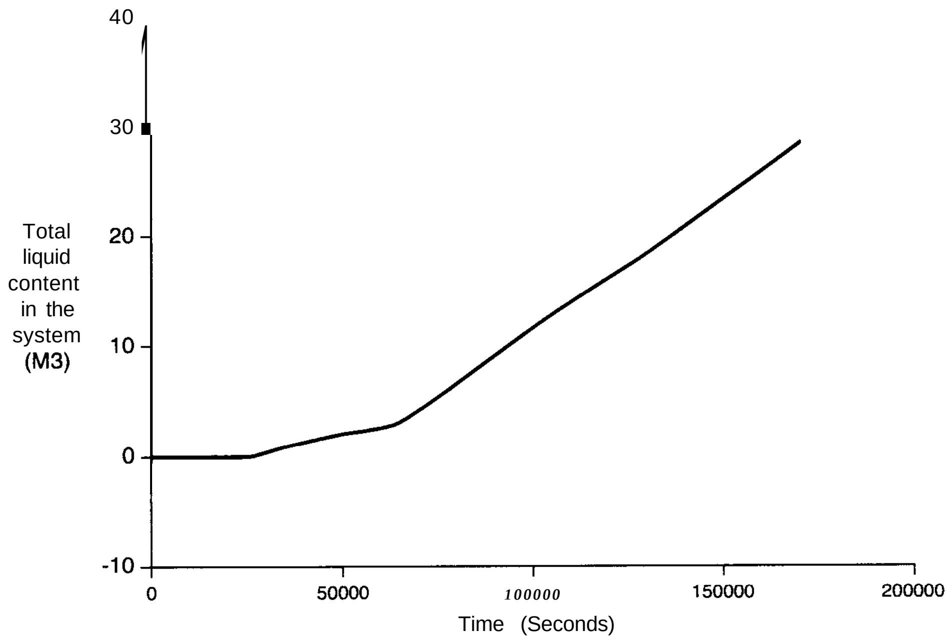


Figure 3. Pipeline liquid content during shut-in

In the first phase of the start-up the gas flowrate is ramped upto 60 mmscfd with an inlet temperature of 80°C. This has the effect of reducing the downstream pressure as the hydraulic gradient is established, and warming the fluid temperature. The warm-up is seen to be slow as the outlet gas temperature has only risen by a few degrees after 13 hours (Figure 4). The slow warm-up and flowrate increase produces liquid sweepout at some of the dips. However the liquid fills subsequent dips and does not exit the pipeline, hence there is still a net increase in the pipeline liquid content during the first phase of the start-up (Figure 5).

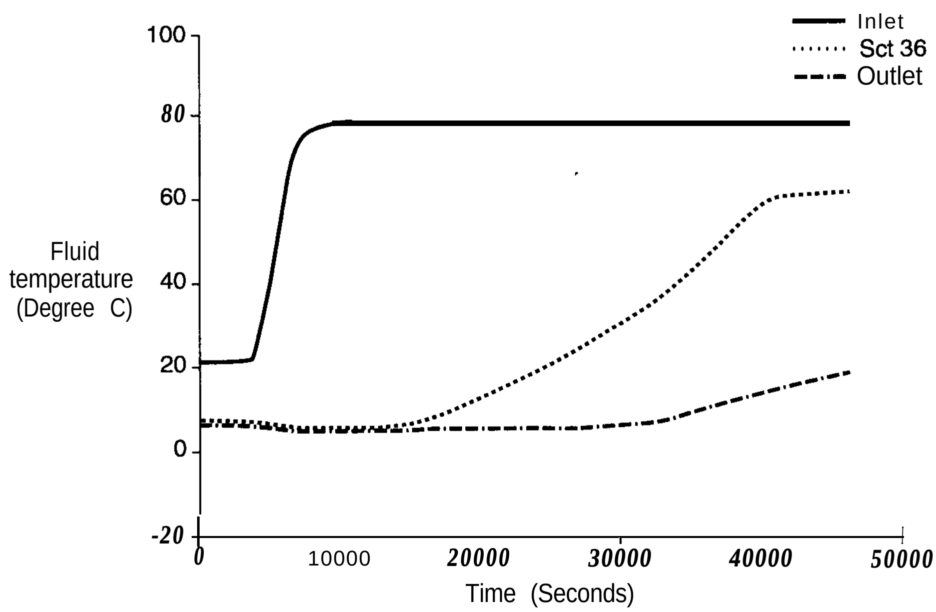


Figure 4. Temperature during first start-up phase

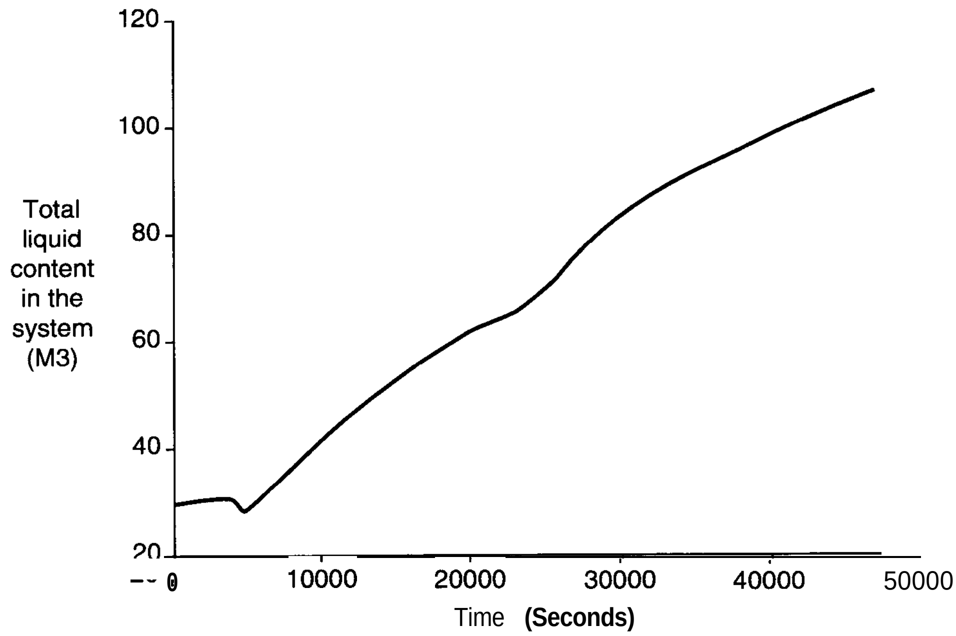


Figure 5. Pipeline liquid content during first start-up phase

In the second start-up phase the gas rate is increased to 215 mmscfd with the gas flowing in at 38°C and 35 bara. The further increase in the flowrate results in a higher pressure gradient and hence a decrease in the outlet pressure. The increase in the cold gas inflow rate causes the warm front to accelerate through the pipeline (Figure 6) causing liquid to be flashed off, and sweeps out the residual liquid from the system (Figure 7). As a result the total liquid content of the system drops sharply removing around 110 m^3 (690 bbls) of liquid from the pipeline (Figure 8), however around 30% of this liquid is evaporated and the remainder flows to the slug catcher.

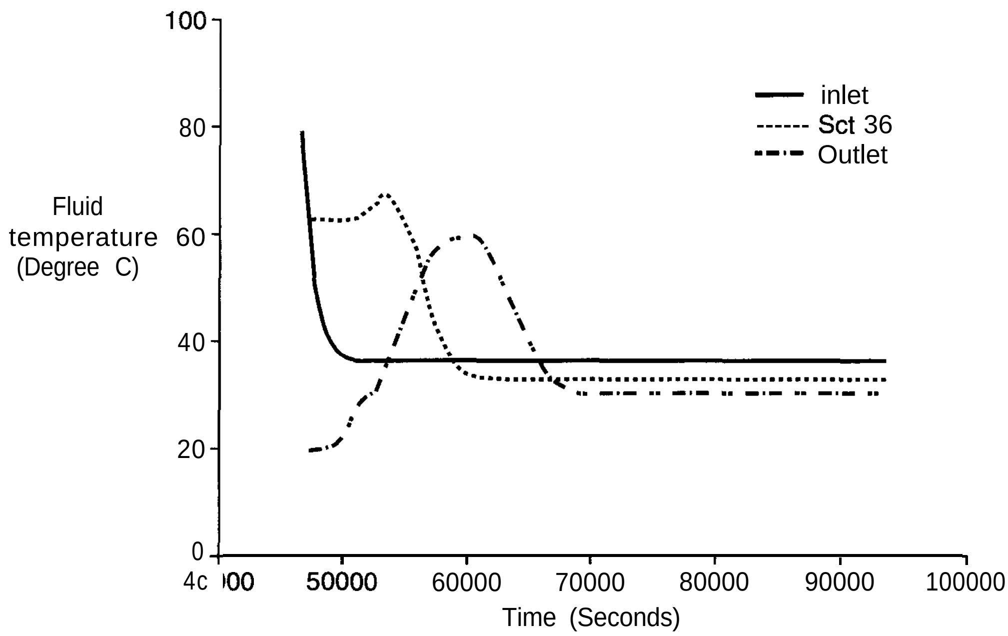


Figure 6. Temperature variation during second start-up phase

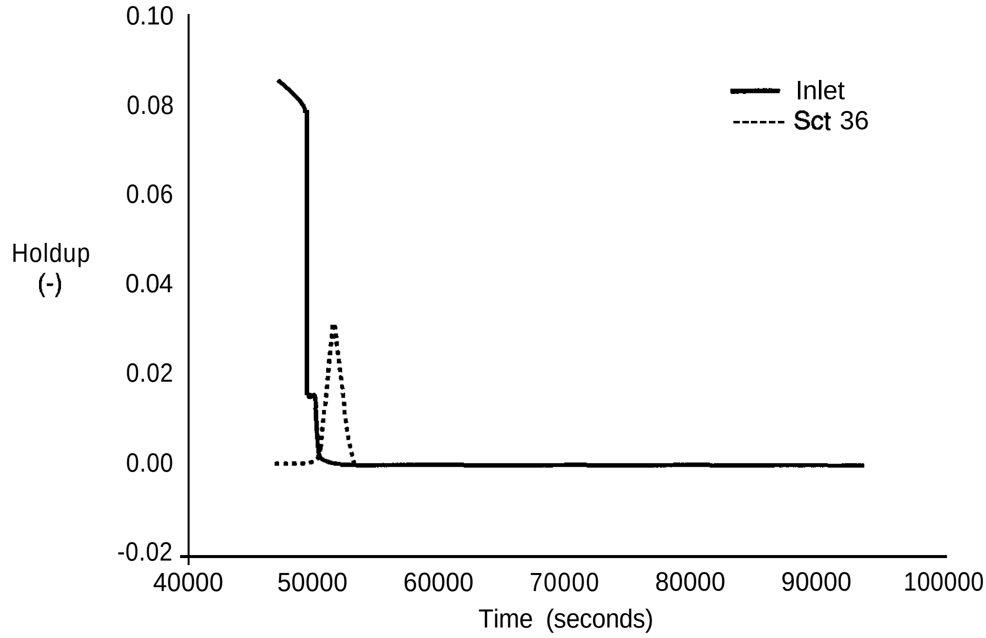


Figure 7. Liquid holdup during second start-up phase

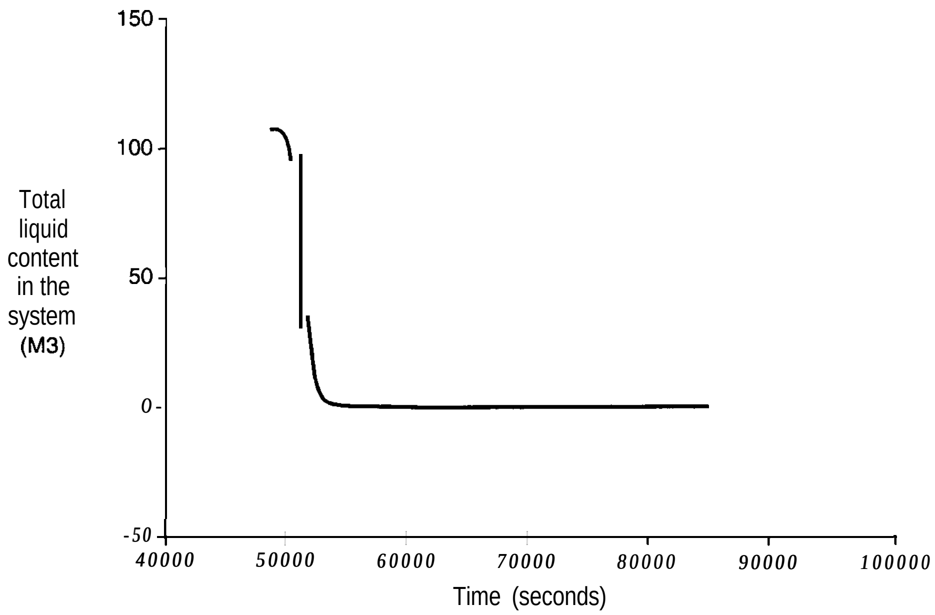


Figure 8. Pipeline liquid content during second start-up phase

Example 7.4 PLAC analysis of severe slugging in a catenary riser

A comparison has been made between PLAC and some experimental results of severe slugging in a Catenaty riser. The experimental data was taken by BHRA in 1990 (Reference 1) and involves holdup, pressure, and velocity measurements in a 2" diameter, 108 ft high catenary riser model using air and water as the test fluids. A 200 ft length of 2" downhill inclined line was used before the riser and an air buffer vessel of 0.126 m^3 was used in the air supply line to model a larger pipeline length. The test rig is illustrated in Figure 1.

PLAC simulations have been performed by XFE for a test condition just in the severe slugging region corresponding to gas and liquid superficial velocities of 2.3 m/s and 0.33 m/s respectively. The measured pressure at various points in the riser are shown in Figure 2 where the top traces are for the transducers in the inclined flowline and at the base of the riser. The experimental severe slugging cycle time is 183 seconds.

The test rig was modelled in PLAC as a TEE component with 86 cells, in which a 31 m horizontal section of pipe was used to simulate the air buffer vessel. The liquid is introduced into the side arm of the TEE, which is located at the end of the horizontal section. The topography is shown in Figure 3. The PLAC simulations begin with an empty pipe and hence some time is required for the severe slugging cycle to be established. Figure 4 shows the pressure in the inclined flowline which is in good agreement with the measured data of Figure 2.

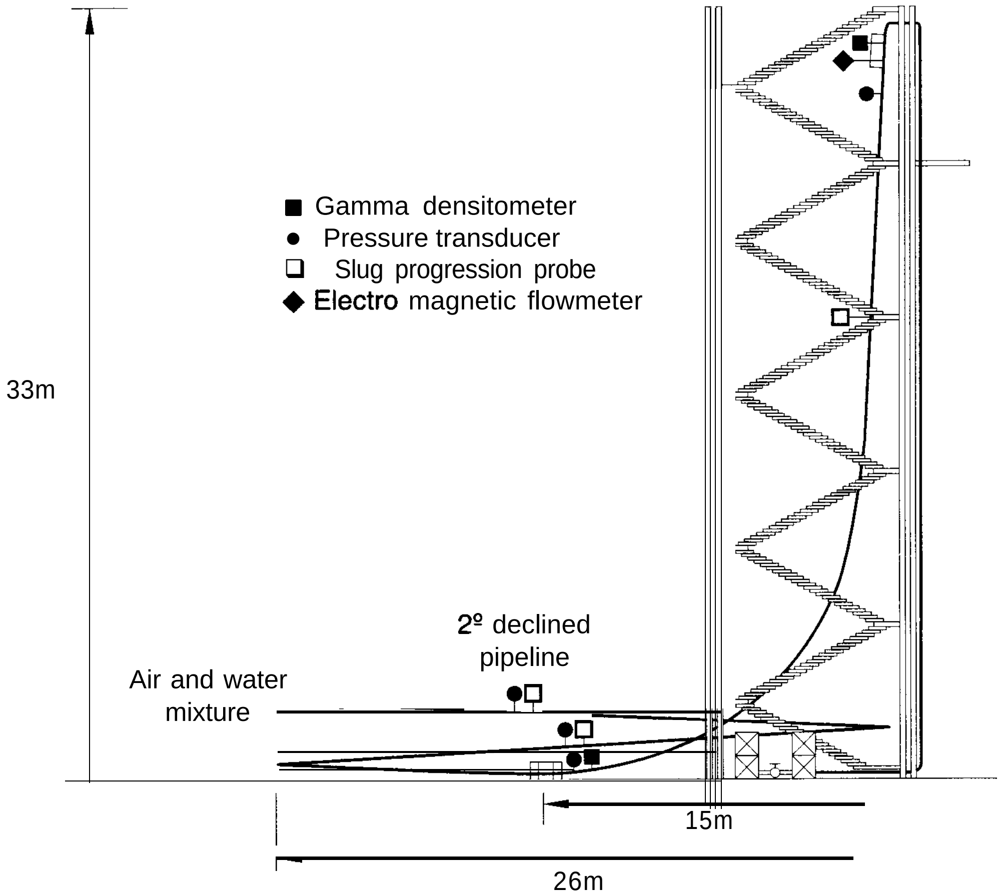


Figure 1. Schematic of experimental riser facility

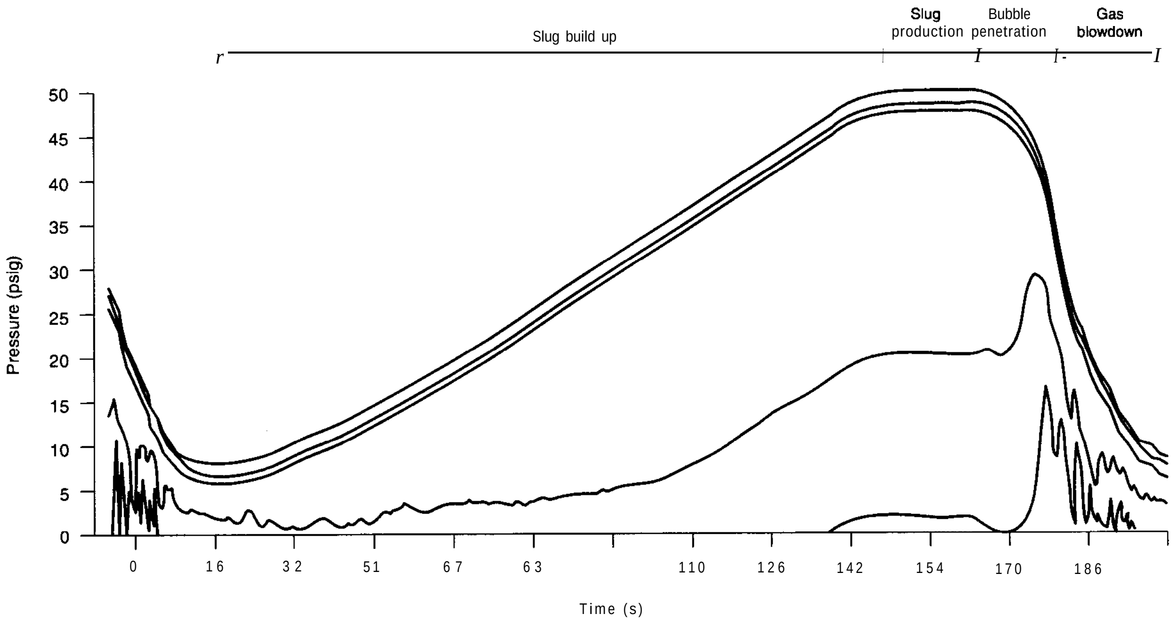


Figure 2. Pressure traces during one severe slugging cycle

The test rig was modelled in PLAC as a TEE component with 86 cells, in which a 31 m horizontal section of pipe was used to simulate the air buffer vessel. The liquid is introduced into the side arm of the TEE, which is located at the end of the horizontal section. The topography is shown in Figure 3. The PLAC simulations begin with an empty pipe and hence some time is required for the severe slugging cycle to be established. Figure 4 shows the pressure in the inclined flowline which is in good agreement with the measured data of Figure 2.

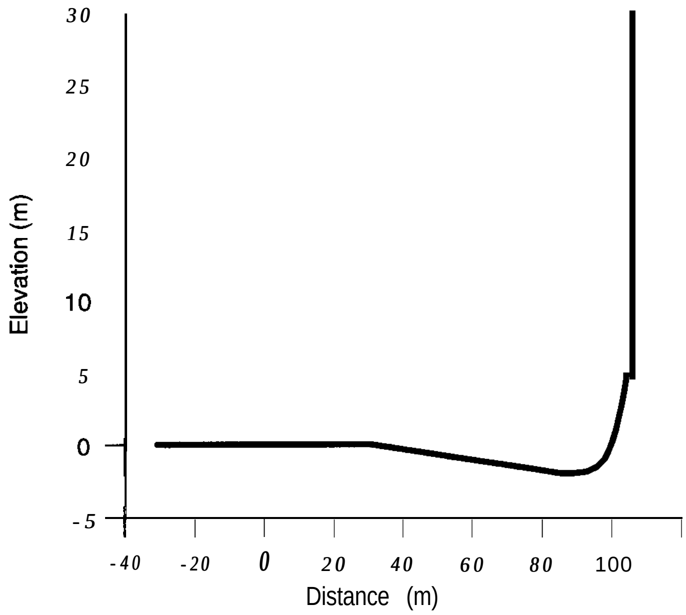


Figure 3. Experimental rig topography used in PLAC

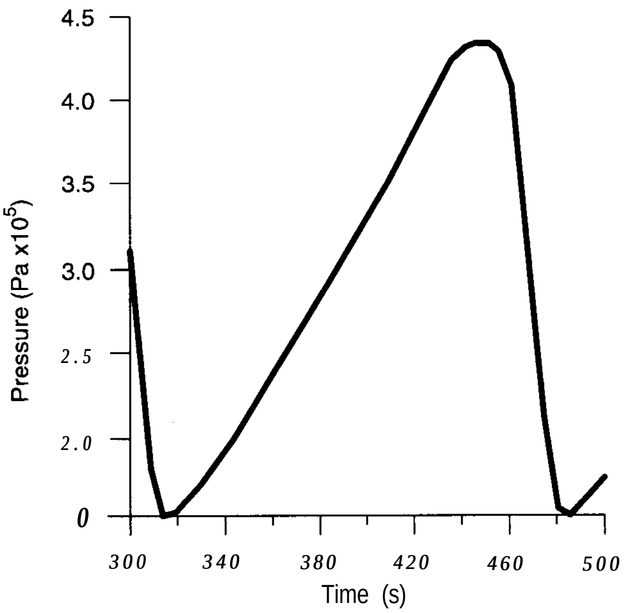


Figure 4. Pressure in inclined flowline predicted by PLAC.

Figures 5 to 7 illustrate the holdup and velocity fluctuations throughout the riser. Analysis of the PLAC results enables the following comparison to be made:

	Experimental measurements	PLAC predictions
Cycle time	**183s**	**172s**
Slug build-up time	**124s**	**127s**
Slug production time	**24 s**	**11 s**
Bubble penetration time	**14s**	**9 s**
Gas blowdown time	**21 s**	**25 s**
Slug length	**59 m**	**57 m**
Maximum pressure at riser base	**47.5 psig**	**48 psig**
Slug tail exit velocity	**3.5 m/s**	**4.5 m/s**

From the above comparison it is seen that in most cases the PLAC predictions are in good agreement with the measured values apart from the slug production time. This time is however relatively short and is difficult to accurately determine from the plots. The experimental conditions were not tabulated by BHRA and hence the estimation of the flowing velocities could also be subject to some error. However, the predictions indicate that the flowing conditions in the riser are close to the boundary for true severe slugging as the riser is only full of liquid for a short period of time before the gas pressure is sufficient to eject the slug. Figure 8 shows the test case point on the experimental flow pattern map and confirms that the PLAC predictions are qualitatively correct.

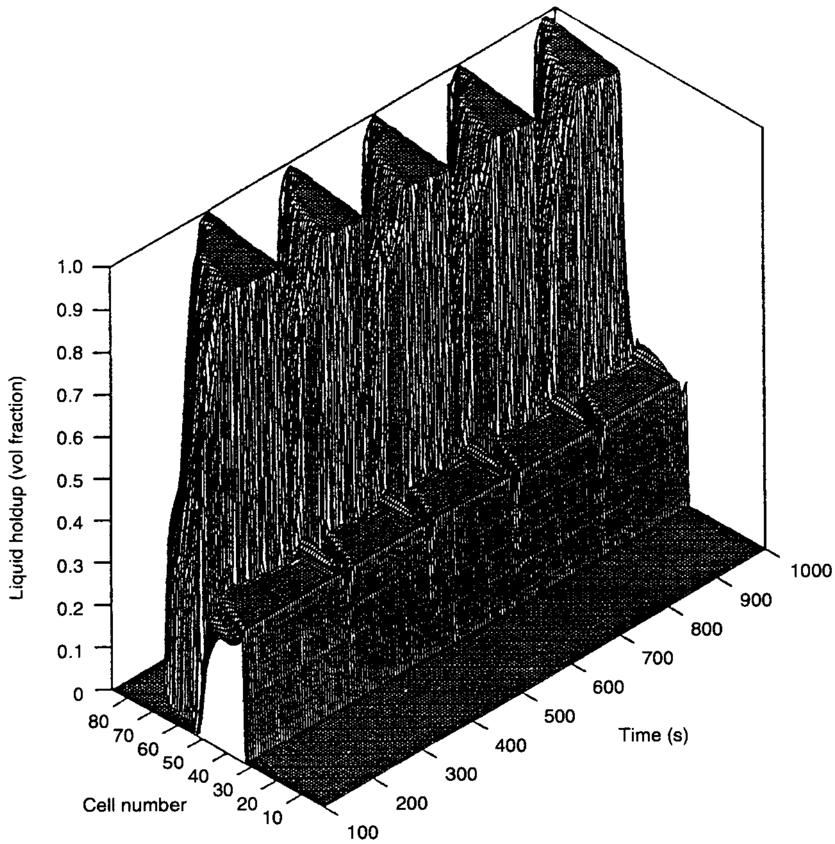


Figure 5. Holdup variation during severe slugging

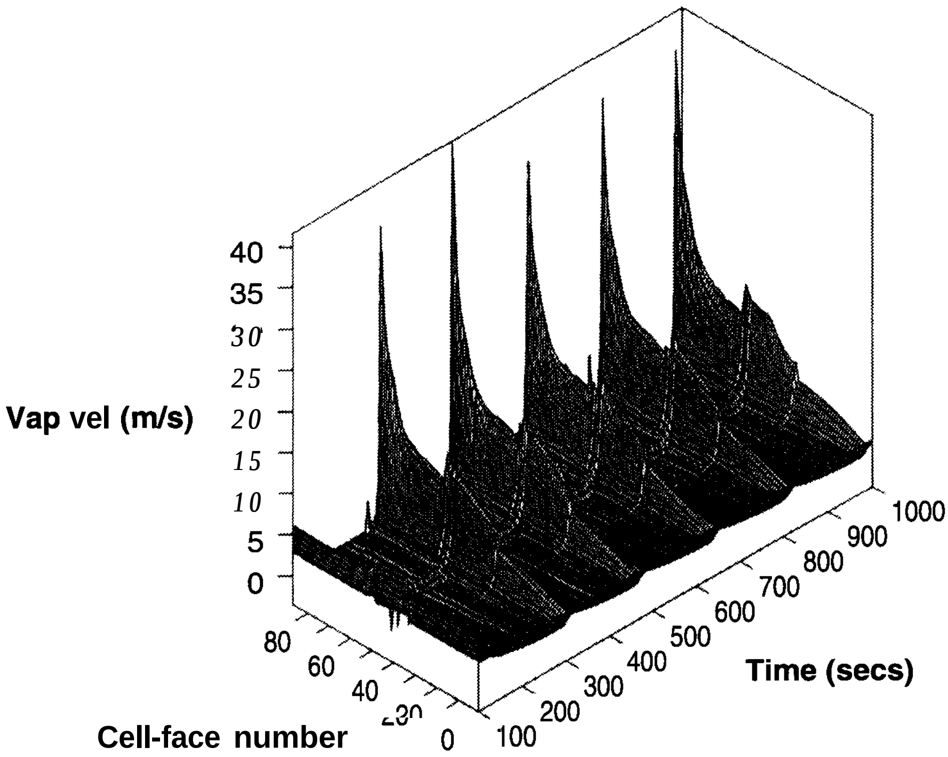


Figure 6. Vapour velocity variation

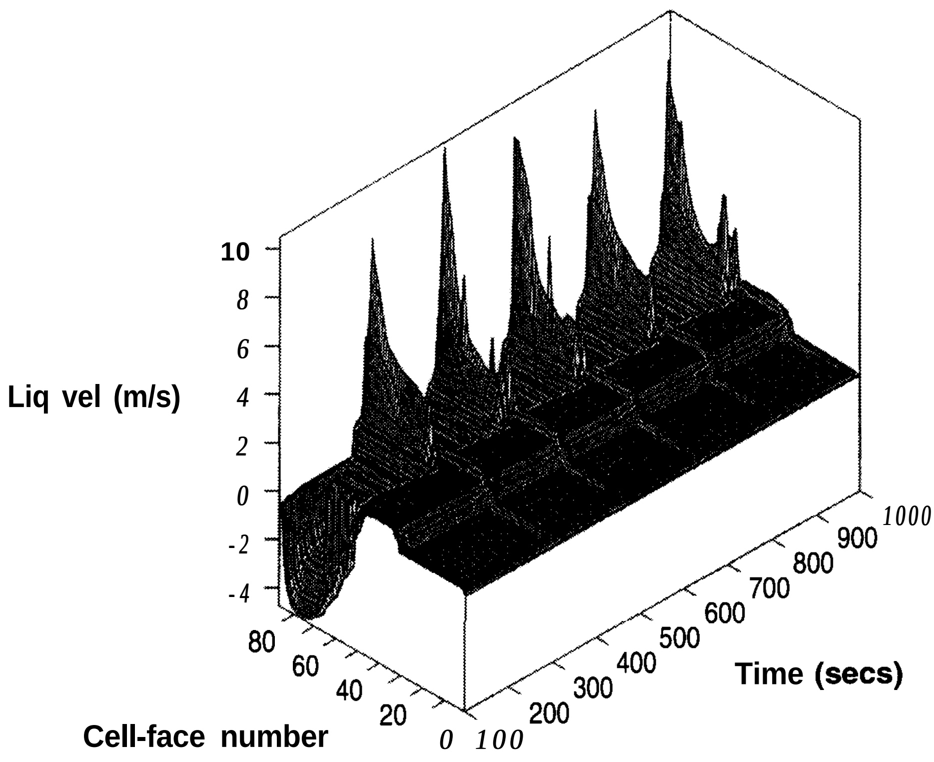


Figure 7. Liquid velocity variation

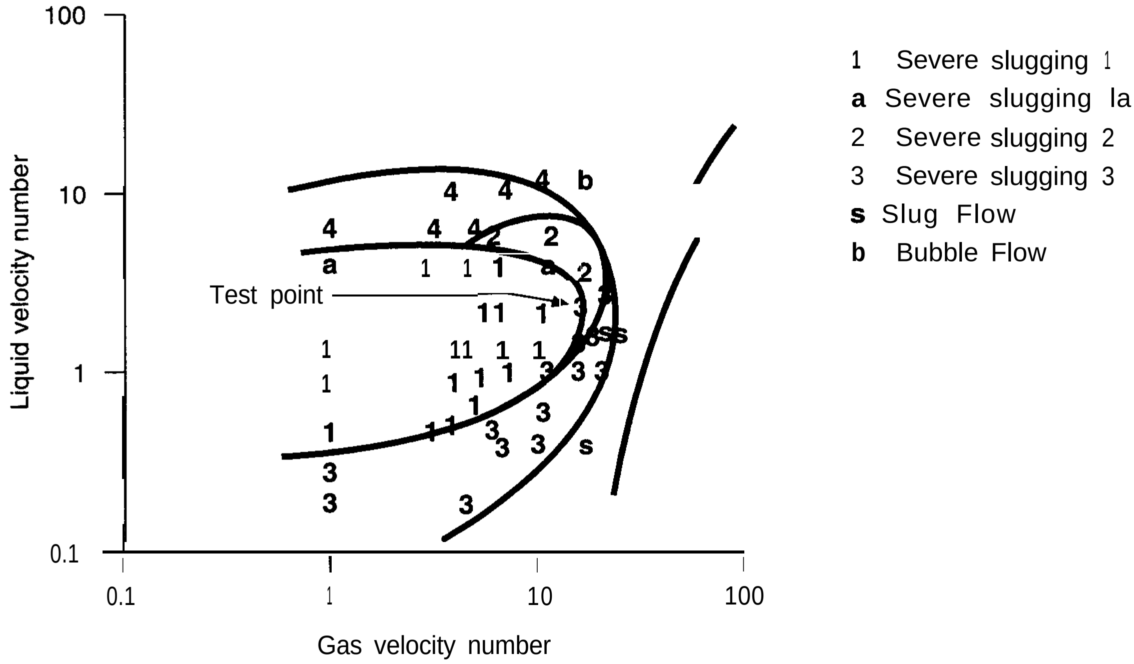


Figure 8. Flow regime observations from catenary riser experiments

References

1. G Henday, V Tin, and Z El-Oun "Severe Slugging in Flexible Risers-Interim Report". MPE 048, Jan 1990, Prepared by the British Hydromechanics Research Association.

Example 7.5 – Analysis of hilly terrain pipeline loop for Cusiana phase 1.

The Phase 1 Cusiana development contractor performed a study in October 1992 to determine the configuration of the production flowlines that were required to transport multiphase well production to the Central Processing Facilities. The flowrates were expected to increase as more wells were drilled in the future, this was accommodated in the design by extensive looping of the pipelines in order to increase the capacity. The pipe sizes were mainly dictated by flowing velocity considerations to avoid erosion limits, however in the study these were incorrectly calculated as half the required value, hence the flowlines were significantly oversized. The Multiphase Flow Group in XFE were commissioned to investigate the proposed multiphase gathering system and to investigate the potential multiphase design issues.

The philosophy of looping was investigated by considering a 20" pipeline loop that was proposed between the original well pad 504 and the Cusiana CPF. The topography of this section is shown in Figure 1 and illustrates the large elevation changes as the pipeline negotiates the hills, the Cusiana river, and the approach to the CPF which is located on a mesa. Both pipes are assumed to have identical profiles. The analysis of the hydraulic operation of the pipeline loop was conducted using MULTIFLO simulations to determine the possible flowing solutions based on a steady state analysis.

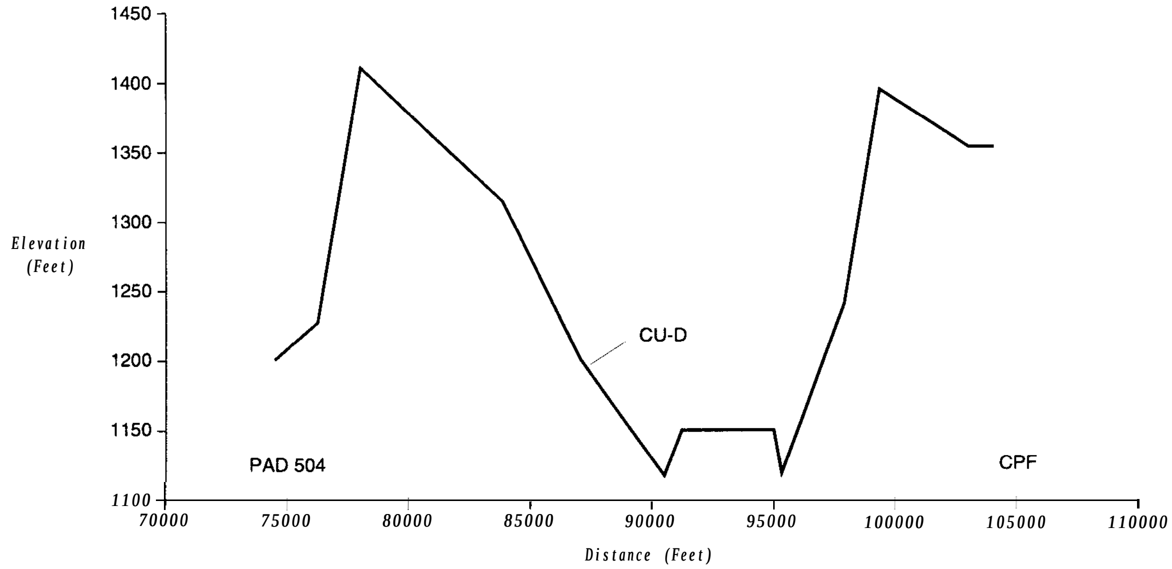


Figure 1. Topography of 20" looped pipeline

The table below shows the calculated gas-oil ratios for a 70 mbd oil flowrate with a producing gas-oil ratio of 2000 scf/stbo. At the inlet conditions to the loop of 600 psia and 175°F the calculated solution gas-oil ratio is 180.81 scf/stbo. α is the fraction of the inlet liquid flowrate in line A and ß is the fraction of the inlet free gas flowrate in line A.

ß	Liquid fraction α						
	0.1	0.2	0.4	0.6	0.8	0.9	1 .0
0.0	181	181	181	181	181	181	181
0.1	2000	1090	636	484	408	383	363
0.2	3819	2000	1090	787	636	585	545
0.4	7458	3819	2000	1394	1090	989	908
0.6	11096	5638	2910	2000	1545	1394	1272
0.8	14734	7478	3819	2606	2000	1798	1636
0.9	16554	8367	4274	2910	2227	2000	1818
1 .0	18373	9277	4729	3212	2455	2202	2000
Qoa	7	14	28	42	56	63	70

Effective GOR's for use in two-phase simulations

The MULTIFLO predicted pressure drops for each gas and liquid flowrate combination are shown in the table below where the pressure drops are calculated for a fixed value of ß at each value of α and hence GOR and oil flowrate Qoa.

ß	Liquid fraction cx						
	0.1	0.2	0.4	0.6	0.8	0.9	1 .0
0.0	163	162	163	164	167	169	171
0.1	118	130	123	109	122	129	135
0.2	91	101	95	96	112	120	128
0.4	68	83	75	94	114	124	134
0.6	65	78	80	104	127	138	149
0.8	68	82	92	118	144	157	170
0.9	71	86	99	127	154	168	181
1 .0	75	91	108	137	165	179	193

Overall two-phase pressure drops for line A

The data can be plotted as pressure drop vs α for each value of ß. If the line sizes or topographies of the loop are different the calculations must also be repeated for line B. However if both lines are the same the results will be symmetrical ie α = 0.9 and ß = 0.9 for line A corresponds to α = 0.1 and ß = 0.1 for line B. The result for the case considered is shown in Figures 2 and 3, where the locus of possible solutions is shown in Figure 3.

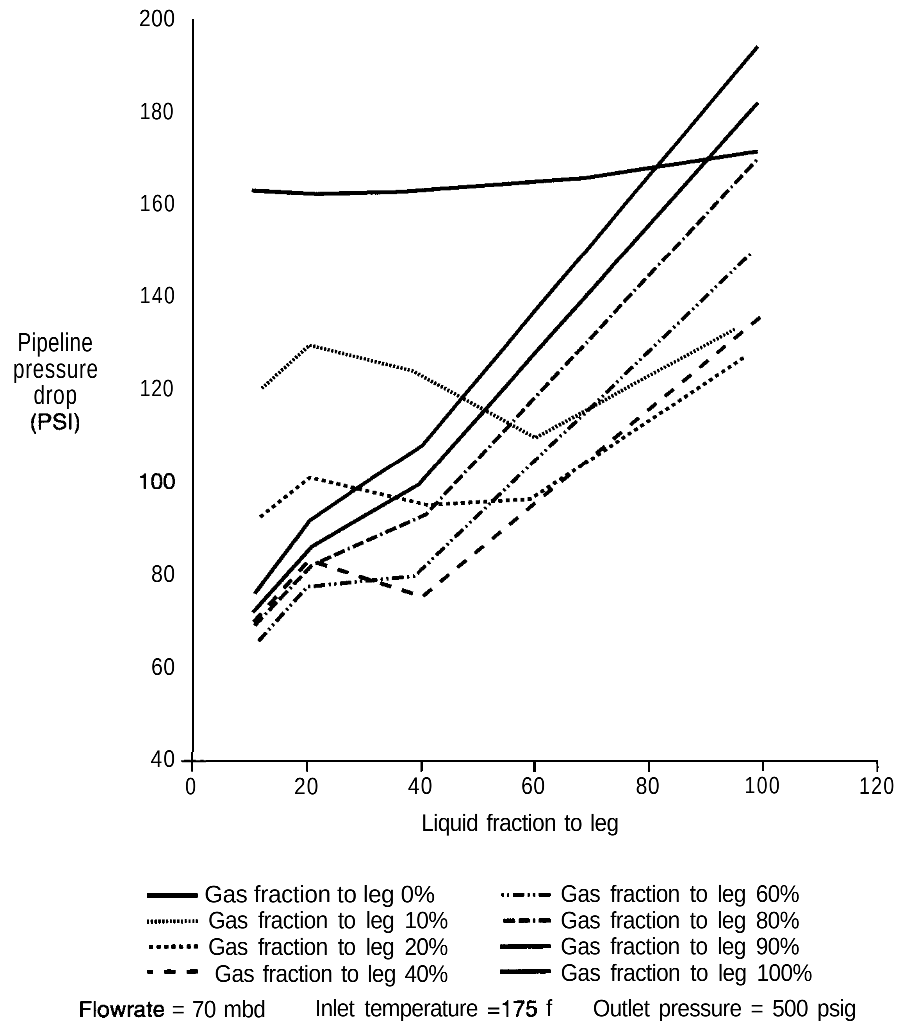


Figure 2. Solution for line A

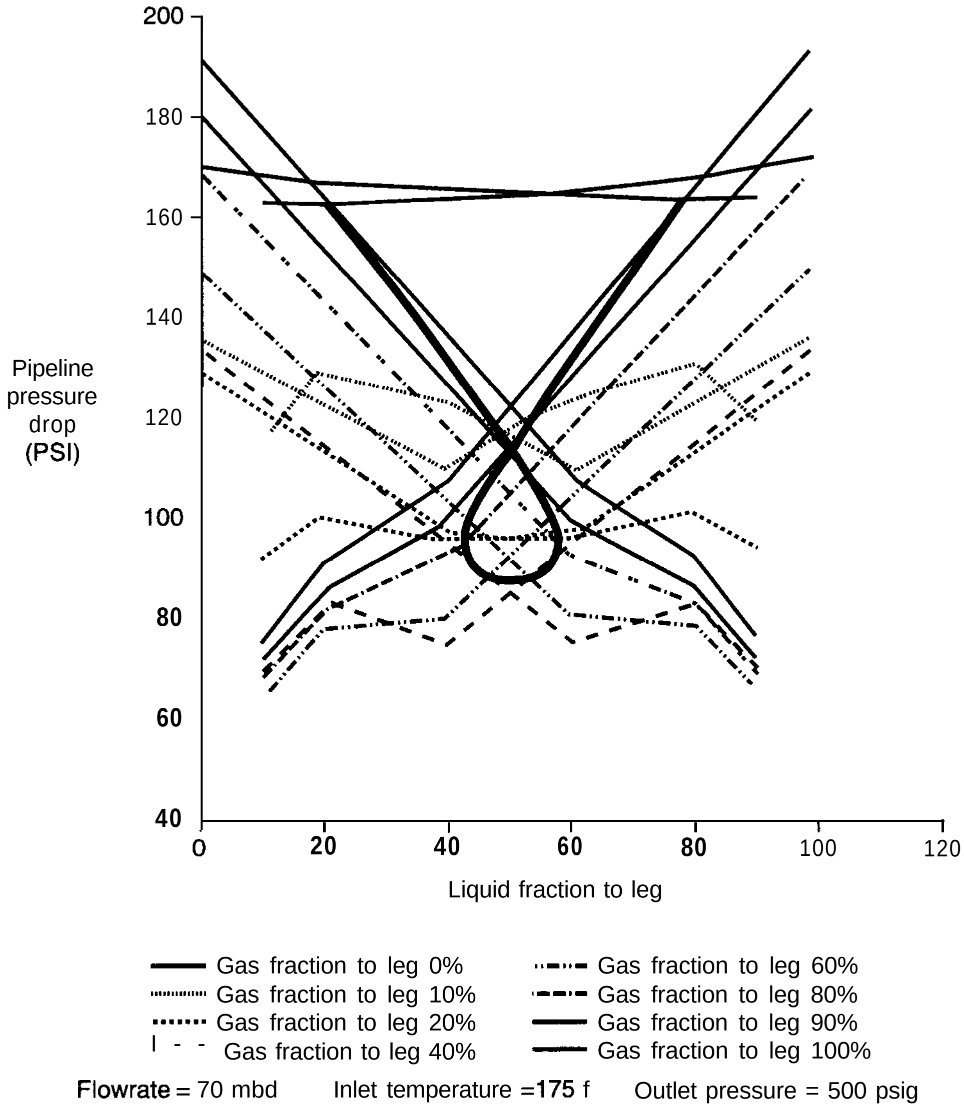


Figure 3. Combined solutions

The figures illustrate that at 70 mbd no zero flow or manometer effect can exist since 0% liquid fraction is not a solution. The nature of the tee junction at the loop is such that the liquid may preferentially take the run and this could give a 100% free gas split into the other leg with 22% of the liquid flowing in the run. Note that between 40% and 60% liquid split to the run several solutions are possible with different overall pressure drops. The main problem is the liquid holdup variations in each leg if the flow switches. The holdup at 0% gas split to the run is 9377 bbls and the flowrate is 15.4 mbd. In the side leg the corresponding holdup is 2264 bbls giving a 7113 bbl slug if the flow in the loop switches. The possibility of unstable operation was part of the reason for recommending that the Cusiana multiphase flowlines should not be looped.

For the example above the sum of the uphill elevations is 504 ft, which gives rise to a hydrostatic head of 163 psi. Figure 4 indicates that manometer effects are possible when the inlet flowrate is reduced below around 60 mbd since the flowing pressure drop could be balanced by a static column of liquid in the other leg.

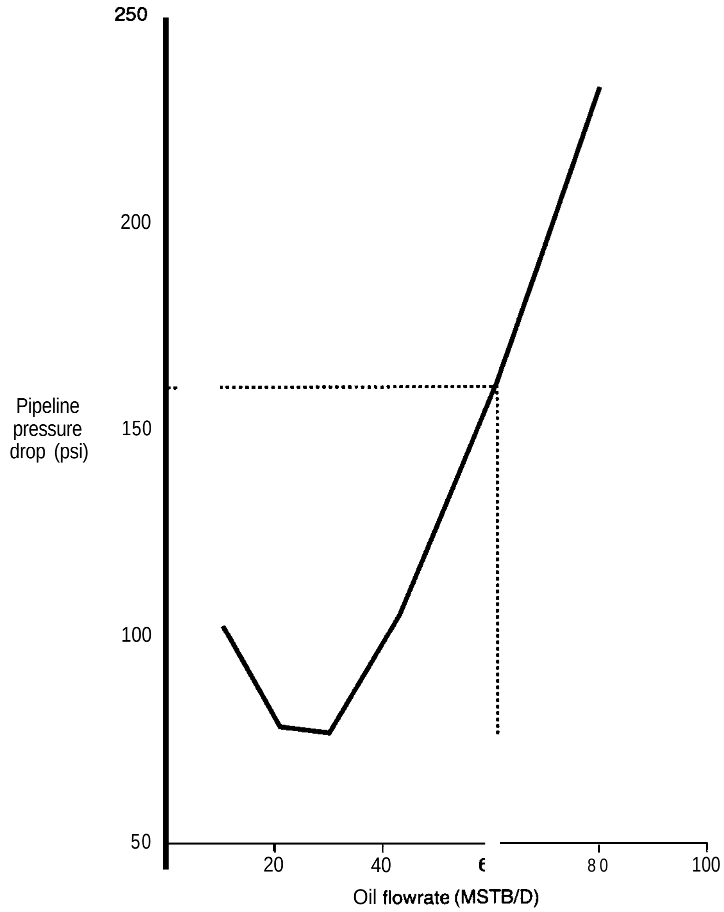


Figure 4. Pressure drop characteristic for total flow in the one leg

Example 7.6 – Using two-phase pressure surge analysis to determine pig-slug loads

The effect of a pig-slug impacting on a closing pressure control valve was studied as part of the design of the Rough field slug catcher at Easington. An isometric of the approach pipework to the slug catcher is shown in Figure 1 where it is seen that the 34" sealine terminates in a 17m riser at 'A' then follows horizontally, and finally rises at 45 degrees between 'D' and 'E' before entering the sphere receiver. The PCV is located just upstream of the slug catcher. The PCV is designed to control the downstream pressure independently of upstream conditions. Flow control is provided downstream of the gas offtake 'J'.

It was required to estimate the loads produced by the pig-slug impacting on the PCV if it closed during pig reception, and hence to determine the need to set the valve fully open during pigging. The slug catcher is located 1 Om above ground level and it was required to assess the structural loads that may be generated by the slug dynamics.

A simple dynamic model of the slug reception process indicated that the slug velocity increases from a initial value of 7m/s to a final velocity of 21 m/s as the tail passes the PCV and the slug is discharged into the slug catcher. This acceleration is due partly to the hydrostatic head loss and the reducing frictional length of the slug as it is produced.

initial simulations using FLOWMASTER were based on the surge pressures generated by closing the PCV during slug reception. A linear valve closure rate was assumed and the effect of the valve closure time investigated. The results of a single phase surge analysis are shown in Figure 2 indicating that surge pressure should not exceed 95 bar, which is well below the maximum allowed pressure of 150 bar. The valve closure times of around 30s are too long to generate significant unequilibriated loads in the short piping runs. What was of more concern was the impact of the slug front on a partially open control valve, which is more difficult to assess.

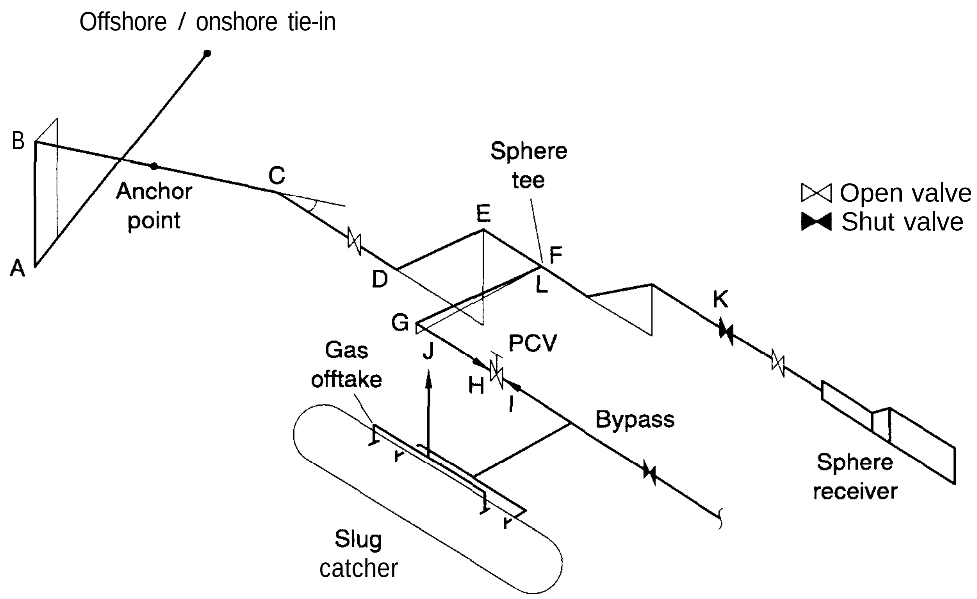


Figure 1. Isometric view of approach pipework to the slug catcher

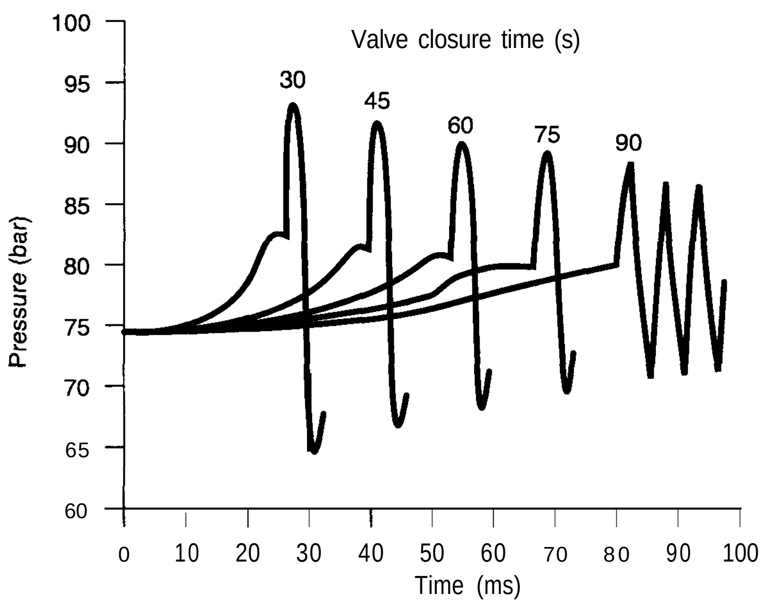


Figure 2. Results of single phase liquid surge analysis

One of the major unknowns is the time over which the slug front impact occurs. The worst case is to assume a vertical slug front and hence an instantaneous impact. The results of using the primer component in FLOWMASTER to simulate this is shown in Figure 3 for locations 'F', 'G', and 'H'. The slug was in this case assumed to be travelling at 6 m/s through a 20% open control valve. An all liquid slug was assumed.

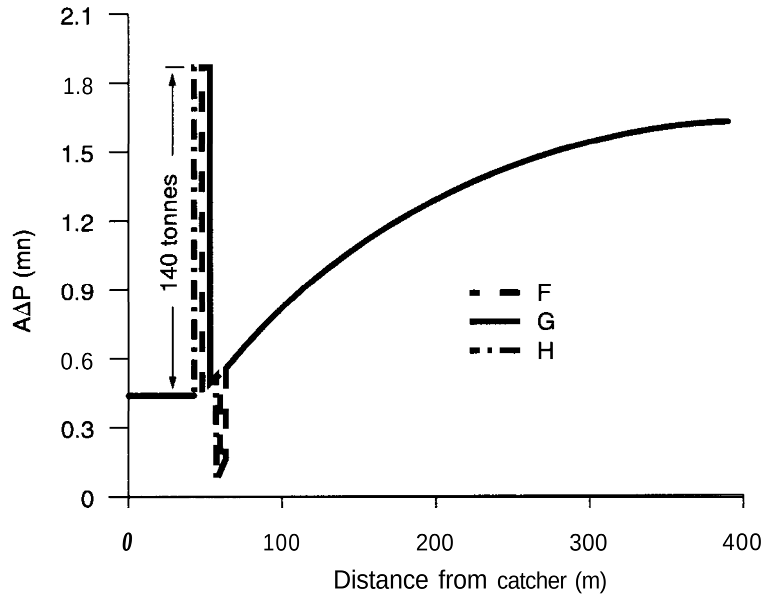


Figure 3. Pressure histories for worst-case impact assumptions

In practice the slug may typically have a front sloping at 30" which will have the effect of increasing the impact time. This was also investigated using FLOWMASTER, showing that increasing the impact time from 0 to 50 ms has the effect of reducing the peak surge pressure from 140 tonnes to 20 tonnes (see Figure 4). At an arrival speed of 6 m/s this is equivalent to a slug front length of one-third of a pipe diameter, and hence is quite feasible. Although it would seem likely that long impact times, and hence small loads, should prevail, the dynamic nature of the slug front makes it difficult to accurately predict the impact time and the possibility of an instantaneous impact should be considered.

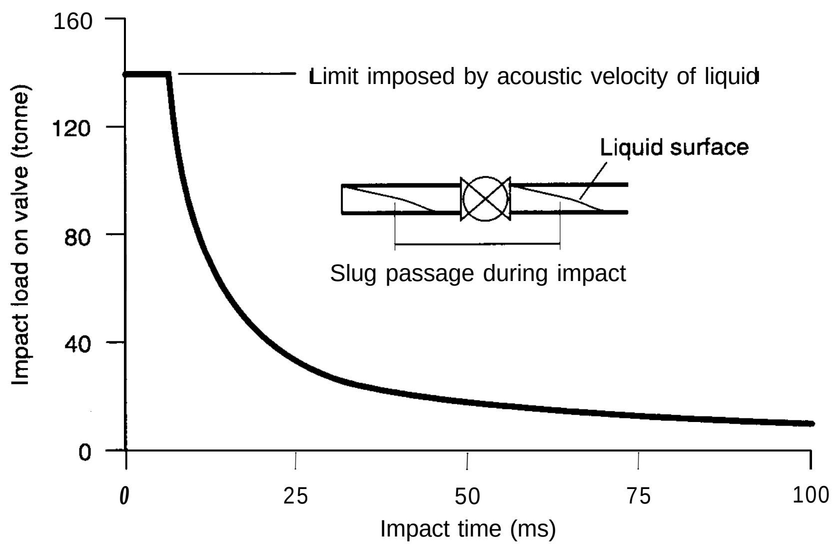


Figure 4. Influence of impact time on load predictions

The reaction loads due to the unhindered slug passage are more easily calculated. These are due to the change in the momentum between the liquid slug and the gas and are related to the slug front and tail velocities. Ignoring elevation effects the slug front velocity is constant and hence the 'coming on' load is fixed. However the slug tail accelerates during slug reception, and hence the largest loads are the 'coming off' loads produced by the passage of the slug tail. These loads are shown in Figure 5 and indicates that the maximum load is of the same order as that produced by a 50ms impact surge. Slug loads are further discussed in Section 11.

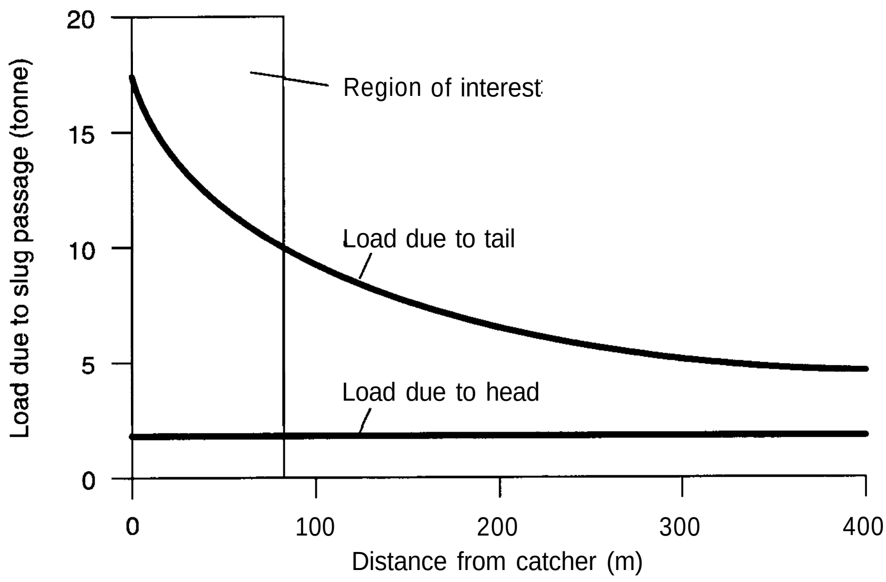
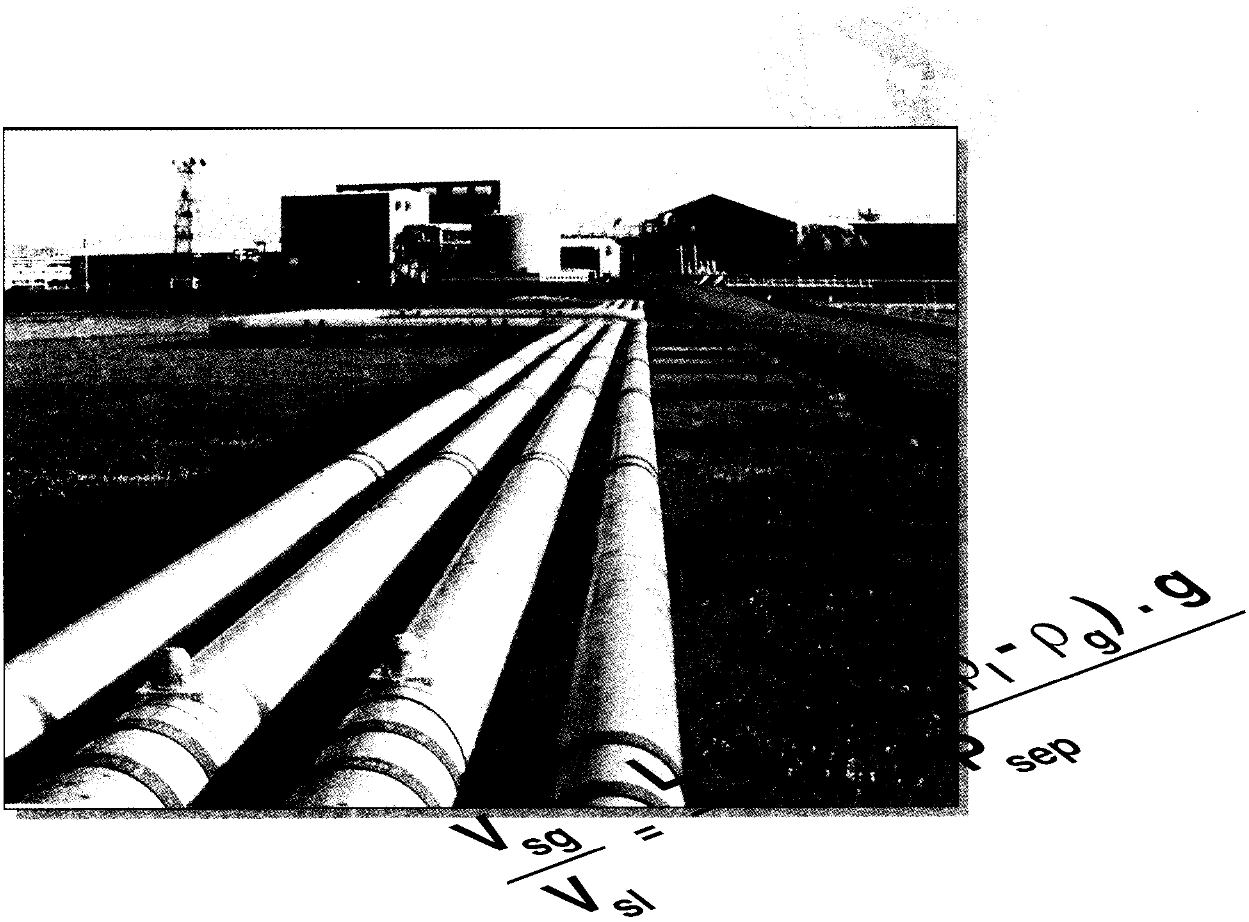


Figure 5. Influence of slug passage on dynamic loads

Section 8 Depressuring and emergency blowdown of single and multiphase pipelines



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8.1 Definition of the terms: Depressuring, Blowdown and Rupture

In this manual the terms depressuring, blowdown, and rupture have the following meanings:

Depressuring

Depressuring is generally used to refer to the controlled and relatively slow evacuation of a pipeline system. Depressuring is usually performed to make the pipeline available for maintenance or repair. Depressuring a pipeline will usually take several hours or even days.

The simulation of pipeline depressuring is discussed below in Section 8.2.

Blowdown

Blowdown is generally used to refer to the controlled but rapid evacuation of a pipework or process plant vessels. Blowdown is sometimes referred to as emergency depressurisation.

Blowdown is used to minimise the escalation of any incident by reducing the risk of pipework or vessel rupture in a fire. In addition blowdown will reduce the hydrocarbon inventory which could feed the fire. Blowdown is normally accomplished in a time of the order of 15-30 minutes.

Blowdown philosophy and modelling is discussed below in Section 8.3

Rupture

Accidents resulting in fluid loss may occur as a result of overpressurisation, material failure and impact damage. The fluid loss occurring as a result of a rupture will depend on fluid composition, hole size, initial pressure, pipeline characteristics, and mass of fluid available in the pipe to feed the rupture.

The fluid obviously escapes from a rupture in an uncontrolled manner. Pipeline rupture modelling is discussed in Section 9 of this manual.

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8.2 Depressuring of single and multiphase pipeline systems

8.2.1 Introduction

The Model discussed here is used to determine a conservative minimum pipe temperature experienced during the controlled depressuring of a pipeline system. The model is applicable for the relatively slow depressuring of a system over many hours/days. For more rapid depressuring events the approach discussed here will produce very conservative temperatures i.e. it will significantly under predict the minimum pipewall temperature. The blowdown model discussed in Section 8.3 below is more appropriate for the rapid emergency depressuring events which occur over a time scale of 15-30 minutes.

The Pipeline model described in this section is applicable for horizontal or slightly inclined single and two phase systems, It also predicts liquid entrainment rates and the critical velocities required to produce slug or annular-mist flow in a riser. The temperature downstream of the letdown valve/orifice is also computed.

8.2.2 Overview

The simulation of the depressuring of a pipeline system is performed in order to determine the minimum temperature experienced, or to calculate the time required to depressure a system such that the temperature does not drop below the minimum specification value.

The purpose of a depressuring calculation is therefore to determine the relationship of pressure and temperature against time for a particular system and depressuring rate.

The calculation has three main elements. These are;

(a) Depressuring Rate

The depressuring rate is a critical aspect. The faster the depressuring the lower the temperature that will be experienced.

The venting rate can be fixed (e.g. 100 mmscfd) by using a control valve and a mass flowrate meter. Alternatively, the system can be vented through an orifice. In most cases the flow through the orifice is critical and therefore the depressuring rate is independent of the downstream pressure.

(b) Auto-refrigeration

As the pressure decreases, the temperature of the system will drop due to Joule-Thompson cooling (auto-refrigeration). The expansion of the gas is normally considered to be isentropic. As the gas expands it does work and the energy required to do this work is ‘removed’ from the gas in the form of heat. The auto-refrigeration is not dependent on the duration of the depressuring, only the physical properties of the fluid.

(c) Heat Input from the Surroundings

To depressure a pipeline system usually requires many hours, and in some cases days. During this period the heat input from the surroundings may be large.

The auto-refrigeration is not time-dependent and is only dependent on the expansion of the gas in the system. The 'heat from the surroundings' is time dependent. The longer the depressurisation, the greater the influence of the surroundings whilst the auto-refrigeration aspect remains (roughly) constant. Therefore, increasing the depressuring time will result in a higher minimum temperature.

8.2.3 Depressuring Rate Calculation

It is usual to depressure a system through an orifice or a valve. The equation for the mass flowrate through an orifice is given below (Equationl). The equation for a valve is of a similar form with slightly different constants (not given here).

$$W = C_d KAP_1 \sqrt{M_W / zT} \qquad (1)$$

(For nomenclature, see the section at end of this document.)

K, the specific heat ratio function is defined by the following equation;

$$K = \sqrt{\left(\frac{\gamma}{R}\right)\left[\left(\frac{2}{\gamma+1}\right) \wedge \left(\frac{\gamma+1}{\gamma+1}\right)\right]} \qquad (2)$$

The equation above is only valid for critical (i.e. sonic) flow. Critical flow only occurs when the following criteria is satisfied (equ.3).

$$\frac{P_2}{P_1} < \left(\frac{2}{\gamma+1}\right)^{(\gamma / \gamma^{-1})} \qquad (3)$$

The occurrence of critical flow greatly simplifies the calculation. If the system is operating under critical flow then the fiowrate is only dependent on the upstream pressure.

It is normal to consider the depressuring to be complete when the system pressure falls below 2 bara. For a downstream pressure of 1 bara, critical flow exists down to pressures of 1.83 bara (γ = 1.3) and so therefore critical flow is usually assumed for the whole of the depressuring. Critical flow may not exist at the lower flowrates where the downstream pressure is significantly greater than atmospheric,

Whilst the pressure ratio satisfies Equation 3, the flowrate (W) is independent of the downstream pressure.

8.2.4 Heat Transfer

(a) Factors Affecting the Minimum Temperature

The temperature of the fluid is determined by the effect of cooling due to the Joule-Thompson expansion of the gas and warming by the influence of the surroundings. The conservatism of the equation is described below;

Isentropic Expansion.
Isentropic expansion means that the gas does the maximum amount of work as it expands. This means that it loses the maximum amount of energy (i.e. heat) and therefore results in the prediction of the minimum possible (i.e. conservative) temperature.

Warming Influence of the Surroundings.
When the gas temperature drops below the ambient temperature the pipeline is warmed by the surroundings. To calculate a conservative (i.e. low) minimum temperature the influence of the surroundings must be underestimated.

By overstating the cooling effects of expansion and understating the heating influence of the surroundings, the calculation of a conservative minimum temperature is ensured.

(b) Internal Heat Transfer Coefficient

The calculation of an accurate internal heat transfer coefficient is important since it greatly influences the prediction of the minimum temperature. The method used must calculate a conservative (i.e. low) heat transfer coefficient for at least the period that the system is at its minimum temperature.

The correlation used to determine the heat transfer coefficient will depend on the type of system and the method of depressuring. Details of the equations used to calculate the internal heat transfer coefficient are given in Appendix 8.2.A.

(c) External Heat Transfer Coefficient

The external medium has a significant effect on the calculated minimum fluid/wall temperature. If the pipe is located in the air (e.g. topsides of a platform) then the overall heat transfer coefficient will be low, giving rise to a relatively low minimum wall temperature. If the pipe is located in the sea the overall heat transfer coefficient will be higher and the minimum wall temperature will accordingly be higher (assuming, for this comparison, that the air and sea temperatures are equal)

(d) Minimum Tube Wall Temperature

The objective of most depressuring studies is the calculation of a minimum pipe wall temperature, or the calculation of a depressuring rate which prevents the system from falling below the minimum tube wall temperature.

In a gas-only system the heat transfer rate between the gas and the pipe wall will be low, and consequently the temperature difference between the fluid and the tube wall will be relatively high. In order to be conservative, it is usual to assume that the fluid temperature and the tube wall temperature are identical throughout the depressuring. This assumption will always result in the calculation of a conservative (i.e. low) minimum tube wall temperature since it ignores the existence of a temperature profile between the fluid and the wall.

In a two-phase system, the heat transfer rate between the liquid and the pipe wall will be significantly higher than the heat transfer rate between the gas and the pipe wall. It is reasonable to assume that where pools of liquid occur (in dips etc.) the high heat transfer rate will result in the pipe wall cooling to the same temperature as the fluid. However the effect of the fluid being warmed by the pipe wall is ignored and hence a conservatively low pipe wall temperature is predicted.

8.2.5 Simultaneous 'Riser' and 'Topsides' Depressuring

In most studies the 'riser' and 'topsides' pipework will be depressured simultaneously. The majority of the length of the riser will be immersed in the sea and have the benefit of the sea as a heat source. Riser pipework is generally much simpler than the Topsides pipework.

The topsides pipework is more complicated (e.g. a large number of branches etc.), and located in the air. The poor heat transfer properties of air coupled with the lower minimum ambient temperature mean that the topsides pipework will drop to a lower temperature relative to the riser. Because of this difference in heat transfer rates the topsides and the riser should be modelled separately.

8.2.6 Critical Velocity for Slugging

Slugging should be avoided where possible. To avoid this the depressuring rate should not exceed the critical velocity for slugging. The angle of the pipe (to the horizontal) is used to calculate the conditions required for slugging.

8.2.7 Flow Regime in Riser

Annular mist flow is unlikely to occur in the riser. The actual velocities experienced in the riser are usually approximately an order of magnitude less than that required to produce annular-mist flow.

8.2.8 Conditions Downstream of Letdown Valve

The temperature downstream of the let-down valve will invariably be significantly colder than that experienced in the topsides or riser pipework. Temperature specifications of the downstream pipework are not usually a problem since stainless steel is usually the material of construction. Liquid drop-out may occur due to the low fluid temperature, and this will need to be accommodated in a liquid KO vessel.

Appendix 8.2a

Heat Transfer Equations

The calculation of an accurate internal heat transfer coefficient is important since it greatly influences the prediction of the minimum temperature. The method used must calculate a conservative (i.e. low) heat transfer coefficient for at least the period that the system is at its minimum temperature.

The correlation used to determine the heat transfer coefficient will depend on the type of system and the method of depressuring.

The internal heat transfer coefficient is calculated using one of the following equations:

Stream Line Flow (Re < 2100)

If the flow is laminar (Reynolds number less than 2100) then the following equation should be used:

$$h_i = \frac{1.86k_m}{d_i}\left[\left(\frac{d_i u_m \rho_m}{\mu_m}\right)\left(\frac{C_{pm}\mu_m}{k_m}\right)\left(\frac{d_i}{L}\right)\right]^{1/3} \qquad (4)$$

["Process Heat Transfer". D.Q. Kern. equ. (6.1). page 103]

As the above equation and those following will usually be applied to non-viscous gases the viscosity correction factor is ignored.

Correction For Natural Convection

For very low flowrates natural convection (rather than forced) convection can become important. The forced convection coefficient (hi) can be corrected by multiplying by

$$\Psi = \frac{2.25\,[1+0.01\,(Gr_a^{\,1/3})]}{\log_{10} Re} \qquad (5)$$

["Process Heat Transfer". D.Q. Kern. Equ (10.4). page 206]

Where Gr_a is the Grashof number at the average conditions, given by:

$$Gr_a = \left(\frac{d_i^3\,\rho_m^2\,g\beta\Delta T}{\mu_m}\right) \qquad (6)$$

["Process Heat Transfer". D.Q. Kern. Table 3.2. page 37]

This factor should be ignored if less than 1.
The Grashof number is difficult to calculate since the temperature difference between the bulk

fluid and the tube wall (AT) has to be determined iteratively. It can be done by guessing a wall temperature and iterating such that the heat flux across the internal film is the same as the heat flux through the pipe and the insulation.

If this correction is ignored, then the internal heat transfer coefficient will be under-predicted and a conservative (i.e. low) heat transfer coefficient will be determined.

Turbulent Flow

If the Reynolds number is greater than 2100, then the following should be used

$$h_i = \frac{0.023k_m}{d_i}\left(\frac{du_m\rho_m}{\mu_m}\right)\left(\frac{C_{pm}\mu_m}{k_m}\right)^{0.4} \tag{7}$$

["Chemical Engineering. Volume 1 ". Coulson and Richardson. Equ (7.50). page 191]

(Note; This is the same method as MULTIFLO, BP's in-house two-phase pipeline modelling program.)

Alternatively, Equation 7 can be re-written as below:

$$h_i = 0.023(k_m)^{0.6}\ (d_i)^{-0.2}\ (\mu_m\ \rho_m)^{0.8}\ (\mu_m)^{-0.4}\ (C_{pm})^{0.4} \tag{8}$$

m denotes a mixture property, e.g. for viscosity.

$$\mu_m = \mu_l\,\lambda + \mu_v\,(1-\lambda) \tag{9}$$

The velocity at which the heat transfer coefficient is to be calculated should be 75% of the exit velocity. This allows for the fact that there is a velocity profile through the system. If possible the velocity should also be based on the average velocity for the time step.

Appendix 8.2b

Nomenclature

A	= Orifice/valve area for venting	m^2]
C_d	= Coefficient of Discharge	I-1
C_{pm}	= Mixture heat capacity	[J/kg K]
d_i	= Internal pipe diameter	[m]
g	= Gravitational constant	[9.814 m/s^2]
Gr_a	= Grashof number	[-]
h_i	= Internal heat transfer coefficient	[W/m^2.K]
K	= Function of specific heat ratio	[kg-mol$^{0.5}$.K$^{0.5}$J$^{-0.5}$]
k_m	= Mixture thermal conductivity	[W/m.K]
L	= Length of System	[m]
M_w	= Molecular Weight	[kg/kg-mole]
P_1	= Upstream Pressure	[N/m*]
P_2	= Downstream Pressure	[N/m*]
R	= Gas constant	[8314 J.Kgmole^{-1}.K^{-1}]
Re	= Reynolds' number	[-]
T	= Temperature	[K]
u_m	= Mixture velocity	[m/s]
w	= Mass flow rate	[kg/s]
ß	= Thermal expansivity	[1/K]
γ	= Ratio of Specific Heats	[-]
λ	= Volume fraction of liquid	[-]
μ_l	= Liquid viscosity	[Ns/m^2]
μ_m	= Mixture viscosity	[Ns/m^2]
μ_v	= Vapour viscosity	[Ns/m^2]
ρ_m	= Mixture density	[kg/m^3]
Ψ	= Free Convection Coefficient	[-]
Z	= Gas compressibility	[-]

MULTIFLO

EXTMED	= External medium h.t.c.	[W/m^2.K]
INSLTN	= Insulation h.t.c	[W/m^2.K]
PIPEWL	= Pipewall h.t.c	[W/m^2.K]

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8.3 Emergency blowdown of pipes containing gas and liquid

8.3.1 Background

This section of the manual is concerned with the simulation and philosophy of emergency blowdown (i.e. rapid depressurisation) of pipes and vessels containing gas and liquid. Conditions experienced during blowdown generally set the minimum design temperature of those pipes and vessels subject to emergency blowdown.

A review of blowdown issues concluded that the only program that could accurately model the blowdown process was developed by Saville and Richardson of Imperial College London. They have been working in the area for many years developing both the experimental facilities to provide accurate data, and the theoretical models to simulate the blowdown process.

Saville and Richardson have been contracted on many occasions by BPX to carry out blowdown analysis of process plant and pipework. However, BP identified a need to have a simplified blowdown model available to assist design teams in carrying out sensitivity studies and in developing optimum blowdown procedures. A blowdown module has consequently been developed within the GENESIS flowsheeting package using a simplified version of the approach taken by Saville and Richardson. This is discussed below in 8.3.2.

8.3.2 Blowdown Philosophy

It is important that a blowdown philosophy document is developed in the early stages of any design, defining not only the relevant cases but also the start conditions from which the blowdown occurs.

(a) Code Requirements

API RP 521 provides the general guidelines for determining the need for emergency blowdown. There are, in fact, no mandatory blowdown requirements. However, emergency blowdown has become accepted in the industry as being necessary for all offshore installations, other than for some unmanned platforms. The BP Code of Practice CP 37 clearly defines this philosophy.

CP 37 sets out the 3 basic reasons for providing emergency blowdown facilities as:

1. To reduce the risk of vessel or pipeline rupture in a fire.
2. To minimise the fuel inventory which could supply a fire.
3. To minimise the uncontrolled release of flammable or toxic gas.

Emergency blowdown facilities are not required for subsea pipelines. The platform topsides section of the pipeline, including pig launcher/receiver and all pipework on the platform side of the ESDV should be blown down through the platform blowdown system. For pipelines fitted with subsea isolation valves (SSIVs) there is also a requirement to provide for emergency blowdown of the riser and subsea pipework situated between the SSIV and the top of the riser.

(b) Effects Of Blowdown

Blowdown of the contents of a vessel, pipework, or sections of a pipeline from high pressure results in auto-refrigeration of the contained fluid due to expansion of the gas and evaporation of the liquid. The resultant fall in the temperature of the pipe or vessel wall metal can result in the need for costly materials to avoid brittle fracture. In addition low fluid temperatures can result in the formation of hydrates, ice, or even solid CO_2, and these can lead to blockages. The determination of minimum fluid and metal temperatures is discussed below in Section 8.3.3.

(c) Blowdown Times and Start Conditions

Blowdown must be at such a rate that equipment subject to high temperature does not fail. API RP 521 sets out the basic requirements for blowdown times and these are further detailed in CP 37. In general blowdown to 50% of design pressure in 15 minutes must be achieved, assuming wall thicknesses of 25 mm and greater. For vessels designed to BS5500, reduction of pressure to 6.9 barg in 15 minutes must be achieved unless the rate can be related to the decrease in strength of the vessel.

Identification of realistic worst case conditions at the start of a blowdown is of great importance in establishing temperature and pressures experienced during the blowdown process. Worst case conditions would normally involve starting the blowdown from the maximum pressure and minimum ambient temperature. However, it is also quite possible to have start conditions where the fluid temperature is below ambient.

The blowdown simulation of Ravenspurn South was a case where the fluid start temperature was below ambient. During platform start-up cold wellhead fluids flash across the choke and cool to temperatures well below ambient. A shutdown of the plant at this time leaves fluids shut-in at a high pressure and low temperature. These conditions had to be taken as the worst case starting conditions for the blowdown modelling.

8.3.3 The GENESIS Blowdown Model

A blowdown module has been developed within the GENESIS flowsheeting package using a simplified version of the approach taken by Saville and Richardson. The GENESIS blowdown module contains all the most important aspects of the IC model, but it is only capable of simulating a single pipe or vessel, i.e. it cannot be used to model the blowdown of an integrated process plant.

In summary, the model uses the following approach:

a. The Vessel is divided into two zones. The top zone contains all the vapour together with any suspended liquid droplets.

The bottom zone contains the liquid pool which has dropped out of the top zone. The bottom zone appears whenever liquid is formed and is eliminated if there is no liquid pool at the bottom of the vessel.

b. Each zone is assumed to be well mixed and separately exchanges heat with the vessel wall with which it is in contact.

Heat transfer to the gas zone is assumed to be by natural convection. Studies carried out at Sunbury have validated the expressions used to model this process.

Heat transfer to the liquid zone from the wall is governed by nucleate boiling.

c. Heat transfer to/from the surroundings is not normally significant over the time periods involved in blowdown, and is therefore not considered.

Heat transfer around the walls by conduction is not considered.

d. The model does not consider a suspension of liquid droplets in the vapour zone. All liquid formed is assumed to instantly fall into the bottom zone of the vessel.

Model Validation

Because of the extensive validation work carried out by IC in the course of the development of their model, it has been assumed that the IC test cases provide accurate bench mark data with which to assess the accuracy of the BP model. The predictions of the BP model were found to be in reasonably close agreement with those of the IC program.

In all the comparisons performed the BPX model provides lower predictions of wall temperature than the IC model. Hence the BPX model is providing more conservative results that the IC model.

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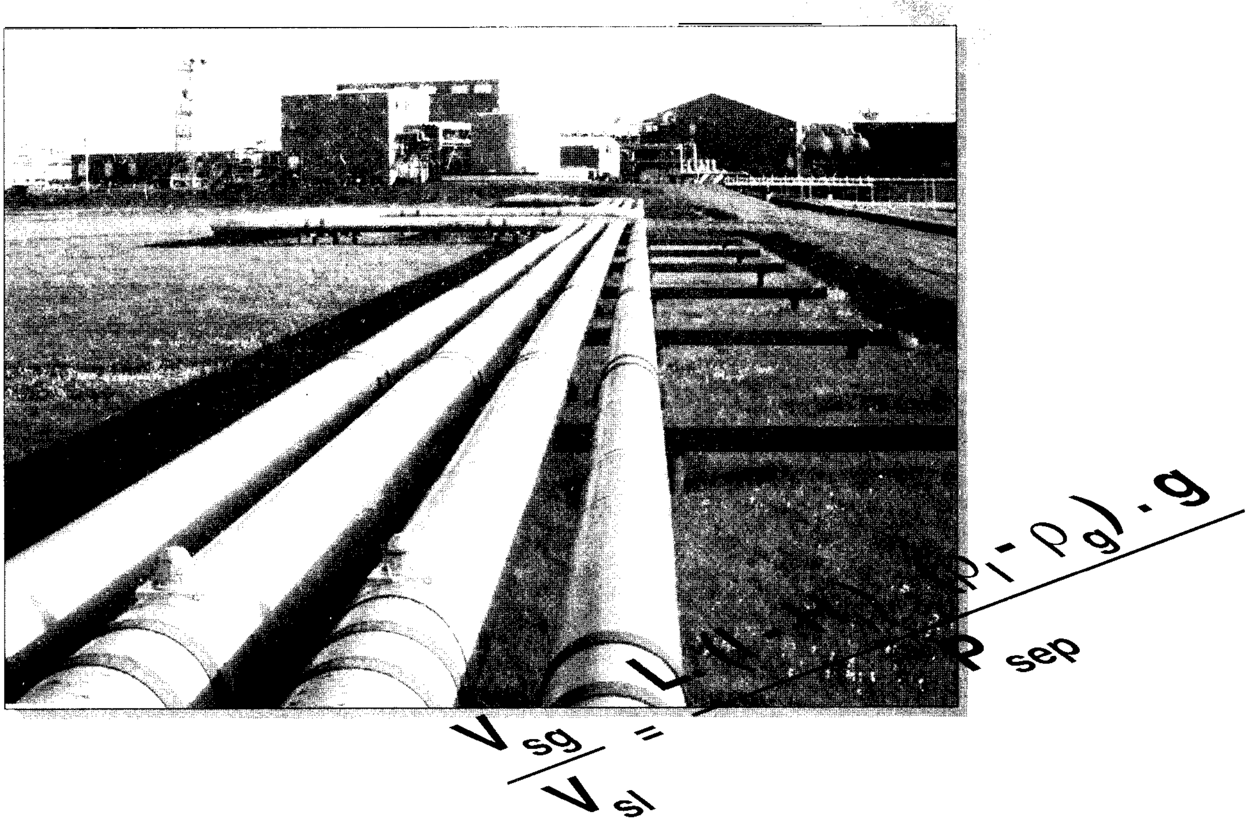
Section 9. Pipeline Ruptures Modelling

Uncontrolled Rupture of Pipeline and Pipework Systems



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9.1 Introduction to Basic Principles

Pipeline accidents resulting in fluid discharges may occur for many reasons including material failure, overpressurisation, and impact damage. If a pipe rupture occurs when a pipeline is transporting a highly volatile liquid, then on depressuring the bubble point can be reached and a two-phase mixture of gas and liquid released into the environment local to the rupture, as illustrated in Figure 1.

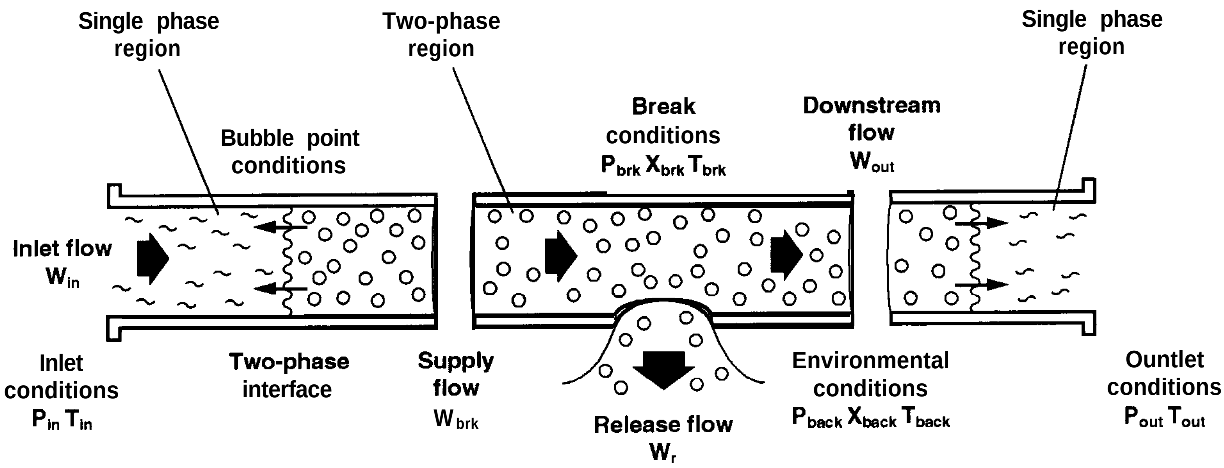

Figure 1. Conditions in the vicinity of a rupture

Single phase pipelines in the above category include those transporting LNG, LPG, gasolines and spiked crude oils, all of which can present a significant hazard if accidentally released into the environment. When pipelines transport a two-phase mixture of gas and liquid a rupture can cause rapid depressuring and critical flow at the break, additionally, with gas/condensate lines the temperature can drop rapidly due to Joule-Thompson cooling, this can have implications on the brittle fracture of the pipeline and safety valves which may cool significantly before being actuated. The analysis of such situations involves complex interactions between the depressuring hydrodynamics and thermodynamics.

Within the multiphase flow group several tools are available to perform this type of analysis ranging from simple specific codes such as PBREAKL to more complex general purpose simulators such as PLAC. Which one to use depends on the type of analysis and the accuracy required. These simulations are usually needed to support safety assessments including the impact of oil spills, gas leaks, jet fires, pool fires, and related consequence analysis.

When depressuring a gas the final temperature depends to a large extent on the type of expansion process assumed. Minimum final temperatures result from an isentropic expansion process which is a reversible adiabatic and frictionless process. If however an irreversible isenthalpic process is assumed then the friction losses result in a higher final temperature. This is illustrated in Figure 2 which shows the various expansion processes for methane expanding from 20 atmospheres and 300 K (27 °C) to atmospheric pressure. The isentropic expansion (line A) results in a final temperature of 140 K (-133 °C) where as the isenthalpic process (line B) results in a much higher final temperature of 273 K (17 °C). Although the isentropic expansion process is often used for a conservative estimate, and is probably appropriate for a full bore rupture analysis, some consideration should be given to the configuration of the upstream pipeline as the close proximity pipeline fittings on topside pipework can result in a throttling type of expansion process which is more isenthalpic. The non-ideality of the throttling process is usually accounted for by an isentropic efficiency term η which is defined as the ratio of the actual enthalpy drop to the isentropic enthalpy drop. For the example in Figure 2 (line C) the isentropic efficiency is given by:

$$\eta = \frac{4530-3770}{4530-3300} = 62\%$$

It is also seen that for a pure component substance the final temperature is independent of the expansion process if the final state is two-phase. This would be typical of a refrigeration plant where a throttle is used to de-pressure the refrigerant and cause the temperature to drop to the saturation value associated with the final pressure, hence the same final temperature is achieved regardless of whether the process is isentropic or isenthalpic. In this case the effect of the irreversible nature of the throttle is to increase the enthalpy of the fluid entering the cold box and hence to reduce the change in enthalpy available for refrigeration.

If a multi-component mixture is expanded into the two-phase region, then there is no longer a fixed saturation pressure and temperature, since the temperature is a function of the liquid mass fraction. The expansion process is hence important with multi-component fluids.

In view of the difficulty of determining the irreversible nature of the expansion process it is recommended that first estimates are based on an isentropic process since this is likely to be conservative, but is also expected to be the most accurate when applied to the de-pressuring of large volumes of fluids such as in the case of oil and gas pipelines. By far the most common request from a transient rupture analysis is an estimate of the fluid lost to the environment and the temperatures of critical components.

The need for safety assessment of the consequences of the release of volatile liquids from a rupture in a pipeline prompted the development of the PBREAKL program within BP. This code is specifically for volatile liquids normally transported single phase, but allows for flashing when depressured. In addition to this code the multiphase group also has a code called PBREAKG which is for the simulating gas pipeline rupture. These codes are limited to the application for which they were developed, where as general purpose transient two-phase codes such as PLAC and OLGA can simulate the depressuring of single-phase and multi-phase lines and are recommended for most cases. PBREAKL has some useful features and is relatively simple to use. Some of these capabilities are outlined below.

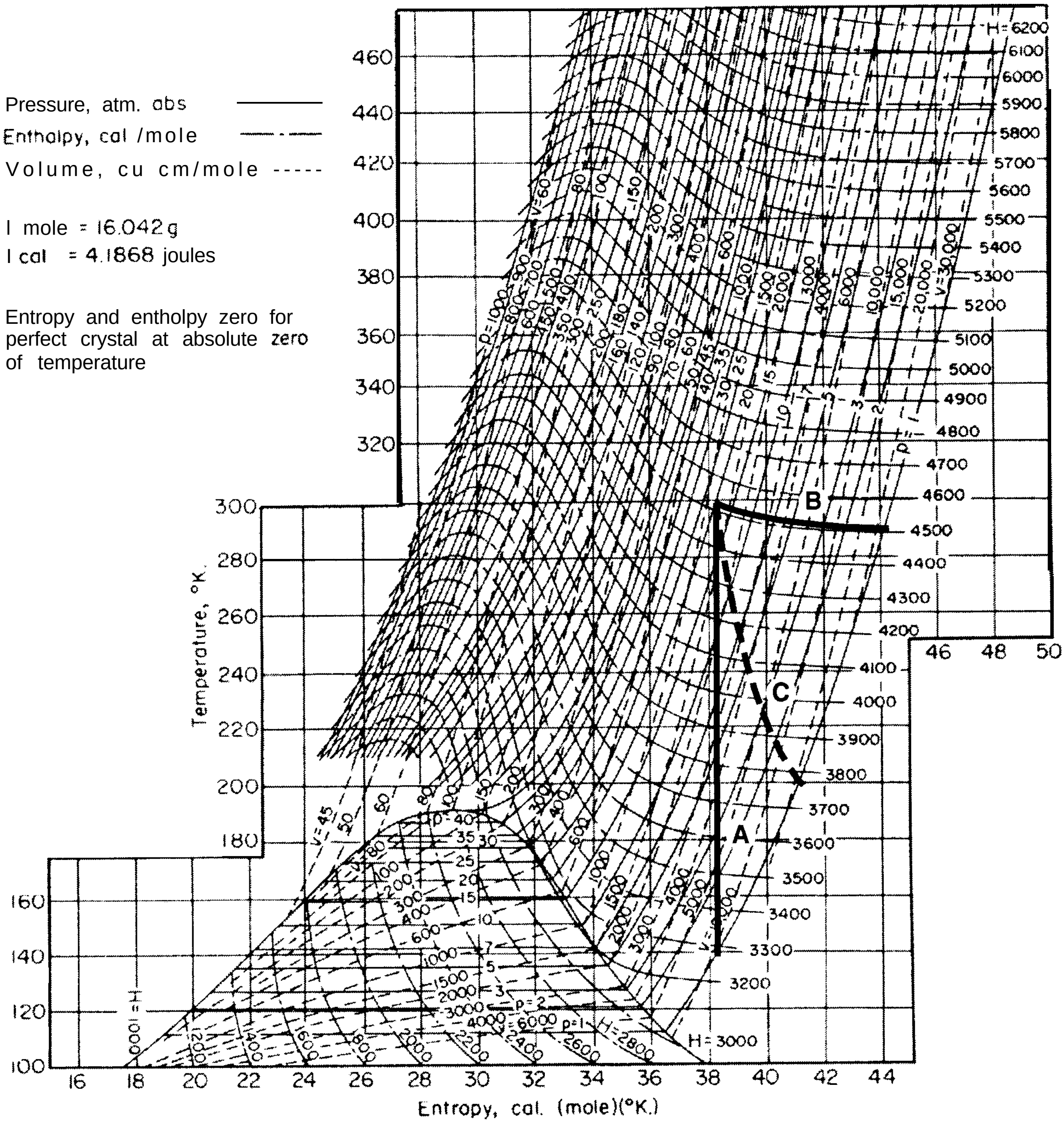

Figure 2. Temperature-entropy diagram for methane

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9.2 The PBREAKL program to model the release of volatile liquids from a pipeline rupture

The PBREAKL program predicts the single or two-phase discharge of fluids resulting from a rupture of specified size in a pipeline, including the single-phase depacking region to the bubble point. The program employs vapour-liquid equilibrium conditions and thermodynamic behaviour of two-phase flowing fluids to predict both critical and sub-critical release rates. The thermodynamic data for PBREAKL is generated using the FlA02 module in the GENESIS package to produce a physical property look-up table. The critical two-phase flow release predictions are accomplished using the method of Henry-Fauske (Ref 1) or Morris (Ref 2), the Morris method is recommended. For subcritical flow a two-phase multiplier method is applied to a standard single phase discharge orifice equation.

PBREAKL can model the effects of a simple pipeline topography with import and export risers and an isolation valve. A schematic of the physical geometry is shown in Figure 3.

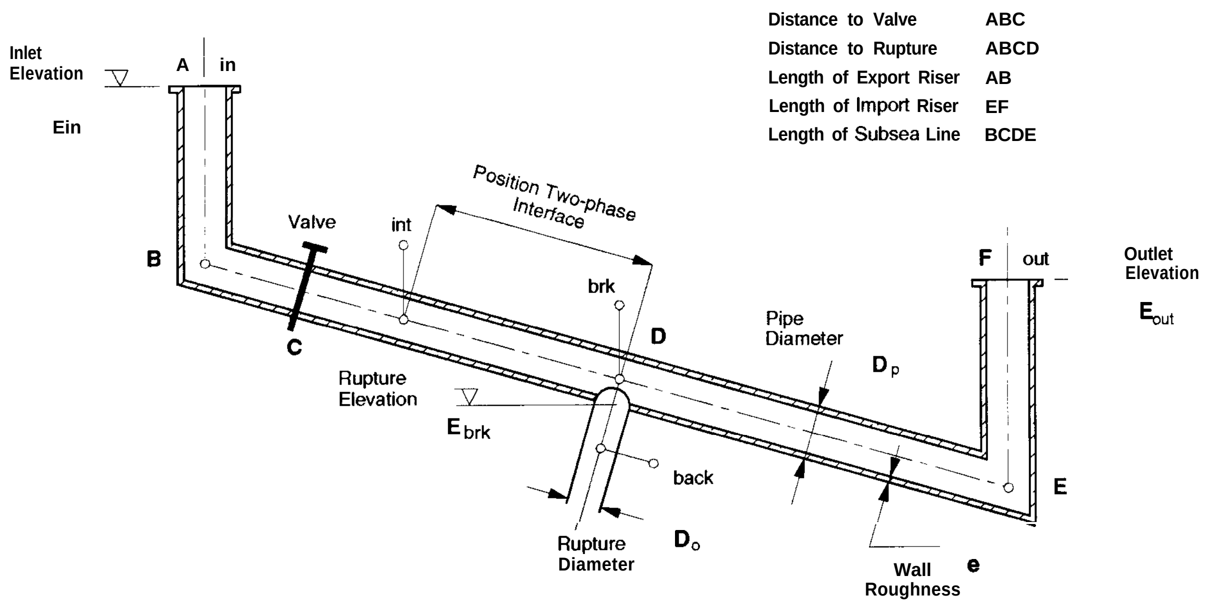


Figure 3. Schematic of pipeline system

PBREAKL has an interactive menu input where the physical dimensions are entered at the pipe geometric details screen. In addition the break diameter, the location of the rupture point and the back pressure are entered at the pipe break details screen. The flowing conditions are defined, such as inlet and outlet pressures, liquid flowrate, viscosity, and the temperature over the pipeline which is assumed to be the average line temperature. The time to valve closure (the closure is assumed to be instantaneous) and other program run time data is also input. Before the program can be run GENESIS phase equilibrium component data and flash data are required.

Prior to a typical rupture the fluid is usually said to to be "packed" within the pipeline at elevated pressures above the bubble point pressure, as a result of pipe wall flexibility and fluid compressibility. When the pipeline is ruptured the initial release is high as the pipeline depacks to the bubble point at the rupture location. When the flow flashes the initial flow is usually choked or critical and the rate of efflux is reduced. As the pressure drops within the pipeline the flow can eventually become sub-critical and the simulation stops if the release rate from the rupture equals the inflow rate.

The depacking calculation in PBREAKL is controlled by an expansion factor and an expansion time constant, which alter the wave propogation effects. These are most significant during the initial release, and as a result several types of depacking calculation are possible. These are outlined in Figure 4.

A description of each depacking method is as follows:

Method A.

This situation occurs when the leak rate reaches a steady state which is less than the inflo rate to the pipeline.

Method B1.

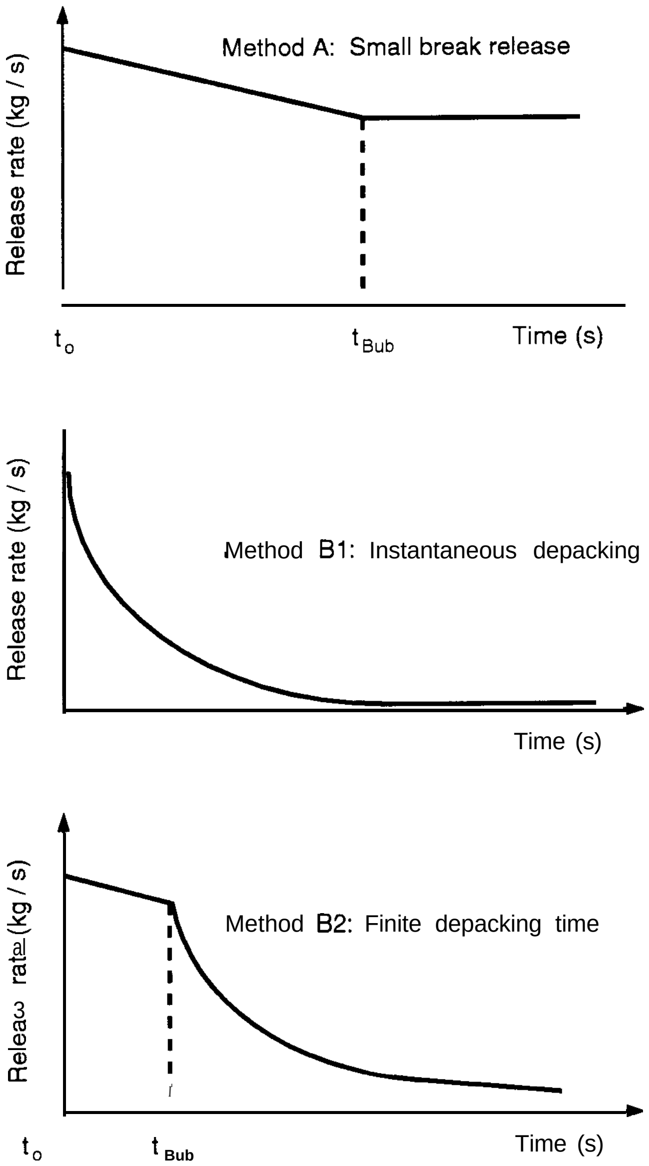
Here the initial leak rate evaluated at the bubble point pressure is higher than the steady inflow to the pipeline, hence the pressure drops almost instantaneously to the bubble point pressure.

Method B2.

The initial release rate exceeds the steady pipeline inflow rate but is less that the maximum rate at which fluid “enters” the two-phase region. Under these conditions depacking can take a finite time to occur.

PBREAKL indicates the type of flow solution and depacking calculation which is used and tabulates rupture gas and liquid rates, the location of the two-phase interface, the total mass removed from the rupture, the rupture pressure and physical properties. All of this data can be plotted graphically.

Some examples of PBREAKL program predictions are given in Figures 5 to 8 for the case of an 8” nominal bore pipeline carrying NGL fluid at 10 C. The pipeline is 10 Km long and the initial pressure is 30 bar at a feed rate of 25 kg/s. Note the different depacking characteristics in Figure 6.



Time of pipe rupture = t_o = 0.0

Time at which bubble pressure reached = t_{Bub}

Figure 4. Types of discharge time history in PBREAKL

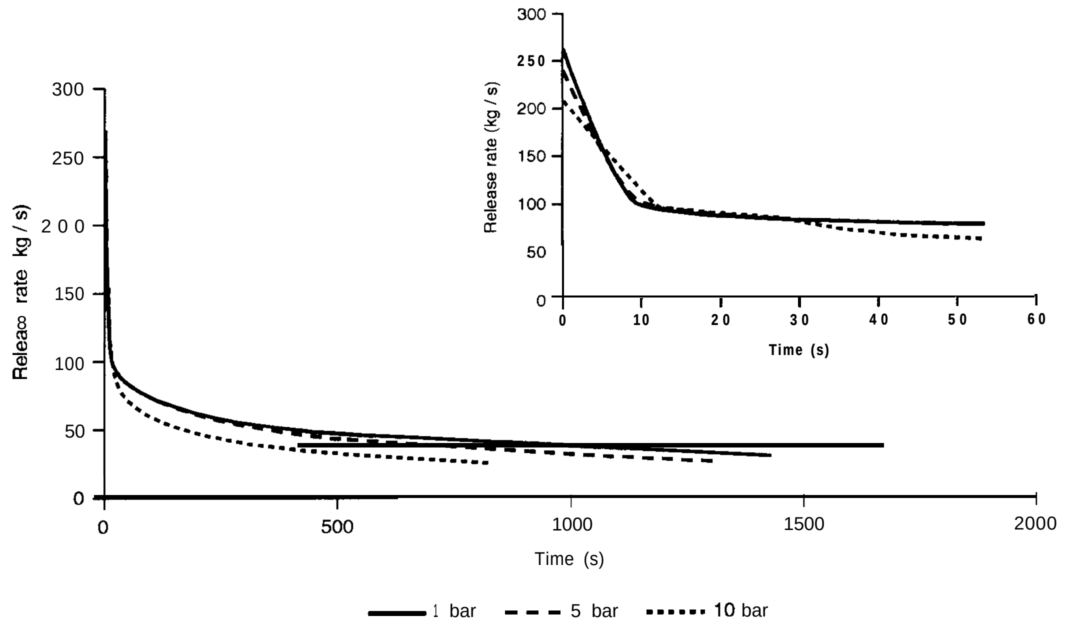


Figure 5 Effect of back pressure

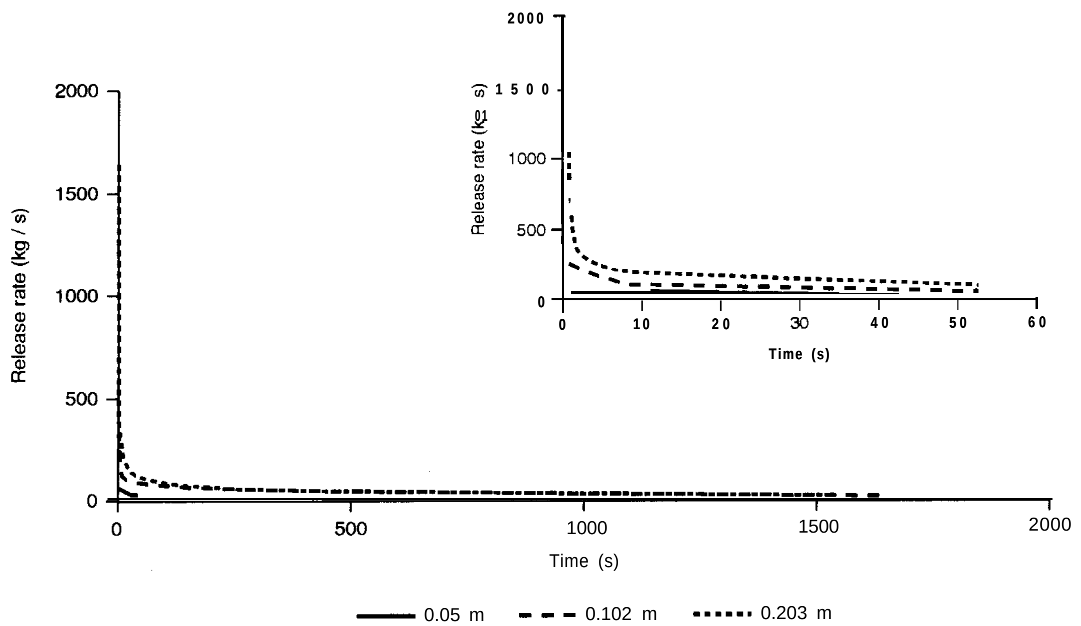


Figure 6 Effect of break diameter

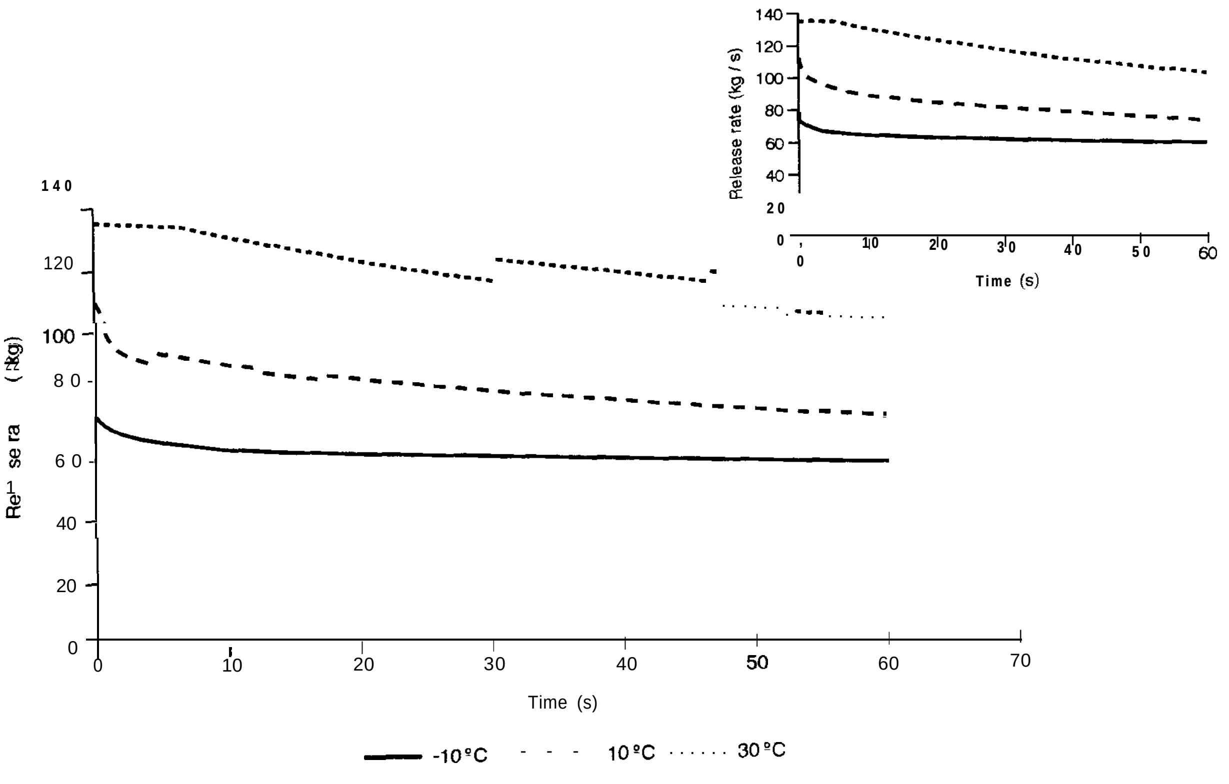


Figure 7 Effect of inlet temperature

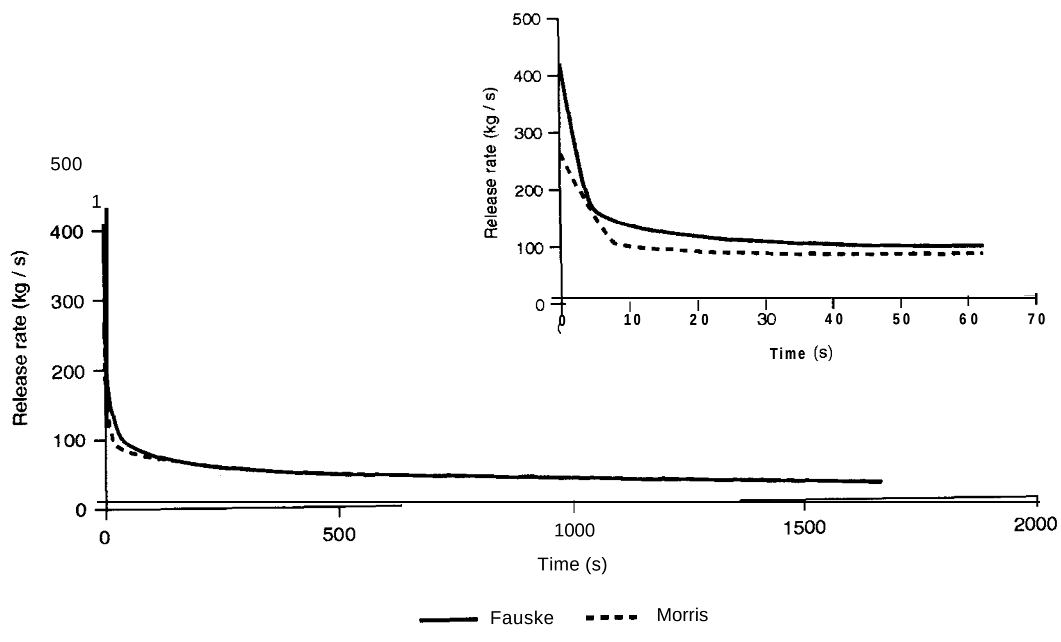


Figure 8 Effect of choked flow correlation

Experience with running PBREAKL has shown that the rupture rate is most sensitive to the break diameter and the bubble point pressure. When carrying out simulations it is recommended that various sensitivity runs are performed in order to identify the range of results and to identify critical parameters. The amount of non-condensible gases in the composition, such as nitrogen, should be checked as this can have a large effect on the bubble point.

PBREAKL is a relatively easy code to use but does have limits of applicability which are listed below. Violating these assumptions may not necessarily lead to poor results, but indicates that the code is being operated outside the range for which it is designed. If the assumptions inherent in PBREAKL are too restrictive a more sophisticated code such as PLAC or OLGA should be considered. This is particularly true if detailed topographic and thermodynamic effects are important.

9.2.1 Limits of applicability of PBREAKL

- The pipe must be long compared to its diameter (length to diameter ratios greater than 100 are recommended).
- The flow conditions upstream of the rupture are largely unaffected by conditions down stream of the rupture.
- The flows can be represented by homogeneous lumped masses.
- Pressure wave and flow inertia effects can be neglected.
- Each pipe section can be considered straight with no significant discrete loss components in the pipeline.
- Friction losses at the rupture can be neglected.
- Conditions in the two-phase region of flow are governed by an isenthalpic flash process.
- The inlet boundaries can be represented by a constant mass flow.
- The flow through small breaks are unlimited by compressibility effects at the rupture.
- The rupture does not vary in size during the analysis.
- Heat exchange to or from the pipeline can be neglected, isothermal flow is assumed.
- The contents of the pipeline are initially all above the bubble point.
- Some of the functionality of PBREAKL is presently available in the PC based loss and consequence model CIRRUS which is supported by the HSE group.

9.3 Factors Affecting the Transient Analysis of Pipeline Ruptures

The transient analysis of pipeline ruptures and depressuring can become very complex depending on what is required from the analysis and to what accuracy. In some cases simple calculations may suffice, but in others large errors can result from the most minor assumptions. The analysis is usually required to determine the liquid and gas flowrates from a rupture with time for the assessment of jet fires and total volume spilled, or for gas release dispersion calculations and determining environmental impacts. The latter is usually in order to study the effectiveness of safety valves. When the fluids are mainly gassy the temperature drop during depressuring can also be a critical aspect of the analysis due to the temperature limits of pipeline materials and components.

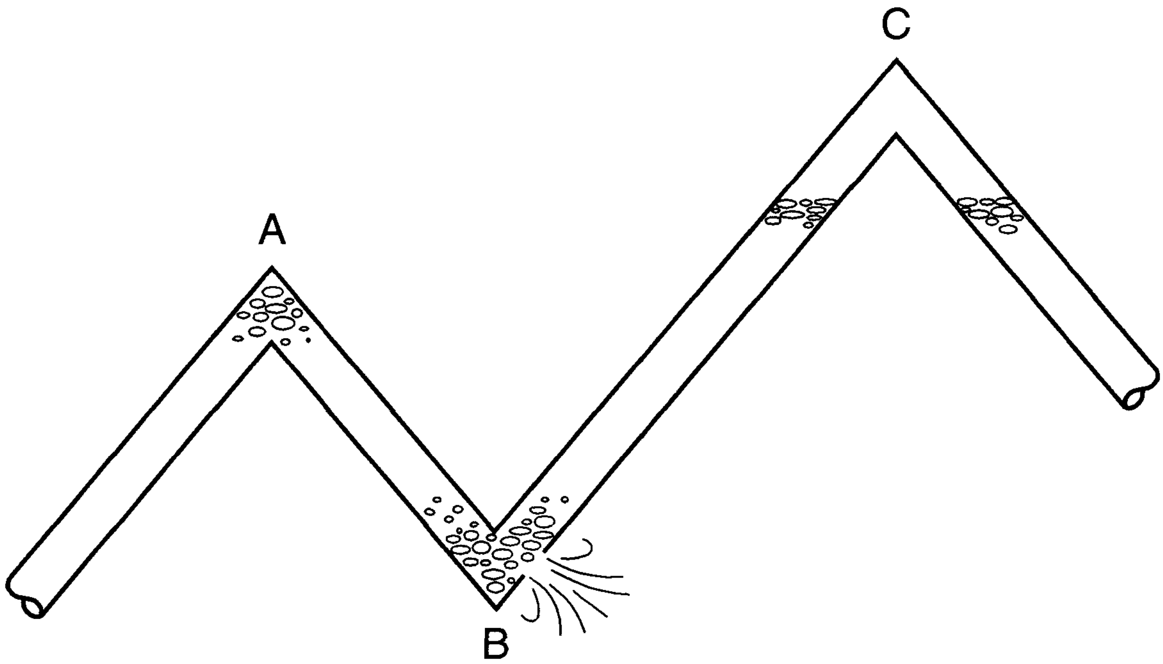


Figure 9. Hilly terrain pipeline schematic

In order to illustrate some of the factors affecting the spill size from a rupture in an oil pipeline, consider the hilly terrain pipeline in Figure 9. A simple approach would be to assume that during depressuring the bubble point pressure is reached at the high points A and C and hence the oil spill size is equivalent to the pipeline inventory between these two points. This volume could be modified by performing a flash calculation from the initial pipeline operating pressure to atmospheric pressure in order to account for the evolution of gas. In practice some liquid could be siphoned over the hills towards the break and on reaching the bubble point the level swell or “champagne” effect produced by gas evolution could result in significant quantities of extra liquid flowing into the dip ABC. The Champagne effect can be halted if downhill elevations are present, since the hydrostatic pressure increases and can rise above the bubble point. The bubble point at the interface may be reduced by the previous evolution of the solution gas. It is hence seen that the analysis is becoming quite complicated without consideration of thermal effects, flowing conditions or the characteristics of the rupture.

Most ruptures are assumed to be circular holes with orifice characteristics, where as in practice the rupture may be jagged and the flow path may be obstructed by the surrounding soil, for example. Because of the difficulty of analysing these effects a free jet issuing from a circular hole is usually assumed. The effect of the back pressure on the efflux rate can be modelled by the codes but the fluid/fluid interaction between the jet and the surrounding medium is ignored. In most cases the critical two-phase flow models are of acceptable accuracy for the calculations required, with the Morris method probably most accurate although not as widely used as the Henry-Fauske method. It is often useful to investigate the upper and lower limits of the critical flowrate using the homogeneous equilibrium and homogeneous frozen models respectively which are available in PLAC.

Work is underway with some of the sophisticated transient two-phase flow codes to enable the individual components of the fluid stream to be tracked. In the case of pipeline blowdown, the lighter components may be expected to leave the pipe in the gas phase during the early stages of depressurisation, with the composition becoming progressively heavier with time. Another situation where composition effects may occur is also when liquid condenses in a dip, leaving the less heavy components in the gas phase. Present developments with some of the codes are attempting to account for this non-equilibrium between the phases.

9.4 Rupture Case Study – Bruce Gas Export Pipeline Temperature Effects Resulting from a Pipeline Rupture

The Bruce gas export pipeline is 32" diameter and 5.8 km long and transports processed Bruce gas from the PUQ platform to a tie-in to the Frigg system, which consists of 360 km of 32" pipeline. In the event of a rupture of the gas export pipeline, an SSIV located 193 m from the platform will close, protecting the facilities if the rupture is on or near the platform. The expansion of the gas at the rupture causes a drop in the temperature, and hence there was a requirement to determine the temperatures at the valve when actuated, to check that the valve stem material remained above the brittle temperature. It was also required to provide valve temperatures for the estimation of the starting torque on the spindle. If the valve stem has cooled sufficiently to become brittle, then when operated it may break, making it impossible to close the valve and prevent the pipeline depressuring.

Under normal steady flowing conditions with a gas export temperature of 55 °C the temperature at the location of the SSIV is 48.5 °C at a pressure of 151 bara. If a pipeline rupture occurs the gas will be expanded to a pressure of 12.0 bara, equivalent to the hydrosystic head of water. An isentropic expansion over this pressure drop yields a final temperature of -76 °C compared to an isenthalpic final temperature of -15 °C. The minimum allowed temperature of the ball valve stem material is -30 °C, hence it is seen that based on isentropic assumptions the gas temperature could be much lower than the minimum allowed. As the pressure drops liquid is condensed in the pipeline giving a liquid mass fraction of around 10% at the break condition. A depressuring line is shown on the Bruce export gas phase envelope in Figure 10, indicating that for most of the blowdown the fluid state is two-phase. This prediction is based on the OLGA simulation described below and includes friction losses as well as heat transfer.

The actual temperature of the valve components can be significantly higher than the flowing fluid temperature due to the heat transfer between the fluid and the inside surface of the valve, and due to the thermal inertia of the valve material which will require sometime to cool by conduction. In addition the surrounding sea water may play a significant role in maintaining temperatures above the brittle temperature.

A dynamic two-phase blowdown simulation was carried out using the OLGA code to assess the temperature of the inner wall of the valve. Initially the simulation was run to steady state to determine the initial pressure and temperature along the pipeline and specifically at the rupture point and the SSIV. The rupture was assumed to be 15 m downstream of the SSIV. The steady state OLGA predictions are shown in Figure 11 where the platform is approximately 6 Km from the Frigg tie-in.

In the first dynamic simulation the program was re-started with the pipe section downstream of the rupture location removed and a fixed pressure boundary of 12.9 bara imposed, hence simulating a full bore rupture. In practice 1 second is allowed for the pressure to fall from the steady line pressure to the seabed pressure, hence the rupture is developed over this time. The ball valve begins to close after 91 seconds and is fully closed after 121 seconds.

Figure 12 shows the temperatures in the vicinity of the SSIV where it is seen that a minimum fluid temperature of -32 °C is obtained during the blowdown and rises rapidly as the valve begins to close. The temperature at the upstream face of the valve rises much more rapidly than downstream as the hotter gas is supplied from the reversed flow in the pipeline. The minimum inner valve wall temperature is -3 "C, which is significantly above the brittle limit, even though the gas temperature is at times below -30 °C (Figure 13 shows the corresponding pressure plots). Even if the valve fails to close Figure 14 shows that a long time would be required for the inner surface of the valve to drop to -30 °C, and a substantially longer time would be required for the valve stem to cool to this value due to conduction through the 6" thickness of the stainless steel ball. The predicted minimum temperature at the rupture point is -45 C using the OLGA code which is halfway between the value calculated assuming isentropic and isenthalpic expansion processes.

A final point about the valve starting torque is worth mentioning. Valves are usually tested in a pressure and temperature controlled facility in which the torque is measured with the whole valve cooled to say -30 C at the pipeline operating pressure, in this case around 151 bara. There is a time lag between the initiation of the rupture and the start of the valve closure, by which time the internal pressure has significantly reduced when the valve is actuated, in this case down to 20 bara, hence the starting torque is much reduced. In addition, the main seals around the ball that contribute to the torque are usually shielded by a debris control ring which will provide some thermal lag. The bulk valve will also be significantly warmer due to the effect of conduction and heat transfer from the surrounding sea water.

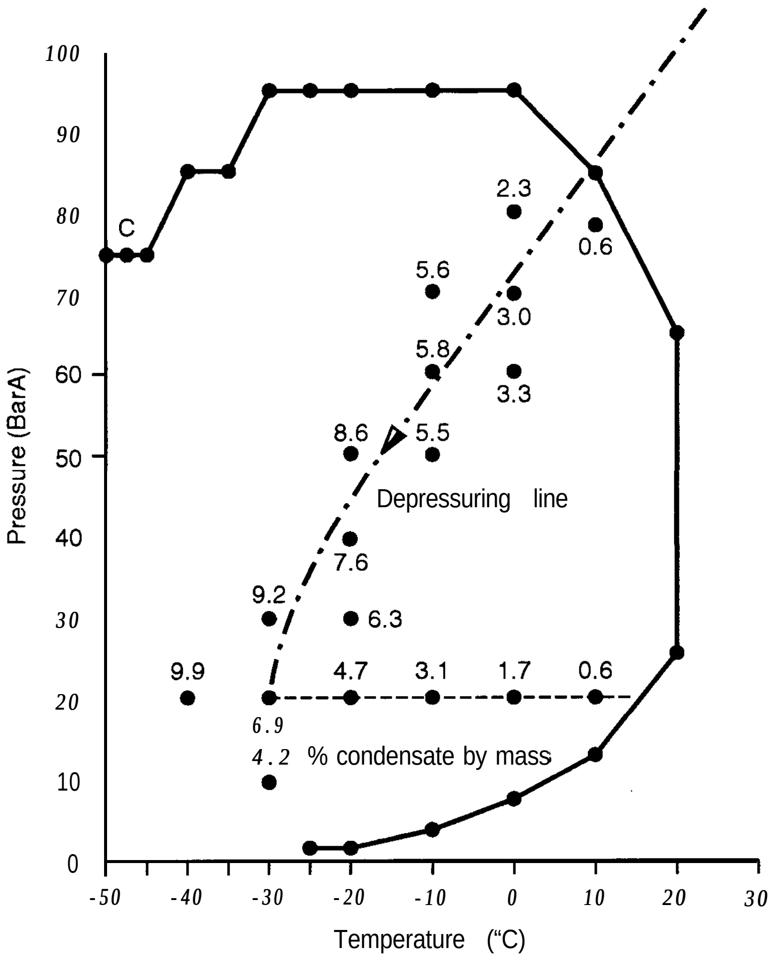


Figure 10. Depressuring line and phase envelope for Bruce gas

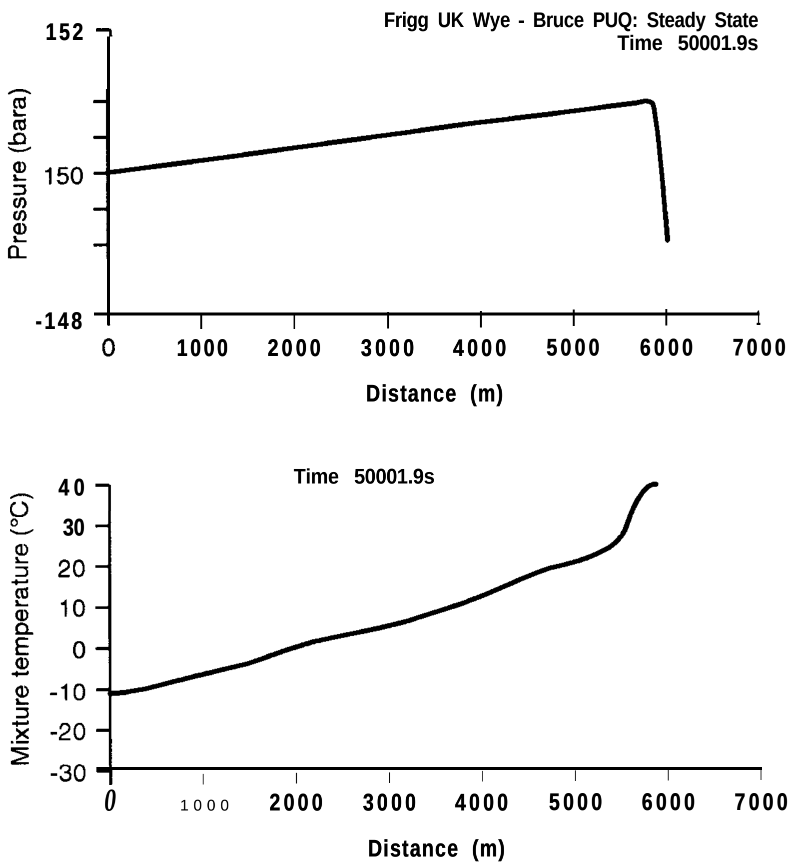


Figure 11. Steady state OLGA predictions

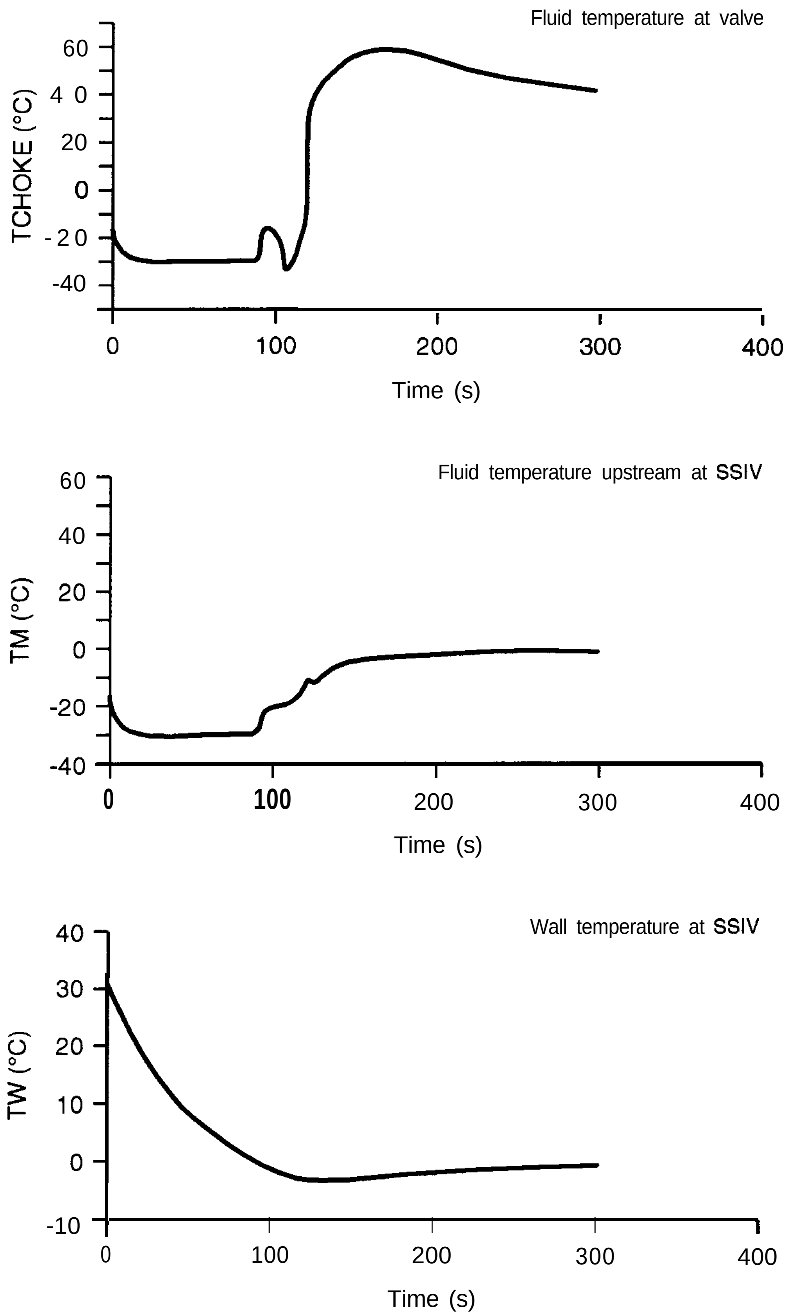


Figure 12. Temperatures in the vicinity of the SSIV

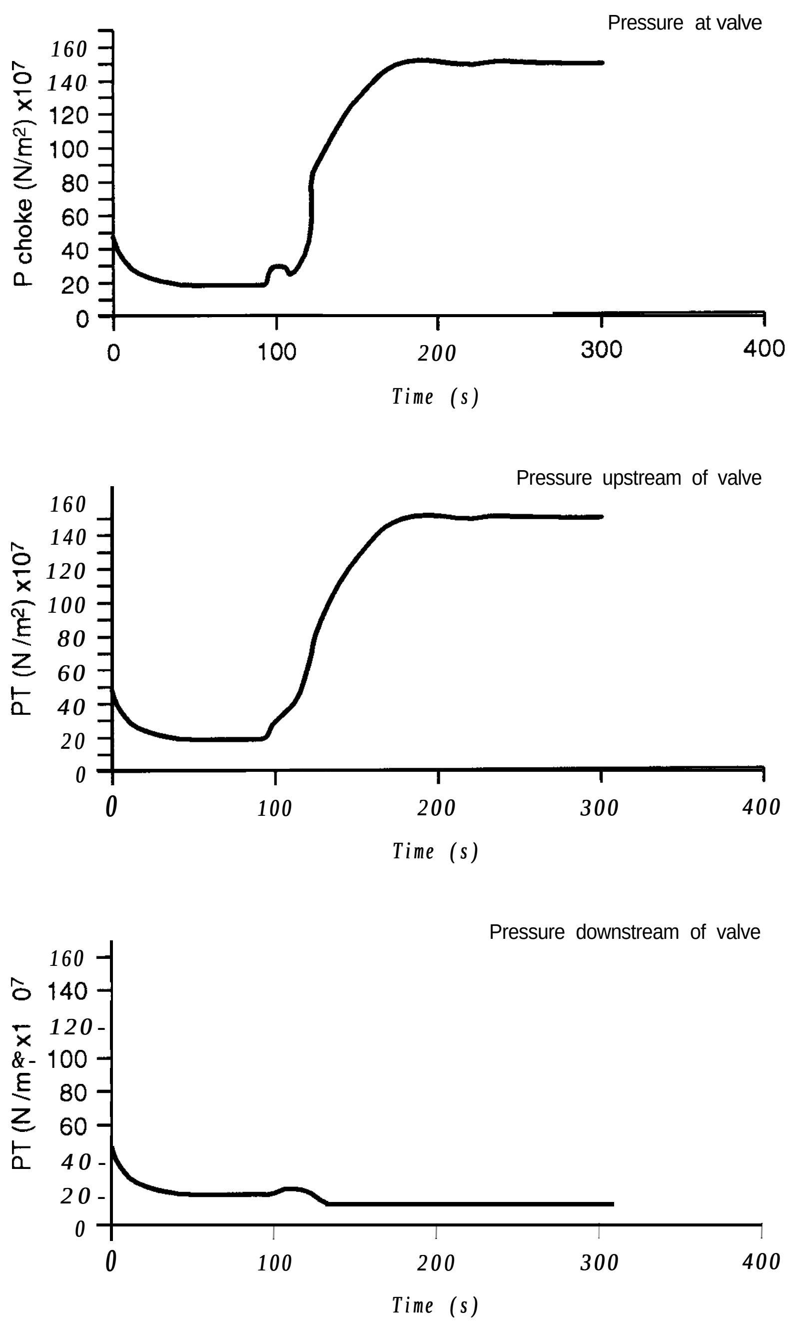


Figure 13. Pressures in the vicinity of the SSIV

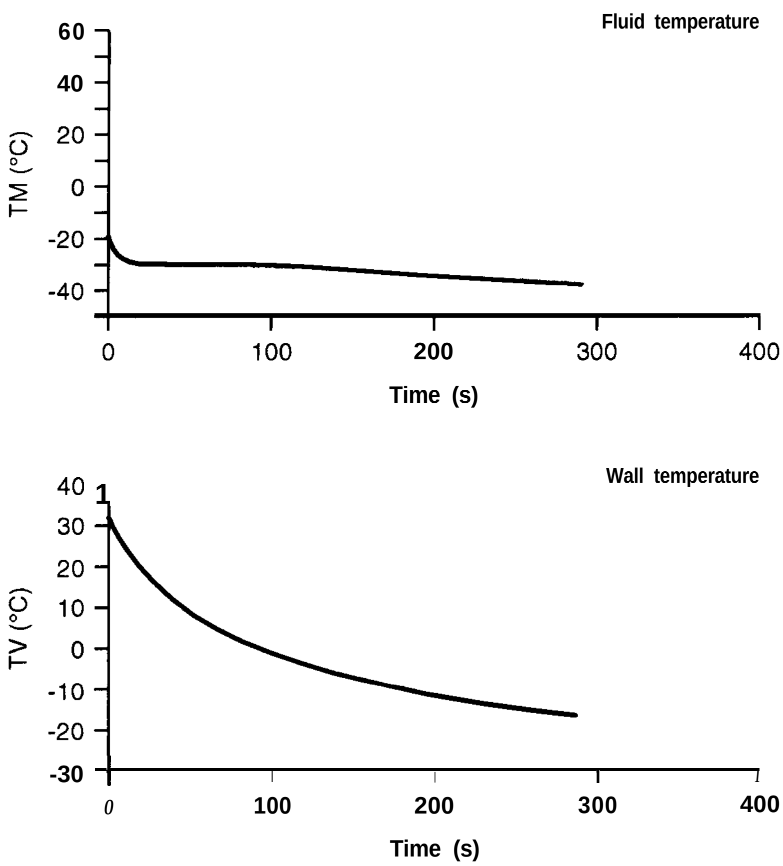


Figure 14. Temperatures downstream of the SSIV with no valve action

References

(1) Henry R.E and Fauske H.K "The Two-Phase Critical Flow of One-Component Mixtures in Nozzles, Orifices and Short Tubes", Trans. ASME-Journal of heat transfer pp 170-l 87, may 1971.

(2) Morris S.D and Daniels L.C "Two-Phase Flow through wellhead Choke Valves A State of the Art review", AERE report G.3710, Oct 1985.

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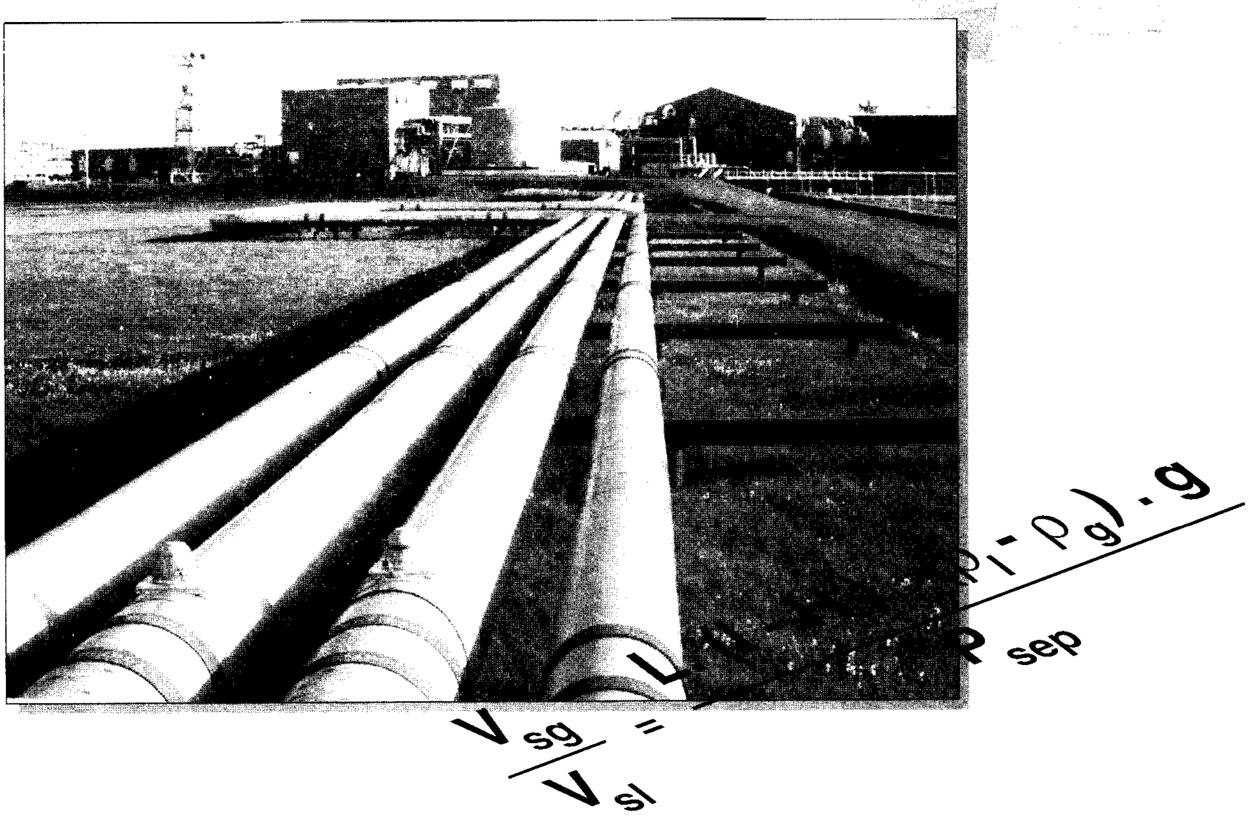
Section 10. Downflow

Design of single and two-phase gravity downflow systems



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10.1 Background

Single and two-phase gravity downflow occurs in a number of areas in BP's operations. The principal examples of such systems are:

- Produced water overflow lines from an offshore separator system.
- Cooling water return lines from offshore exchangers and condensers.
- Rain and firewater collection systems from, for example, offshore helipad decks.

Over the last 5 years the Multiphase Flow Group has been asked to investigate operating problems and design methods for a variety of downflow systems. As the amount of design data for such systems was very limited, and their operation poorly understood, XTC sponsored a programme of experimental work at Sunbury.

A large scale transparent test rig was constructed and operated by the Multiphase Flow Group to study downflow in a number of pipework configurations. This work is reported fully in (1).

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10.2 Basic Design Approaches

A number of components of the design of gravity flow systems must be specified.

- Liquid may be supplied to the top of the downcomer pipe by a header tank, or by a pipe of the same, or different, diameter.
- The bottom of the downcomer pipe may be open to atmosphere, or submerged beneath liquid.
- The top of the downcomer pipe may, or may not, connect to atmosphere.

The behaviour, or flow regime, of the flowing liquid will depend on the liquid flowrate, the combination of the above inlet/outlet options, and the diameter and routing of the actual downcomer pipe. The two stable regimes are single phase liquid, in which the whole of the downcomer is full of liquid, and annular liquid, in which the liquid flows as a film down the inside wall of the downcomer, with a central gas core.

The selection of the combination of inlet/outlet options will be influenced by the source of the liquid flow, the desired flow regime, and safety considerations related to pipework support and/or failure consequence analysis.

There are five principal recommended design options, given the assumption of pipe discharge to atmosphere:

- Single phase downflow from pipe supply.
- Single phase downflow from header tank supply.
- Annular liquid flow from pipe supply (top of downcomer closed).
- Annular liquid flow from pipe supply (top of downcomer vented).
- Annular liquid flow from header tank supply.

These options are illustrated in Figure 1 and discussed in detail in the following sections.

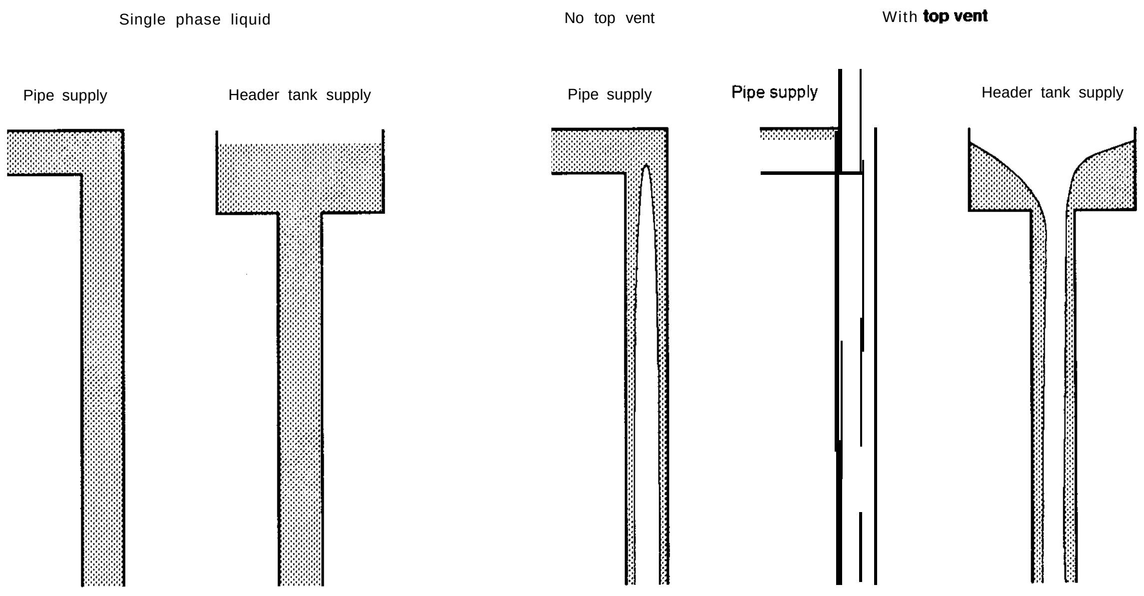

Figure 1. Downflow design options.

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10.3 Single Phase Downflow

Downflow gravity systems can be designed to operate in single phase liquid flow. This type of approach is often applied to the design of high capacity firewater drainage systems. However, in most cases it is not possible to ensure full pipe flow at all times. For example, a firewater drainage system will generally be maintained in a dry condition. When the system is called upon to dispose liquid it will go through a transient stage during which gas displacement will occur.

There are two design cases that can result in single phase downflow. The difference between the two is the source of the liquid. In both cases, once the gas initially present in the pipe has been displaced, the flow is single phase liquid through the system.

10.3.1 Supply by Header Tank

If the liquid is supplied to the top of the downcomer by a header tank then it is necessary to maintain a liquid depth in the tank high enough to produce the liquid flowrate required to keep the downcomer inlet, and the whole of the downcomer length, flooded.

(a) For downcomer entry at the base of the header tank

To avoid gas entrainment, liquid height, h, needs to be:

$$h \geq 0.89 \left[\frac{G^2}{\rho_L^2 \, g \, D} \right]^{0.25}$$

where:

h = liquid height in vessel above outlet (m)
D = outlet pipe diameter (m)
G = mass flow (kg/s)
ρ_L = liquid density (kg/m^3)

(b) For downcomer entry at the side of the header tank

$$h \geq 0.811 \frac{G^2}{(g \, D^4 \, \rho_L^2)}$$

where:

h = liquid height above the top of the downcomer entry.

In order to determine the capacity of single phase downcomer systems supplied by header tank, an energy balance is applied between the inlet, point 1, and outlet, point 2.

$$P, + \frac{\rho U_1^2}{2} + pgh = P_2 + \frac{\rho U_2^2}{2} + 0 + \text{friction} \qquad \textbf{Equation 1}$$

The static pressures acting at points 1 and 2, P, and P,, are both equal to atmospheric pressure. U,, the downward velocity in the header tank, will generally be very much less than U,, the liquid velocity at the pipe outlet, and so the above equation reduces to:

$$pgh = \frac{p U_2^2}{2} + \text{friction}$$

Friction losses are the sum of losses over the pipe, and over any fittings.

$$\text{Friction} = K \frac{\rho U_2^2}{2} + f \frac{L}{D} \frac{\rho U_2^2}{2}$$

where:

f = Darcy-Weisback friction factor
L = pipe length
D = pipe diameter
K = total number of loss coefficients for fittings.
so:

$$pgh = \frac{\rho U_2^2}{2} + K \frac{\rho U_2^2}{2} + f \frac{L}{D} \frac{\rho U_2^2}{2}$$

$$gh = \frac{U_2^2}{2} \left[1 + K + f \frac{L}{D} \right] \qquad \textbf{Equation 2}$$

Hence, it is necessary to ensure that the required capacity (i.e. outlet velocity U,) can be achieved with the available head, h.

As already discussed, a drainage system designed on the basis of single phase flow, will probably pass through a stage where gas is being displaced by the liquid. The system must then be designed to handle slugging flow which will occur as gas is swept from the pipework (see below).

10.3.2 Supply by Pipe

If the liquid is supplied to the top of the downcomer by a pipe then the design requirement is simply that the liquid flowrate is high enough through the downcomer to initiate flooding and prevent de-flooding.

On the basis of the BPX tests this flowrate is highly dependent on the geometry (routing) of the downcomer. For a straight vertical downcomer with a pipe inlet the required Froude number to initiate flooding is 1.07. For a downcomer that includes a horizontal section the required flooding Froude number varied, with position of the horizontal section, from 1.21 (section near bottom) down to 0.85 (horizontal section near the top of the downcomer).

The Froude number is the quantity:

$$V_{sl} / \sqrt{gd}$$

where:

V_{sl} = superficial liquid velocity (m/s)
D = pipe diameter(m)

So in order to design for single phase flow it is useful to have a horizontal section of pipe in the downcomer, near the top of the system. This will give the greatest turndown on flowrate whilst maintaining single phase flow.

During the period when the liquid flow is filling the downcomer there will be a transient stage of gas displacement from the line that may result in surging and slugging, and some mechanical vibration to the pipe. This must be allowed for in siting pipe supports.

In this case, where feed to the downcomer is directly from a pipe, the full energy balance given above in Equation 1 still applies, but the static pressure at the top of the pipe, P, is not necessarily equal to atmospheric pressure.

If P, is greater than atmospheric pressure, then the capacity of the downcomer will be increased to a value in excess of that given by Equation 2.

However, if the flowrate of liquid into such a system is reduced to a value below that given by Equation 2, then the pressure at the top of the downcomer, P_2, will fall to a value below atmospheric pressure. As the system being considered here has no vacuum breaker (ie is not vented) there is the danger of a very low pressure developing in the downcomer.

This risk of developing low pressures is greater in a system where the exit is below sea level, as there is no opportunity for air to be sucked in to the bottom of the downcomer to help break the vacuum.

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10.4 Annular Liquid Flow

Liquid flows down the downcomer as an annular film, with a gas core in the centre of the pipe that is open to atmosphere at the bottom of the downcomer. The downcomer pipe may, or may not, also be vented to atmosphere at its top end. The liquid flowrates achievable are different depending on whether there is a top vent.

Offshore water overflow systems, such as cooling water return lines, and produced water downcomers, are generally designed to operate in the annular flow condition.

10.4.1 Without Top Vent

This case occurs with a pipe supply to the top of the downcomer. The classical design method used to ensure that a downcomer remains in annular flow is to maintain the superficial liquid velocity (Vsl) below a critical value given by:

$$V_{sl} < 0.35 \sqrt{gd}$$

The term $V_{sl} / \sqrt{gd}$ is referred to as the Froude number, Fr. So, for annular liquid flow systems the relationship becomes:

Fr c 0.35

The experimental work at Sunbury was aimed at determining the criteria for maintaining annular flow in various piping configurations, and for downcomers whose exits are above sea level, and for those which are submerged.

The experimental studies illustrated a number of points that had not previously been quantified. These are:

- The capacity of a vertical downcomer is significantly greater than that of one which has bends and dog-legs.
- The capacity of a downcomer is reduced by submerging the exit below the water line.

Note that the capacity of a downcomer in this context is defined as either, or both of, the liquid flowrate leading to flooding, i.e. liquid fills at least part of the downcomer, or the liquid flowrate which causes air to be drawn down the downcomer.

Based on the work at Sunbury, our recommended maximum flowrates which avoid flooding and air removal are as follows:

Downcomer configuration	Maximum Froude No. exit above sea level	Maximum Froude No. exit submerged
Vertical	**0.57**	**0.3**
With bends/dog-legs	**0.35**	**0.3**

It can be seen that the presence of bends significantly reduces the downcomer capacity. Ideally, bends should be avoided. If bends are unavoidable any near horizontal section of pipework should have a minimum inclination of 1 in 40. To avoid slugging in any inclined lines it is usual to specify that lines run no more than 1/2 full of liquid (for pipe internal diameters < 0.2 m). For pipes > 0.2 m in diameter, the line can be allowed to run up to 3/4 full. Liquid depths are calculated using the relationship :

$$V_L = \sqrt{32gmi}\ \log 10 \left[\frac{e}{14.8m} + \frac{0.22k}{m\sqrt{gmi}} \right]$$

where:

V_L = mean actual liquid velocity (m/s)
g = 9.81 (m/s*)
m = hydraulic mean depth = area for flow / wetted perimeter (m)
i = inclination of pipe to horizontal
e = pipe roughness
k = kinematic viscosity (m^2/s)

With this type of system in which only the bottom of the downcomer system is open to atmosphere, the gas (air) in the core is only circulated very slowly.

10.4.2 With Top Vent

This case can occur with a vent line put into the top of the downcomer supplied by a pipe, and with a header tank system in which the flow is insufficient to maintain a liquid level in the tank, or if the tank has a hollow annular flow stabiliser. The behaviours of the systems are, however, very similar.

The experimental work at Sunbury illustrated that allowing air to be drawn down the downcomer in a top-vented system increases the amount of liquid that could be handled, i.e. increases flowrate to a level in excess of the rates given in the table in Section 10.4.1. Air is drawn down through the core by the liquid flow, thereby giving a rapid replacement of the core.

For a header tank without a hollow annular flow stabiliser the flow limit will be reached when the liquid bridges the top of the downcomer in the tank.

For the cases with a vent line (vented pipe supply, or header tank with hollow annular flow stabiliser) the limit will be the onset of surging caused by some liquid bridging in the downcomer.

In the experimental work at Sunbury, as the liquid flowrate increased, a point was reached where a slugging/surging type of flow was produced.

10.5 Transition From Annular Liquid Flow

As the liquid flowrate is increased in a system operating in annular liquid flow, a point will be reached at which liquid bridging occurs at one or more locations in the downcomer. Liquid slugs are formed that move downwards through the system.

10.5.1 In Non Top-Vented Systems

In systems that have no top vent the slugs will begin to pull a vacuum as they fall, beginning the process of gas displacement from the downcomer. The gas is sucked downwards, and its previous position occupied by the liquid flowing into the top of the downcomer pipe. Some slugs of liquid will propogate right through the downcomer, drawing gas with them out of the bottom of the system. Other slugs will break up, leaving a churn flow region in the downcomer.

If the liquid flowrate is increased sufficiently, a single phase liquid flow will eventually be produced. The instability of annular flow at these higher liquid flowrates is amplified if there are any restrictions in the downcomer, or reductions in diameter.

In a practical situation this type of flow might also cause liquid in any of the lines feeding the downcomer to surge intermittently (as the slugs decrease the pressure at the top of the downcomer) and this, in turn, could affect the operation of the plant producing the liquid. Apart from the effect on the plant feeding the downcomer, slugging in the downcomer will produce out-of-balance forces which will require the pipe to be well supported, especially at bends. If a downcomer is operated at rates above the maximum values given in 10.4.1, slugging flow can be expected.

Calculation of the velocity of a falling slug requires solving a complex force balance. The Multiphase Flow Group, BP Exploration should be consulted if such a calculation is required. Once a slug velocity is determined the force on any bend is calculated using the equation given in Section 3.4.1.

10.5.2 In Top-Vented Systems

In systems with a vent into the top of the downcomer, the effect of liquid bridging on the plant supplying the liquid is much less of an issue. The pressure at the top of the downcomer remains constant at atmospheric. The only change will be in the gas rate drawn into the downcomer through the vent.

Any liquid slugs formed will, however, still have the potential to cause vibration to the pipework system, and the caution on design of supports, as given in the previous section, still applies

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10.6 Operating Experience

If there is any chance that hydrocarbon vapours could enter the system (for example, through entrainment in the cooling water from a leaking exchanger) it is essential that the entire downcomer retains the central gas core so as to enable explosive gases to escape through a safe route to atmosphere. In the design of the Bruce platform, the FEED contractor undersized the downcomer at 48". When it was realised that a 60" downcomer would be needed to maintain a self-venting system, it was too late to alter the design. As a consequence, hydrocarbon vapour detection equipment had to be installed in the downcomer to warn of the presence of any hydrocarbon gases being drawn down through the system as a result of a cooling water exchanger leak.

In the Clyde produced water system the liquid flowrate was in excess of the values given above in 10.4.1. As a result slugging and surging conditions were experienced, and this led to intermittent backing up of liquid in the lines that feed the system.

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Reference

(1) Single Phase/Two Phase Downflow Studies, Brister, D, and Turner, A L, Sunbury Branch Report 124 179 dated 28.6.91

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Section 11. Piping Layout

Mechanical Design of Pipelines and Supports for Multiphase Flow

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11 Piping Layout

This section describes a number of ways in which the pipework layout may influence the behaviour of the flow, and other ways in which the flow will affect the behaviour of the pipelines themselves.

The influences of pipeline inclination angle, and of line diameter, are described in Section 3, in the text on flow regimes and slug flow. This section (11) contains the current understanding on more geometrically complex issues.

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11.1 Converging Networks

Pressure loss calculations for converging networks may be made using the MULTIFLO programme.

The principal issue here is how the upstream flows from separate lines influence the behaviour of the combined downstream flow.

Each line section in the network should undergo a flow regime prediction based on the in-situ gas and liquid superficial velocities in that section. Where the upstream flows are predicted to be not intermittent, but the downstream flow is intermittent, then the lines into and out of the point of combination may be treated as separate.

However, if one or more of the upstream lines is predicted to be in slug flow then this may have an effect on the nature of the downstream flow. If the downstream flow is predicted to be stratified then it may be forced into slug flow because of the incoming slugs. This case may occur where the downstream line is of a larger diameter than the incoming lines. If the downstream line is predicted to be in slug flow then the slug frequency in this line may be affected by the frequency of the incoming slugs. This will be most likely in systems where the incoming slugging line is the same diameter as the outlet line.

The main example of converging networks of interest to BP is in the Kuparuk field in Alaska (1). During one period of observation, slugs arriving at a junction of two 16 inch lines with a 24 inch out flow line were seen to decay, not propogating down the 24 inch line.

Further work is required to quantify the areas of the flow regime map (for a line downstream of a junction of two or more lines) that are susceptible to influence from upstream conditions.

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11.2 Diverging Networks

11.2.1 Occurrences

In general the main requirement of flow division by a diverging network is to split the incoming gas and liquid flow equally between two branches of the same diameter. This occurs where flow from a pipeline or manifold needs to be divided between two or more separation trains, and also in the inlets of finger-type slugcatchers.

Another instance of diverging networks is found in gas transmission systems, in which small amounts of condensate may be present - the requirement is to know where the condensate will appear. This second example is not of current interest to BP. The key research work on this topic has been done by Oranje (2).

11.2.2 Equal Division of Flow

For a diverging network, the key points to note are that in order to divide a multiphase flow equally the following provisions must be made:

- A dead-end tee configuration must be used, with the two exit arms both having the same diameter

- The inlet line must be straight for at least 15 diameters upstream of the tee

- The exit arms should be horizontal, with subsequent downstream pipework also horizontal if possible

- Exit arms should follow parallel routes downstream, to maintain equal pressure loss

- If the exit arms require a significant (10-15 D) downwards elevation change after the tee then a line diameter equal or larger than the inlet line to the tee should be used, to prevent potenttial imbalance due to bridging (see Section 10) in the downflowing lines

- Exit arms should not be routed upwards after the tee

These recommendations have arisen out of three pieces of work conducted in BP. Firstly, the evaluation of flow splitting problems on Forties platforms Alpha and Delta (3); secondly, the evaluation of a design for a new production manifold on Forties Alpha (4); thirdly, a general study of splitting at a tee at varying distances downstream of a bend (5).

These tests involved a number of different tee/bend configurations, as illustrated in the isometric drawings below.

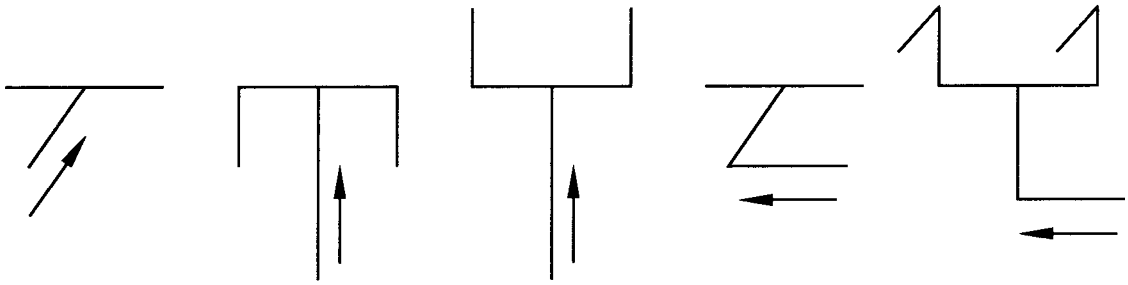

Figure 1.

If the exit arms rise soon after the tee it is possible that very slight imbalances in the liquid head in the rising sections may feed back to the tee and affect the splitting. If the arms drop after the tee it is possible that a siphon effect may be set up, also affecting the flow division at the tee. It is therefore recommended that the exit arms be made horizontal, or downwards of larger diameter if a significant downward elevation change is required.

A bend in the same plane as, and close to, the tee may cause very significant maldistribution, especially of the liquid (up to 6:1). To eliminate this effect the inlet line to the tee should be straight over a length of 15 diameters.

To ensure good operability over a wide range of gas and liquid flowrate combinations, all these design recommendations should be followed.

11.3 Forces on Bends

It is necessary to know the forces that might be applied on pipework by internally, and externally, flowing fluids, in order to be able to design adequate supports and restraints.

11.3.1 Internal fluid flow

Any fluid or moving object that is constrained to change direction requires a force to be applied to make it change direction. If the fluid has a steady density, and is moving at a steady velocity, then the force required is constant, and may be evaluated from a force/momentum balance.

For multiphase flow pipelines in stratified, bubble or annular flow the fluids have a reasonably steady density and velocity. The changing momentum of the liquid phase is the principal contributor to the force exerted on bends. In these cases the liquid momentum is not that great as the flowing velocities are not very high. In a bubbly flow pipeline it would be very unusual for the liquid velocity to be above 3 m/s. In annular flow and some high flowrate stratified flows the fluid velocities are much higher than that of bubbly flow. The gas velocity is much higher than that of the liquid film. However, the average fluid density is much lower than in bubbly flow due to the small liquid fractions in such flows. A steady, and moderate, force is applied by the pipe on the fluids to change their direction.

However, in slug flow there are two differences from the other flow regimes. Firstly, the flow density is not steady, but varies between a low liquid hold-up in the film region, to a high hold-up in the slug body. Secondly, the velocity of liquid in the slug is roughly equal to the combined gas and liquid superficial velocities. Therefore the force required to change the direction of a slug (high velocity and density) is much higher than the forces required during the other regimes. In addition, the intermittent nature of the flow means the force is coming on and off at frequent intervals. This can lead to pipework vibration and the possibility of fatigue of the pipework supports.

The slug frequency and holdup may be predicted using the methods available in the BP multiphase design code, MULTIFLO (shortly to be re-issued as the UNIX code THoR). The detail on the calculation of forces due to slugging around bends is given below in Section 11.3.2.

There have been instances (Prudhoe Bay, Alaska) of slug flow causing pipelines to jump off their supports. These cases occurred at expansion loops (a series of four 90 degree bends in a horizontal plane). The force required to change the direction of the slugs is provided by a flexing of the bends, which are not fixed to their supports because of the thermal expansion require ments. In some cases the movement of the pipe because of the flexing bends has been so great that the pipe moved off its supports.

In most cases the cause of the pipe leaving the supports could be attributed to unusual operations rather than normal slugging flow. In these cases a shut-in (but pressurised) line was opened to a depressured vessel causing rapid acceleration of the liquids accumulated in the low sections. The forces that can be generated are greater than a reasonable support design can accommodate. The problem was solved using operational procedures as well as safety interlocks to prevent opening major valves with an unbalanced pressure.

In one flowline the piping jumped from the supports during normal operations on three separate occasions. This line was a tie-line diverting excess production from one production facility to another. The line was operating at high velocities and would have been predicted to be operating in annular flow. The velocities were higher than would have been considered for slug restraint design. One possible theory is that a slug developed in one of the individual flow-lines feeding this tie-line and was passed through the tie-line at these high velocities. This scenario hasn't been proven but it highlights the need to consider high velocity slug flow in similar "net work' systems.

Pipe vibration is also noticeable on Forties Alpha where the 12 inch line from Forties Echo passes through a bend upstream of the pig receiver, and then on into the production manifold.

In addition to rigid near-horizontal lines, the other systems in which the forces exerted by slugs are significant are those involving flexible risers. Flow-induced vibrations may affect the life of the riser, and of the riser connections and supports. For reference BHR group have undertaken a number of studies of severe slugging in flexible risers. Further discussion on the behaviour of flexible risers under forces exerted by internal (and external) fluid flow are to be presented in Section 20 of this manual (Multiphase Flow Through Deepwater Systems).

11.3.2 Forces due to slugging

(a) Near horizontal lines

The force, Fg, exerted by the flow of gas only through a bend would be given by:

$$F_g = \sqrt{2}\, \rho_g\, V_g^2\, A$$

where:

F_g = force exerted by gas only (N)
ρ_g = gas density (kg/m^3)
V_g = gas velocity (m/s)
A = pipe cross sectional area (m^2)

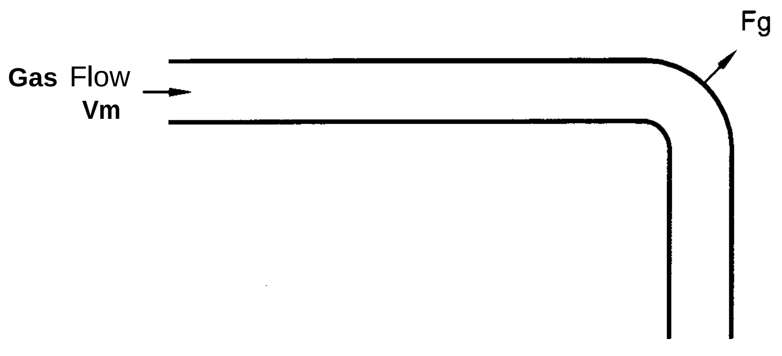

Figure 2

If the pipe bend is carrying liquid only flow then the equation to determine force exerted is of the same form, but written as:

$$F_l = \sqrt{2}\, \rho_l\, V_l^2\, A$$

where:

F, = force exerted by liquid only (N)
ρ_l = liquid density (kg/m^3)
V, = liquid velocity (m/s)

with $\rho_l > \rho_g$ and V, < V,, and a typical F, usually greater than a typical F_g.

When slug flow occurs in an essentially horizontal line the mean velocity of the liquid in the body of the slug is equal to the mixture velocity, V_m, hence:

$$V_s = V_m = V_{sg} + V_{sl}$$

where:

V_s = mean slug velocity

It is this velocity, V_s, which should be used when evaluating the forces imposed by a slug as it travels through a bend.

As a slug passes through the bend at velocity V_s, the total composite force exerted on the bend, F,, is given by :

$$F_t = \sqrt{2}\,\rho_s V_s^2 A$$

where:

ρ_s = slug density

with $\rho_s \sim \rho_l$, $V_s \sim V_g$, and therefore F_t > F,.

If the slug hold-up is H,,, then ρ_s is given by :

$$\rho_s = \rho_l H_{ls} + \rho_g (1 - H_{ls})$$

The difference between the two steady forces F_t and F_g is the difference between the force required to change the direction of the fluids in the predominantly liquid slug, and the force required during the gas bubble. This is a little conservative as the gas bubble will usually contain some liquid flow. On slug arrival into the bend this change in steady force required is termed the slug force, F_s, and is given approximately by:

$$F_s = \sqrt{2}\,(\rho_s - \rho_g) V_s^2 A$$

Depending on piping configuration a considerable portion of this force may result in bending the pipe, giving a ‘pipe-spring’ which will then be released when the force due to the slug passage is removed as the slug passes out of the bend.

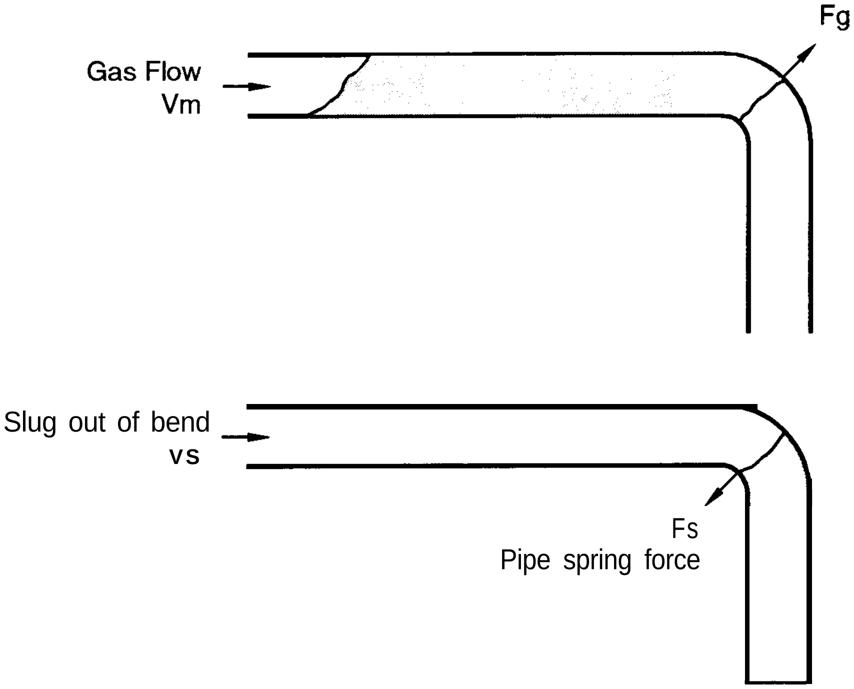

Figure 3

These forces are based on normal quasi steady-state slug flow. They are not adequate to analyse the dynamic problems previously described at Prudhoe Bay. In the largely dynamic situation mentioned there are also large unbalanced pressure forces that occur as the slug is accelerated.

It should be noted that in determining design loads for fixed supports to withstand these forces due to slugging, a dynamic factor of 2 (minimum) should be used unless a rigorous dynamic analysis is performed (as discussed below). This results in:

$$F_d = 2\sqrt{2}\,(\rho_s - \rho_g)\,Vm^2\,A$$

The dynamic effect of slug loads on bends is illustrated in the following figure. This shows a load cell output for a bend subject to slug flow. Note the short duration dynamic load (roughly two times the steady load) when the front of the slug enters the bend, and also when the end of the slug leaves.

The nature of slug flow means that these forces produce a tendency for the bend to move firstly in one direction, as the slug enters the bend, and then in the opposite direction as the slug exits the bend.

If the piping is not adequately designed then these forces may cause excessive pipe movement or stress, and this may make the pipe and supports susceptible to movement induced fatigue.

(b) Pipeline Risers

It was mentioned in Section 3.3 that although a slug will progress through a riser at the down stream end of a pipeline, it will tend to decelerate as the line upstream of the slug packs up to provide the pressure to overcome the increasing hydrostatic head in the riser.

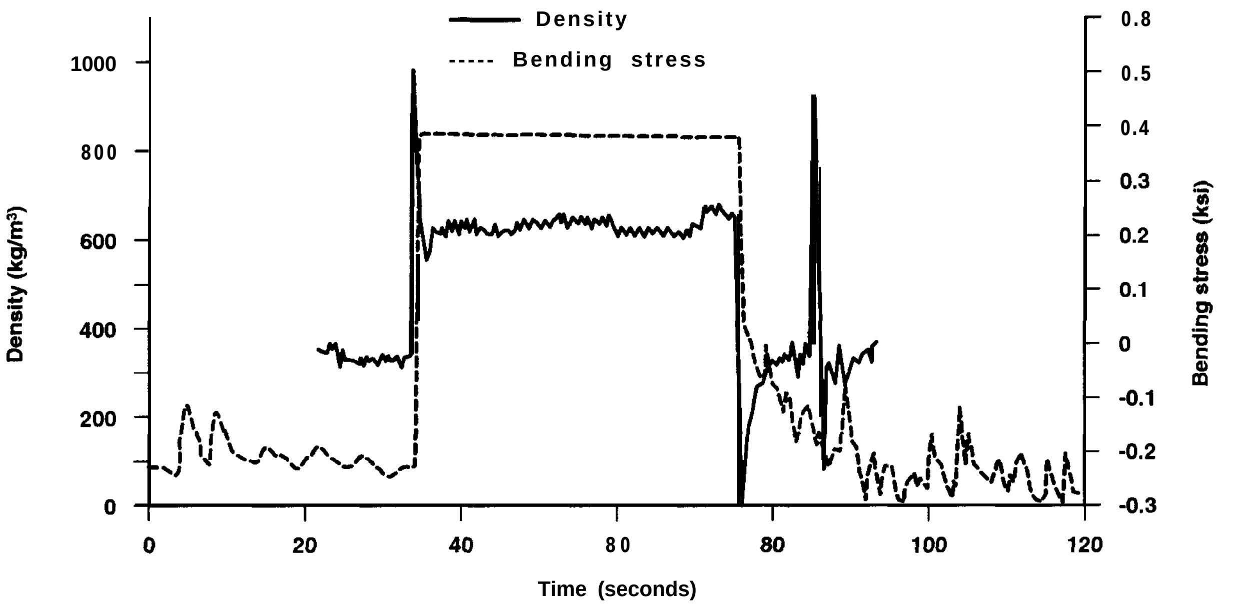

Figure 4. Slug flow around a 90" bend.

As the slug leaves the riser the hydrostatic head loss in the riser reduces so that the upstream gas bubble expands and accelerates the slug into the process plant. This phenomenon is described in more detail in Section 3.

In order to model the effect of slug flow in a pipeline- riser, BPX have developed a dynamic program known as NORMSL. This can be used to determine the velocity of the slugs as they pass through the riser and into the topsides pipework. This information can then be used in the determination of forces exerted.

In order to assess the interaction of slugging with the topsides process plant the output file from NORMSL can be used as the input to a dynamic simulation model of the topsides plant. Process dynamic simulations can be performed using such software as the SPEED-UP and PLOT programs, and there is now a link between NORMSL and both SPEED-UP and PLOT. In this way any pressure changes in the process plant arising from the production of slugs or gas bubbles are fed back directly into the pipeline-riser model, thereby giving a better prediction of the slug velocities. This feedback effect becomes more significant as the height of the riser increases, i.e. for deep water developments.

(c) Areas for Further Work

Analysis of test rig data is continuing and will be used to improve the slug frequency model, which will affect fatigue predictions.

Collection of data from hilly terrain pipelines operating in normal slug flow is required to validate/modify the existing models for slug frequency and velocity. Collection of data from the Cusiana field should go some way towards achieving this aim.

A very simple model is currently used for determining slug liquid hold-up. Work is required to develop a model which more closely reproduces the observed data. In particular the model will need to account for the water content of the oil.

11.3.3 Support and Restraint Design Notes

For the initial design there is often adequate data providing the piping engineer where to look and what to look for. The biggest problem is to ensure that the piping engineer knows that the line will be dynamically loaded. In many cases' field problems are the result of either break down in communication between process and piping engineers or unexpected dynamic loads that exceed or visibly test the inherent factors of safety. Hence it is important to ensure an efficient transfer of data and information between the fluid system designer and the piping designer.

There are several analysis techniques available to avoid or analyse these dynamic loads, and some dynamic effects may be approximated by static loads.

The piping layout should contain adequate guides and line stops where practicable and long unguided sections should be avoided.

The piping layout connecting to relatively rigid pipework and equipment (manifolds, vessel nozzles, etc.) should be carefully designed to protect these locations from the dynamic slugging forces. Possible solutions include:

- Adequate rigid restraints and anchors to protect these locations.

- Ensuring that force points (bends, etc.) are some distance from the critical location allowing both piping flexibility and less robust restraints to provide protection for the equipment.

Where reasonable, piping should be supported or restrained at all heavy masses such as valves, etc.

Include a dynamic factor of 2 minimum on design loads for fixed supports when using simple static analysis to calculate forces.

Avoid or minimise number of spring hangers.

11.3.4 Piping Stress Design Notes

- **Do not use un-reinforced branch connections or fittings.**
- **Do not use threaded connections.**
- **To perform a quasi-static analysis:**

Calculate momentum slug forces as previously outlined. If conservative values for density and velocity were not used in calculating the slug force an additional 25% safety factor should be included. (Note that velocity is the most critical factor and as a result determining a safe but reasonable maximum velocity is very important).

Perform steady state stress analysis of the possible load cases. There are several different force combinations depending on slug length and location of the slug (arriving and departing). For a piping system with n bends there are *n* load combinations assuming a single slug provided it is possible to have slug lengths longer than the distance from the first bend to the last bend. Several of the cases may be trivial or redundant so that a limited number of the cases require further analysis.

Total stresses should be calculated and compared to allowable stresses. The cyclic (slug force) and steady state components of stress should be calculated for use in fatigue life cal culations.

Calculate permissible fatigue cycle-life using the appropriate code (BS 5500 for the UK) and steady state and cyclic stresses calculated as outlined above. Estimate fatigue life based on slug frequency and duration.

This analysis does not include the possibility of resonance. Resonance due to normal quasi-steady-state slug flow is unlikely in that the slug frequency is not constant even during steady state flow. Slug lengths and frequencies are very distributional rather than constant.

11.3.5 Field Problem Solving

When confronted with a specific case of pipeline response to the dynamic loads, a first step is to assess the need for immediate action. In a few cases the situation may be sufficiently serious to cause a shutdown for correction or measures to limit the undesirable effects. On the other hand it is frequently possible and entirely adequate to reduce pipe movements by provid ing temporary braces, supports, ties, snubbers, or similar restraints, so long as care is taken to avoid possible thermal expansion difficulties on shutdown or during temporary operational upsets.

More permanent corrective measures include an assessment of the movement(shakes) charac teristics by field measurements and designing a permanent corrective solution.

Accurate measurements of piping deflections and frequencies can be used to gauge the severity of the problem and determine whether action is required. For low frequency (very flex-ible) systems adequate measurements can generally be taken by simple manual methods (Don't trust your visual impressions without actual measurement against some nearby stationary object). High frequency rigid systems require more sophisticated measuring tools such as accelerometers.

These deflections and frequencies can be used in a steady state analysis to estimate maximum and cyclic stresses. Areas of high stress intensification should be carefully assessed when con sidering fatigue live. These include branch connections, nozzles, equipment connections, girth welds, etc.

At this point a permanent solution can probably be determined. It is possible that a more rigorous analysis is required; involving dynamic stress analysis, strain gauge measurements, etc.

There are several, well established, state-of-the-art piping flexibility programs, such as ADLPIPE, Autopipe, PSA5, CAESAR II, CAEPIPE, etc. which can be used to evaluate the dynamic responses of the system and can be used during field investigations.

There are numerous other considerations (e.g. type of pipe material, local stress raisers, type of welds, strength of supporting structure, etc.) which are considered during the field 'trouble shooting'. It must be pointed out that any field solutions to an existing piping system suffering from the effects of dynamic loads should be preferably left to an experienced piping engineering specialist.

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11.4 Diameter Changes, and Fittings

This section is concerned with the forces exerted on pipework by multiphase flow passing through diameter changes and fittings. The significance of the diameter change or fitting is related to the flow regime, as was the case in the previous section.

Slug flow is the principal source of problems, especially where there are flow restrictions or diameter reductions. The problems are two-fold. Firstly, a force is required to accelerate the liquid into a restriction and to overcome the increased pressure gradient through it - this force is provided by an increase of pressure of the upstream fluids, and has a reaction that is met by the pipe wall and the pipe supports. The reaction force is intermittent, matching the flow behaviour, thereby creating the possibility of pipework/support fatigue.

The second problem arises from the compression of the upstream gas. As described above the upstream gas pressure has to increase in order to accelerate the liquid into the restriction and overcome the increased pressure gradient through it. The gas is therefore packed to a higher pressure and density. As the end of the slug enters the restriction the acceleration force is no longer required. When the length of slug contained within the restriction or section of reduced diameter pipe starts to decrease (e.g. as the front of the slug passes out of the end of the restriction) then the pressure drop across the slug begins to drop. However, due to the packing of the gas there is now excess pressure available upstream. This excess pressure accelerates the slug.

The severity of the consequence of this is dependent on the position of the restriction relative to other restrictions or diameter increases. For a restriction (e.g. a partially closed valve) located some distance (several slug lengths) away from other influences then the overall effect of a slug passing through is just a force on the pipework containing the restriction, and some variation in the velocity of the slug.

However, if there is a diameter reduction a distance of the order of a slug length upstream of the separator, then the effects may be serious. The upstream gas pressure rises to force the slug into and through the restriction. As the slug passes into the separator the pressure drop across it reduces - but the upstream gas pressure then exceeds that required to overcome the pressure drop, and is therefore available to accelerate the remaining length of slug. This acceleration continues until the end of the slug passes into the vessel. The upstream gas pressure is still likely to be higher than that in the separator, resulting in a rush of gas into the separator.

This process is occurring in Kuparuk field, Alaska, where, in order to save money on the initial installation, two lines were reduced from 24 inches to 20 and 18 inches, for distances of 150 and 30 metres respectively, upstream of the separator (1). The cost saving was achieved by having smaller valves. However, the separator internals have had to be replaced on seven occasions due to the mechanical damage caused by slugs being accelerated into the vessel. The gas compressors are also continually coming in and out of recycle to cope with the gas flow surges.

Designers of multiphase flow systems should minimise the number and extent of restrictions (and elevation changes), especially over the last few hundred metres of line. On no account should the line diameter be reduced within several slug lengths of the separator.

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11.5 Elevation Change

The effect of an inclination change is similar to that of a diameter change, in that the pressure gradient of the flow is altered. The consequences are described in the section on terrain and geometry dependent slugging (Section 3.4.2)

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11.6 Slugcatcher/Separator Inlet Nozzles

This section is an extension of Section 11.4 (Diameter Changes and Fittings) in that it deals with the final section of pipework leading to the nozzle of the receiving vessel.

The main recommendation is that account be taken in the design of pipe supports of the load imposed by slug flow. The number of bends close upstream of the inlet nozzle should be kept to a minimum. Ideally the inlet line should pass the fluids straight into the nozzle.

In Kuparuk field there has been an incident of weld cracking at a vessel inlet nozzle. The configuration of the inlet pipework is shown in the following figure. There are two 90° bends close upstream of the nozzle. Slug flow through these imposes a bending moment due to the bend in the vertical plane, and a torque due to the bend in the horizontal plane. There is some scope for compensation of the bending moment by flexing of the vessel shell. However, there is no scope for compensation of the torque other than by stress in the nozzle weld. Eventually this weld failed.

It should also be noted that such a configuration may also result in maldistribution into the two 'halves' of the vessel if the distance of the arm marked Z-Z is less than 10–15 diameters (see Section 11.2.2).

The remedy to this case was to construct a pipe support to take the loading on the bend in the horizontal plane, thereby preventing any further movement and stress on the weld

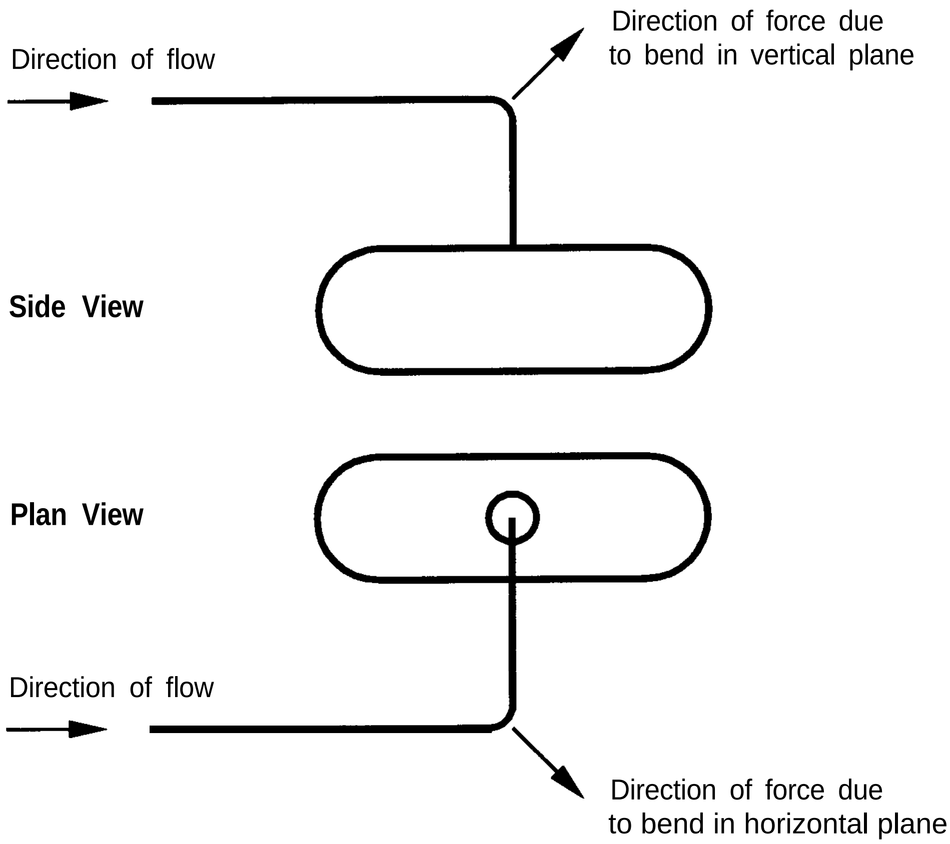

Figure 5. Kuparuk CPF-3 Case

An alternative inlet configuration is shown in the Figure 6, where there are no bends immediately upstream of the inlet nozzle. However, this type of vessel (end inlet) will have different separation characteristics to that shown in Figure 5 (central inlet), and so the balance of requirements between separation and inlet piping needs to be considered.

To design for the pipe support loadings on any bends in the upstream pipework system it will be necessary to produce force-time diagrams for the pipework configuration based on the nature of the slug flow through the pipework. The Multiphase Flow group can provide the necessary slug characteristics information.

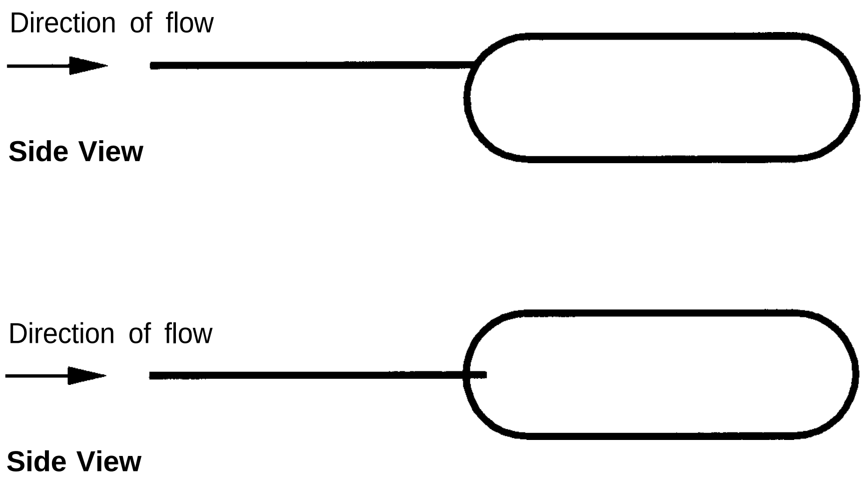

Figure **6. Alternative configuration**

11.7 Liquid Distribution Through Bends

The main area where this topic has provoked interest is in the distribution and stability of corrosion inhibitor on pipewalls. At Prudhoe Bay problems have been experienced with excessive corrosion just downstream of bends (in the vertical plane) of 30, 45 and 90 degrees (6).

Investigative work is at an early stage and recommendations on the optimum bend configuration have yet to be finalised. Initial experimental work has been carried out for a road crossing configuration, where the pipeline drops down to go under a road. In this instance there is usually a liquid build-up at the base of the riser at the downstream end of the road crossing.

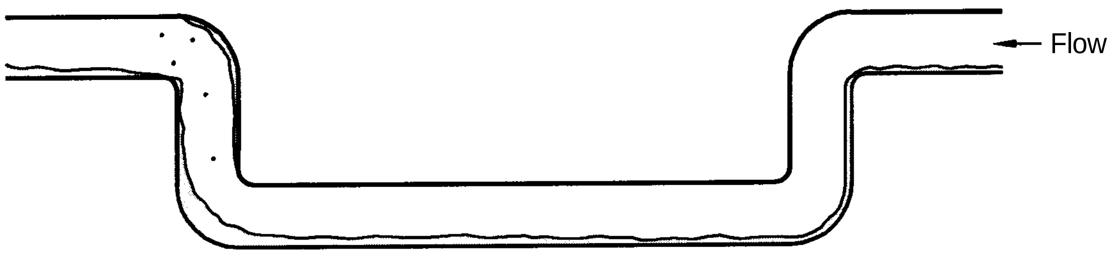

Figure 7.

This liquid build-up reduces the gas flow path area, causing acceleration of the gas to a high in-situ velocity. This velocity is sufficient to entrain liquid as droplets, and accelerate them towards the gas velocity. The droplets then impact on the inside of the bend at the top of the riser. If the drops contain any sand this will add to their inhibitor-stripping, and erosive, potential.

Bend angles (turning up from horizontal) of 90, 45, 22 and 11 degrees have been tested. The restriction of the gas flow path by liquid accumulation occurs at each angle, with some droplet entrainment resulting. However, the extent of droplet impingement at the top of the bend that brings the line back to horizontal is reduced as the bend angle is reduced (7).

Further discussion of corrosion in multiphase systems is given in Section 21 of this manual (Corrosion Prediction in Multiphase Flow).

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11.8 External Fluid Flow

The two main cases to consider are (i) wind flow over pipe suspended between two supports, and (ii) water currents flowing over unsupported sections of pipeline laid on sea, lake or river bed, or around rigid or flexible risers.

For work on these aspects you will be referred to other groups in the Research and Engineering organisation.

Within fields of BP interest in Alaska there have been problems with wind-induced vibration. At Kuparuk field there has been a study of the use of helical strakes to reduce pipeline vibration. More recently damped mass restraints have been used. These consist of weights attached to the pipe with specially designed elastomer bands.

The subject of the interaction of forces due to intermittent internal flows with externally generated forces has not been a major topic of interest to date. Some work has been proposed on the study of the behaviour of flexible risers under slugging flow, in which external currents could be considered.

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(1) 'Multiphase Flow Field Tests in Kuparuk Field, Alaska, November 7th to December 1 st, 1990', Sunbury Report No 21 164 dated 3.5.91.

(2) 'Condensate Behaviour in Gas Pipelines is Predictable', Oil and Gas Journal, July 2, 1973, 39-44.

(3) 'Two Phase Flow Physical Modelling - Flow Splitting at a Tee', Sunbury Branch Report 122 210, dated 30.11.83.

(4) 'Physical Modelling of Multiphase Flow Splitting Effects at a Dead End Tee for Forties Artificial Lift Project', Sunbury Branch Report 123 534, dated 10.3.89.

(5) 'The Splitting of Two Phase Flow Regimes at an Equal Tee, at Varying Distances Downstream of a Bend', Sunbury Branch Report 123 395, dated 28.9.88.

(6) 'Simulation of Multiphase Flow Characteristics Through the Sag River Crossing (Prudhoe Bay)', Sunbury Branch Report 123 713, dated 9.11.89.

(7) 'Prudhoe Bay Road Crossings', All-in-l document issued by CJ Nelson, dated 25.1.93.

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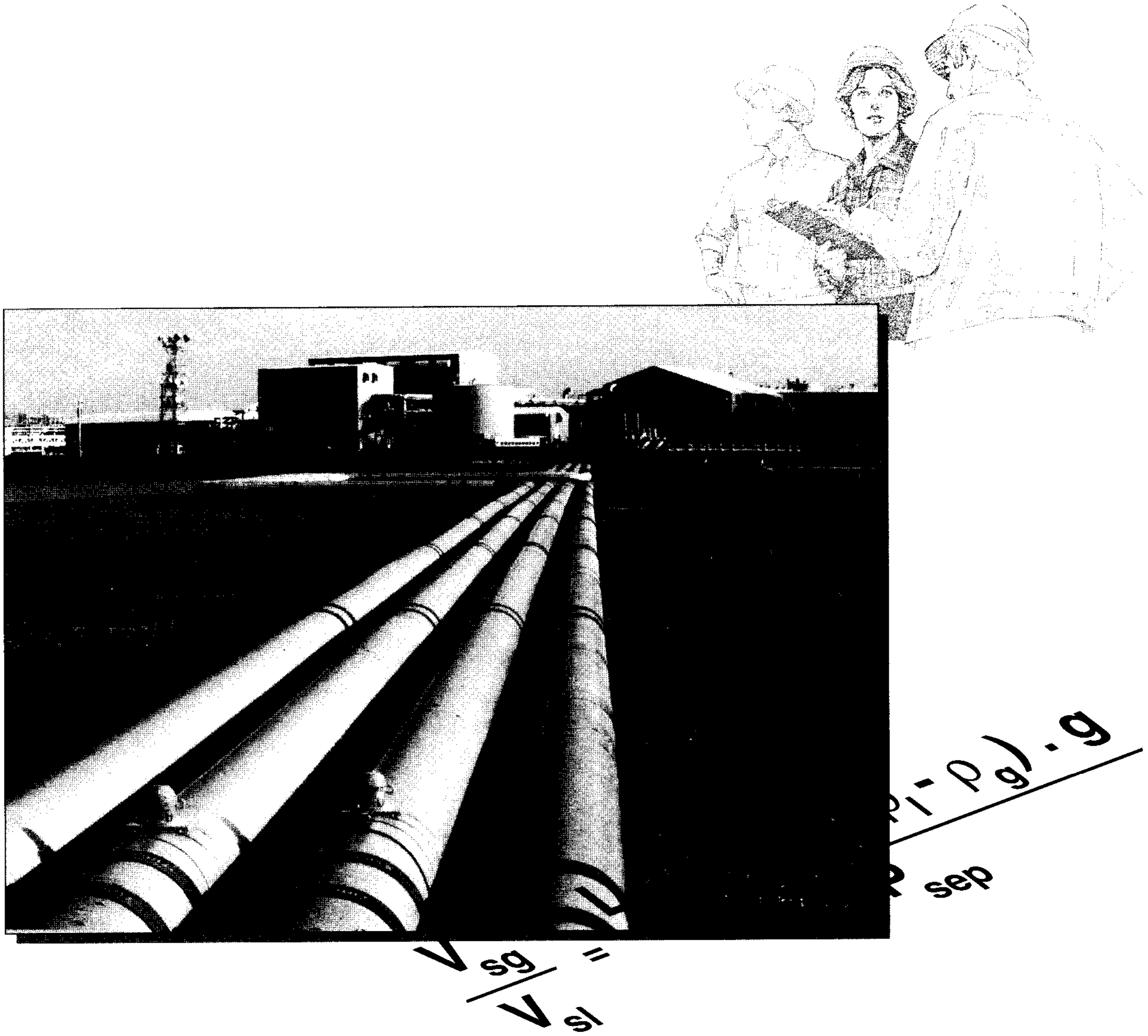
Section 12. Solids Transport

Behaviour of Solids in Multiphase Systems



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12.1 Introduction

The hydrocarbons produced from a reservoir are sometimes accompanied by small quantities of solids such as sand or fracturing materials. For example, the Forties field produces in the region of 5 to 40 pounds of sand for every thousand barrels of oil produced. This sand normally collects in the separators and is either removed by manual intervention during maintenence periods or flushed out using a jetting system. When the South East Forties development was considering using seabed templates connected to the existing Forties Alpha platform by two 5 km pipelines, there was concern that the sand produced might settle out in the pipeline causing pigs to become stuck. The removal of sand may be relatively simple using pigging, provided that only small amounts are deposited. The removal of larger quantities may be difficult and time consuming. To design such systems required knowledge on how the sand is transported and when it will accumulate. This prompted experimental work to be undertaken at BHRA in 1983 (Reference 1). Again in 1990 the Forties Foxtrot development highlighted the need for a better understanding of solids transport in multiphase pipeline systems.

In 1989 the three Ravenspurn South platforms were experiencing proppant production from the well fracturing operation that led to the design and installation of separation facilities to prevent the carry-over of solids into the wet gas transport pipeline to the Cleeton platform. This study required the prediction of conditions to transport the solids steadily to the separators without causing undue erosion in the pipework.

Following from these studies BP has sponsored a project at the University of Cambridge to investigate the flow of solids in two-phase flow pipelines. Data has been collected for sand transport in a horizontal and inclined 40mm ID pipe. This data has been used by the Multiphase Flow Group in XFE to develop preliminary design methods for predicting solids transport conditions in multiphase flowlines.

The results of this preliminary work are presented here as a guide to predicting the critical conditions required to prevent solids accumulating in multiphase oil and gas pipelines. This is necessary to prevent pigs from becomming stuck and to prevent possible corrosion under solid deposits in pipelines. Because of the possible stabilisation of solid deposits by heavy hydrocarbons, and inhibitors, and the potential for accelerated corrosion under deposits, it is recommended to operate multiphase flowlines above the settling velocity to avoid solid deposition, and below critical erosion velocities to limit material loss.

This work is limited to solids which are heavier than the carrying fluids, ie sand and proppant, and may not be applicable to other solid substances formed by chemical reaction such as hydrates, asphaltenes and waxes.

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12.2 Solid Particulate Settling Characteristics and Flow Regimes

Flow regimes for solids transport in liquid/solid and liquid/gas/solid systems are illustrated in Figure 12.1.

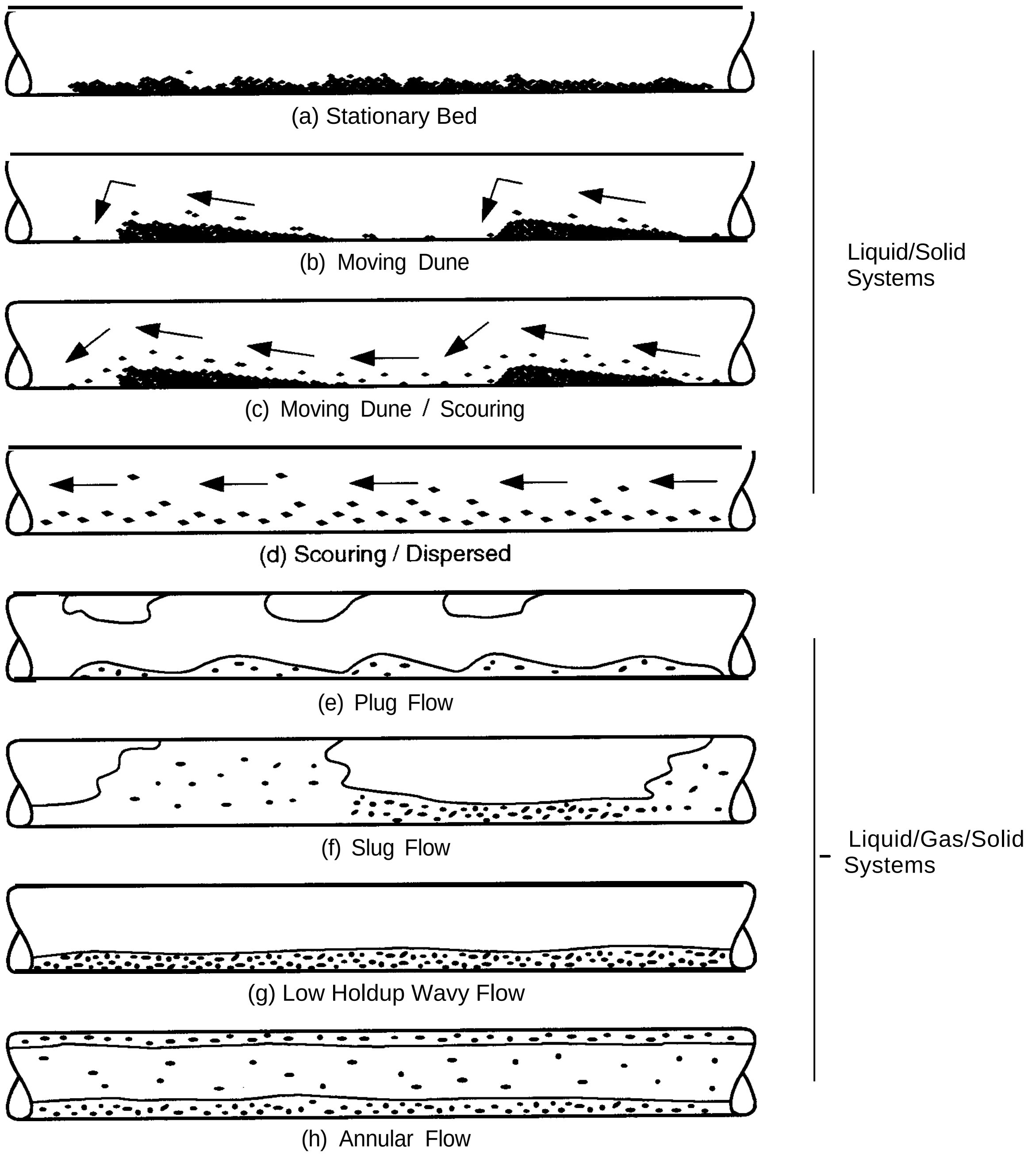

Figure 1. Flow regimes for the transport of solids in multiphase pipelines

12.2.1 Liquid/solid systems:

(a) Stationary bed

At very low liquid flowing velocities a stable solid bed is formed with particles at the bottom and no grains move at all. With an increase in the velocity a stable bed height is reached where the particles at the top are transported further downstream to increase the length of the bed. The upper surface of the bed is flat at very low flowrates but becomes wavy as the flowrate increases. At higher liquid flowrates the height of the stationary bed decreases. An equilibrium bed is reached when the shear at the upper surface of the bed transports solids downstream at a rate equal to the solid inflow rate.

(b) Moving dunes

If the liquid flowrate is increased further the bed breaks up and the particles arrange themselves into moving dunes in which the grains on the upper surface of the dune are rolled along from the shallow back region to the deeper downstream front, where they fall into the sheltered region at the front . The dune passes over these particles until they are once again on the top surface. The motion of dunes is similar to sand dunes in the desert and to snow drifts. Smaller dunes move faster than larger ones and a given length of stationary deposit will break up into a number of dunes, each with a characteristic length and velocity.

(c) scouring

As the flowrate is increased further the grains roll along the top of the dunes with sufficient momentum that they escape from the sheltered downstream region and are swept away as individual scouring grains. Dunes can still survive in this erosion environment by replenishment from upstream particles.

(d) Dispersed

At high liquid flowrates the dunes are dispersed and the flow consists of solids particles bouncing about in the flow and dispersed more in the liquid phase. A strong concentration gradient is still however usually observed.

12.2.2 Liquid/gas/solid systems

Since the solids are heavier than the carrying fluids they are usually transported along the bottom of the pipe when the concentration is low. For this reason the flow patterns observed in single phase solid/liquid flow are similar to those seen in stratified solid/gas/liquid flow since the liquid occupies the lower part of the pipe and the flowing velocity is steady. This is not however the case when the gas/liquid flow regime is plug or slug flow, as the depth of the film and the velocities vary.

(a) Plug flow

In plug flow the gas bubbles can flow along the top of the pipe and have little effect on the solids flow with the full range of regimes already mentioned possible. As the amount of gas is increased the bubble depth increases and the fluctuating velocities affect the transport similar to that described in slug flow.

(b) Slug flow

In slug flow the transport of solids is complicated as the solid may settle during the passage of the film region and may be transported in the slug body. There can be a large diameter effect as the depth of the film varies and shields the bottom of the pipe from the turbulence of the slug. A bed can be formed if the soild is not transported by either the slug or film. In cases where the solid is transported in the slug only the motion is obviously intermittent. The frequency between slugs may be a factor if bed compaction and stabilisation by other products is a possibility.

For slug flow in slightly uphill inclined pipes the solid may be transported backwards due to the reverse flow in the film region, and hence the overall motion of the sand depends on the efficiency of the forward transport by the slug and the reverse motion caused by the film region.

(c) Low holdup wavy flow

In wet gas pipelines the liquid can be transported as a thin film along the bottom of the pipe, in which case the solid concentration in the film can be high, and in the extreme may appear as a wet solid bed. In this case little is known currently about the conditions required to remove the wet solids.

(d) Annular flow

In annular flow the solids may be transported in the liquid film and the gas core. Since the velocities are high in annular flow the usual concern is whether the erosion rate is excessive rather than if the solids will be transported or not.

Several factors can significantly complicate the analysis of the conditions required to prevent the accumulation of solids in multiphase pipelines. These include:

- Three phase flow effects (gas, oil, and water flow)
- Preferential wetting of the solids by another phase ie water wet solids removal by the oil phase
- Bed stabilisation by other products, ie wax
- Effect of inhibitors

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12.3 Model for predicting the limit of a stationary deposit in horizontal and slightly inclined multiphase pipelines

The conditions required to prevent the formation of stationary deposits in multiphase pipelines can be estimated using a method developed by XFE in 1993 (Reference 2). The model is based upon a series of equations derived by Thomas (Reference 3) for calculating the friction velocity at the limit of solid transport in a liquid/solid system. The friction velocity is related to the pressure gradient and has been extended by XFE to the case of transporting solids in multiphase systems by estimating the flowing conditions that give rise to the same pressure gradient that is required to transport solids in the liquid/solid system. The model is hence called the minimum solids transport pressure drop model. The Thomas equations are used to predict the flowing pressure gradient associated with the minimum transport condition in liquid/solid flow where enough energy is passed to a solid particle to enable it to remain in the bulk of the fluid phase and to be transported downstream. Using this pressure gradient, a locus of points can plotted on a two-phase flow pattern map for a constant pressure gradient equal to the pressure gradient at the minimum transport condition.

In the XFE model the two phase pressure gradient is predicted using the method of Beggs and Brill by guessing values for the gas superficial velocities for a given liquid superficial velocity and calculating the two-phase pressure gradient. Iteration is performed until the velocities produce a pressure gradient equal to that for the minimum transport condition caluclated by the Thomas equations for the same liquid flowing velocity. The calculation is repeated for a range of liquid velocities to yield a locus of velocities above which the pressure gradient should be sufficient to transport the solids along the pipeline.

12.3.1 Determination of pressure gradient at the minimum transport condition

Thomas derived several equations for the minimum transport condition depending on whether the solids particle diameter is smaller or larger than the laminar sub-layer in the liquid, and depending on the solids concentration. The first step in the analysis is hence to determine the solids particle diameter and the thickness of the laminar sub-layer, however, since the thickness of the laminar sub-layer depends on the Reynolds number, some iteration is required. We initially assume that the particle diameter is greater than the thickness of the laminar sub-layer and check for this condition after the friction velocity has been calculated.

12.3.2 Particle diameter

It is important to use the correct particle size in the analysis as this affects the calculation of the particle settling velocity and also determines which method is used, depending on whether the particle is smaller or larger than the laminar sub-layer. For single sized particles this is no problem. However in practice the solids produced with oil and gas usually contains a spread of particle sizes. Figure 12.2 shows the particle size distribution for Forties sand and shows that the size varies from around 1 mm to 45 micron. For most cases it is recommended to use the mean particle diameter or d50 value (in this case 255 micron) for the determination of the minimum transport criteria. However it is also recommended to investigate the sensitivity of the results to the particle diameter used.

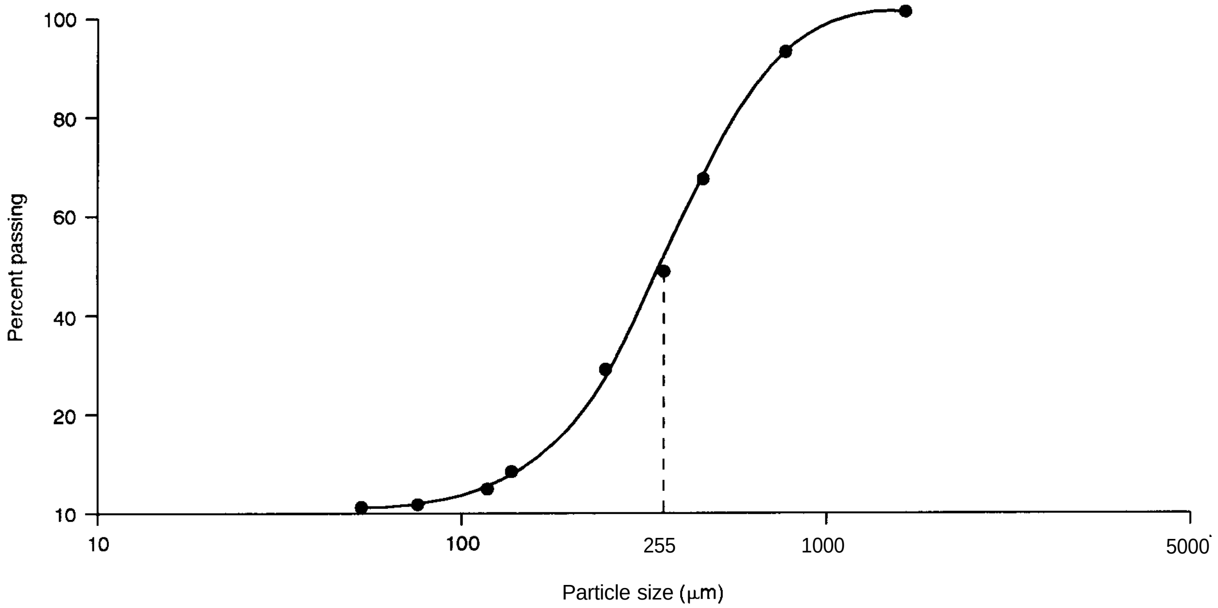

Figure 2. Particle size distribution of Forties sand

12.3.3 Thickness of laminar sub-layer

The thickness of the laminar sub-layer is simply related to the pipeline diameter and the Reynolds number for the case of a smooth pipe with a Reynolds number $<10^7$. The relationship is as follows:

$$\delta = 62\ D\ Re^{7/8}$$

where:

$Re = [1488\ D\ V_{sl}] / \mu$
δ = thickness of laminar sub-layer in ft
D = pipeline diamater in ft
V_{sl} = superficial liquid velocity of interest in ft/s
p = density in lb/ft^3
μ = dynamic viscosity in cP

12.3.4 Particle settling velocity

The particle settling velocity is the velocity at which particles will settle under gravity in the stagnant fluid. This velocity is primarily determined by the relative magnitude of the gravity and the viscous drag froces acting on the particle. Three settling laws are required to cover the possible range of settling conditions from low Reynolds numbers (ie small particle diameter/high viscosity fluid) to settling with high Reynolds numbers (large particle diameter/low viscosity fluid).

The three relationships for the settling velocity are as follows:

(a) Stoke's law

For Re, < 2

$$w_s = [1488\ gd^2\ (p_s\text{-}p_l)\ /\ 18\mu]$$

(b) The intermediate law

For 2 < Re, < 500

$$w_s = [3.54\ g^{0.71}\ d^{1.14}\ (p_s\text{-}p_l)^{0.71}]\ /\ [p_l^{0.29}\ \mu^{0.43}]$$

(c) Newton's law:

For Re, > 500

$$w_s = 1.74\ [gd\ (p_s\text{-}p_l)\ /\ p_l]^{0.5}$$

where:

the particle Reynolds number is given by:

$$Re, = (1488\ d\ w_s\ p_l)\ /\ \mu$$

Since the Reynolds number depends on the particle settling velocity the correct equation to use is found by calculating the settling velocity and Reynolds number by each equation and comparing the Reynolds number with the applicable limits for each method. For particles of between 50 and 1000 micron in oil the appropriate law is likely to be the Stoke's or intermediate laws.

The particle settling velocity can be used to estimate the flowing conditions required to transport solids in vertical pipes. For liquid/gas/solid flow it is required to consider in which phase the solid particles are transported.

12.3.5 Friction velocity at minimum transport condition

When the particle diameter is larger than the laminar sub-layer then the friction velocity at deposition for the limiting condition of infinite dilution is correlated by:

$$u_o^* = [0.204\ w_s\ (v/d)\ (\nu/D)^{-0.6}\ \{(p_s\text{-}p_l)\ /\ p_l\}^{-0.23}]^{0.714}$$

where:

w_s = particle settling velocity (ft/s)
u_o^* = friction velocity at minimum transport condition for infinite dilution(ft/s)
d = solids particle diameter (ft)
ν = kinematic viscosity (ft^2/s)

When the solids concentration is high the friction velocity is modified by the following relationship:

$$(u_c^*/u_o^*) = 1 + 2.8\,(w_s\,/u_o^*)^{0.33}\,.\,\Phi^{0.5}$$

Where u_c^* is the friction velocity at the minimum transport condition for a given concentration and Φ is the solids concentration volume fraction in ft^3/ft^3.

For most cases of interest in oil and gas pipelines the solids concentration is low and the first equation is usually sufficient to determine the friction velocity for minimum transport. However the concentration correction may be required if the liquid holdup is very small.

When the solids particle diameter is smaller than the laminar sublayer the expression for the friction velocity at the minimum transport condition is:

$$u^* = [100 w_s\,(\nu/d)^{2.71}]^{0.269}$$

Given the friction velocity the Reynolds number and the thickness of the laminar sub-layer can be calculated and the appropriate friction velocity expression checked.

12.3.5 Pressure gradient at minimum transport condition

Following the above procedure determines the friction velocity at the minimum transport condition for the liquid phase. This is easily used to calculate the associated single phase pressure gradient at this condition using the expression:

$$\Delta P_{mtc} = (4\,\rho_l\,u^{*2})\,/\,[144\,g_c\,D]$$

where:

$g_c = 32.174$

A two phase flow pressure drop method can now be used to determine the liquid and gas velocitiy combinations which result in the same two-phase flow pressure gradient. It is useful to plot the locus of these points on a flow pattern map to indicate the conditions under which solids may or may not be transported. Alternatively comparing the two-phase pressure drop for the conditions of interest with that for the minimum transport condition will indicate whether solids are deposited or not. Figure 12.3 shows a comparison of the model predictions with some of the data obtained at BHRA.

XFE have PC computer programs available to perform these calculations and to plot the results. For more information contact Paul Fairhurst on Sunbury 2983.

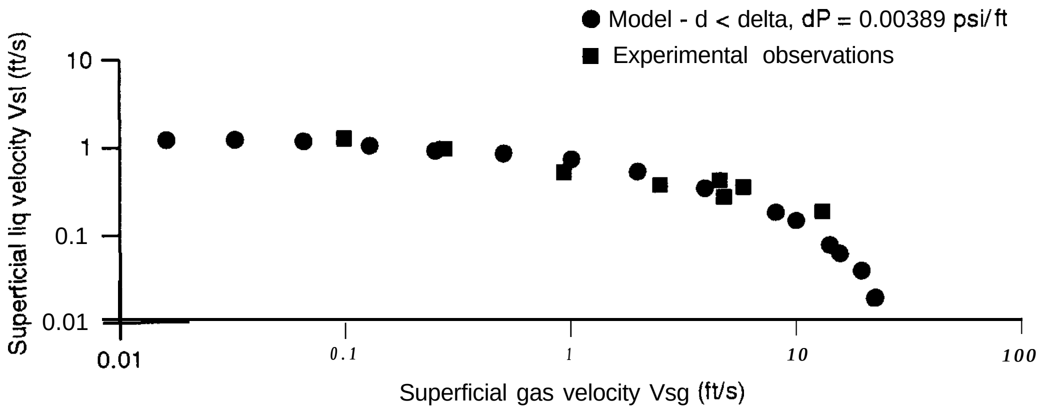

Chart 1 - BHRA Case A - 1" Pipe (air/water)

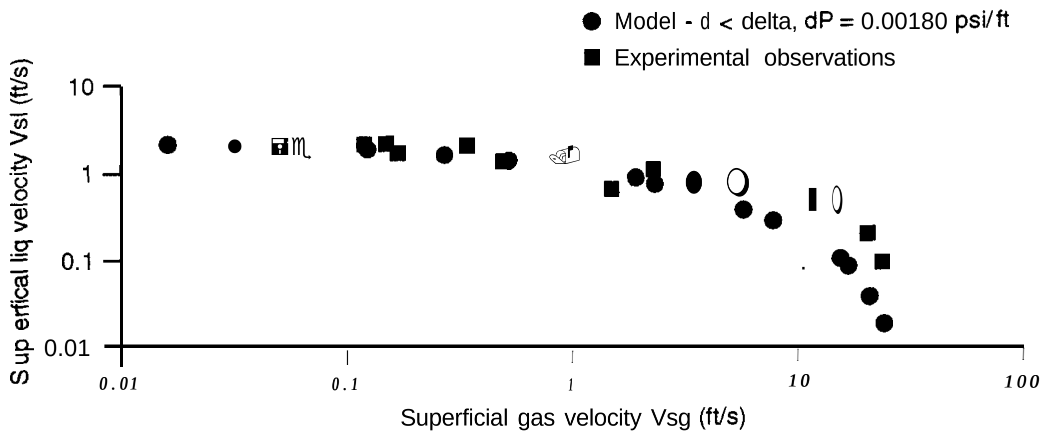

Chart 2 - BHRA Case B - 2" Pipe (air/water)

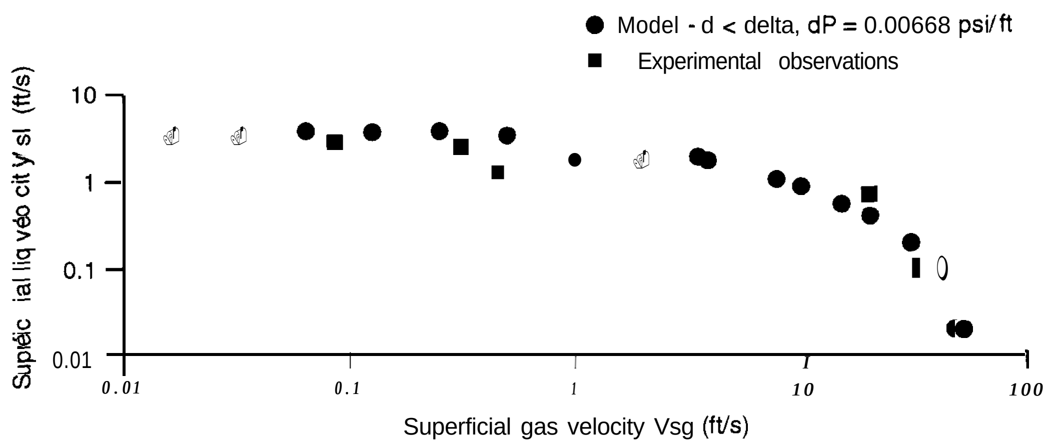

Chart 3 - BHRA Case C - 2" Pipe (water/glycerol mixture)

Figure 12.3 Comparrisons of the model with BHRA data

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References

(1) Fairhurst,CP "Sand Transport in the South East Forties Pipe Line", BHRA,1983.

(2) Smith,M "A Model for Predicting Solids Transport in Near Horizontal Multi-phase Oil and Gas Pipe Lines", XFE report 8/2/1993.

(3) Wasp, Kenny & Gandhi "Solid-Liquid Flow Slurry Pipe Line Transportation", Gulf Publishing Company, Clausthal, Germany 1979.

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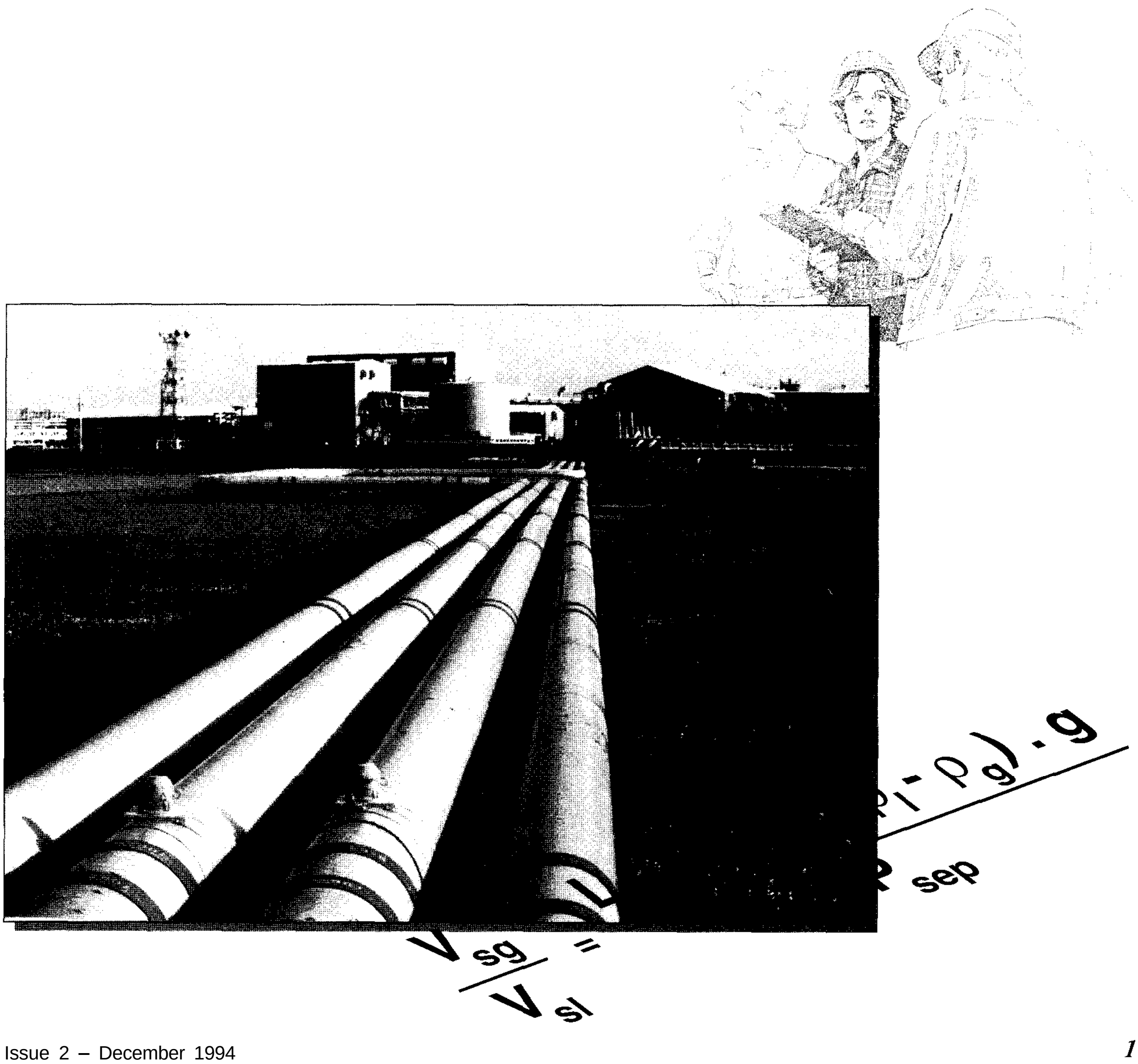
Section 13. Three **Phase** Flow

Effect of Water on Oil/Gas and Gas/Condensate Systems

Section 14. Flow Monitoring

Monitoring of Multiphase Flow



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14 Flow Monitoring

This section of the guidelines will first describe the requirements for instrumentation of multiphase flowlines. The context of this is of data acquisition above and beyond the normal measurements of pressures, temperatures, and single phase flowrates after gas/oil/water separation. In these descriptions of requirements the section numbers of appropriate measurement techniques are given for easy reference.

In the second half of this section brief descriptions of available techniques are given. Some are non-intrusive, and may be fitted temporarily to systems. Other measurement techniques will require a spoolpiece or tapping in the line.

For monitoring of phase flowrates in multiphase flow, please refer to Section 15 (Metering).

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14.1 Instrumentation Requirements

14.1.1 Flow Regime Identification

The original requirement for flow regime identification in oilfield multiphase flowlines arose from the R&D activities set up to develop and verify flow regime prediction techniques. The design of processing facilities is affected by flow regime, in that slug flow may cause unacceptable gas and liquid flowrate variations in the plant. Therefore reliable prediction of flow regime over a wide range of gas and liquid flowrates, pressures, temperatures, fluid compositions, and line topographies are very important. See Sections 1 and 3 of this manual for further discussion of this issue.

A series of field trials has been conducted within BPX to collect data on flow regime. This activity will continue on major new oilfield developments to ensure that the prediction methods remain applicable in new combinations of conditions.

Please refer to Sections 14.2.1 – 14.2.4 and 14.2.6 – 14.2.7 (for information on the different measurement techniques available for this activity.

14.1.2 Line Holdup Estimation

The next step beyond identification of flow regime is to be able to determine the hold-up of liquid in the pipeline. This is required in the assessment of pigged liquid volumes, and in the prediction of slug volumes generated by flowrate or topography changes. Some pressure loss prediction methods also rely on predictions of liquid hold-up.

As with Flow Regime Identification this activity is largely R&D based, to verify hold-up prediction methods for future design requirements, rather than for a current or future normal oilfield operation. For this activity the techniques described in Sections 14.2.3 – 14.2.7 may be appropriate.

14.1.3 Slug Flow Characterisation

This activity has two main thrusts - firstly the provision of data and subsequent verification of slug flow design methods, and secondly the provision of information for feed-forward control of slug reception facilities (slugcatchers, separators).

(a) Slug Flow Design Methods

The instrumentation requirements are threefold – to measure the frequency of slugging, the velocity of slugs, and the phase volumes (gas, oil and water) contained within the slugs and subsequently delivered to the process. Data acquired is used to validate the predictive tools for these parameters.

For the development of slug flow design techniques it has been important to gather data over a wide variety of conditions of gas and liquid flowrate, line diameter and topography, and fluid properties. Continuing data acquisition on lines with new combinations of conditions will be carried out as appropriate. See Section 3 of this manual for an in-depth discussion of the development of the actual design methods. The current thrust is in the development and validation of methods for hilly terrain systems.

A pair of measurement positions is required for the determination of slug velocity. Techniques suitable for slug flow characterisation are described in Sections 14.2.1-14.2.4 and 14.2.6-14.2.7. The techniques in Sections 14.2.1-14.2.2 do not give the phase fractions within the slug.

(b) Feed-Forward Slug Flow Control

The requirements for this application are twofold - firstly to measure the velocity of slugs, and secondly to determine phase volumes contained within the slugs. This allows determination of the arrival time of slugs into the reception facilities, and gives the liquid inflow rate value to an assessment of the likely effect of the slug on oil and water levels, and of the effect of any gas surge following the slug on the gas plant.

Appropriate instrumentation techniques are the same as those listed in the Slug Flow Characterisation section.

Location of the measurement position has to be carefully selected in order to give sufficient time for any feed forward activity (e.g. increasing pump-out rates) to take place. The biggest benefit will be obtained from such feed forward control in systems where the slug frequency is quite low, where a slug arrival will cause a significant change in separator vessel conditions. This will be the case in larger diameter lines, and lines operating close to the slug-stratified boundary.

14.1.4 Reservoir Management

A knowledge of the volumes of fluids produced from different wells in a field is generally required in order to be able to manage depletion of the reservoir. Water breakthrough needs to be identified, as does gas coning. The current industry standard is to flow wells one at a time to a test separator via a dedicated test line. The flowrates of gas, oil and water are measured after separation. Wells are tested typically once a month. Required accuracy is of the order of +/- 5 to 10% on each phase.

In an increasing number of satellite tiebacks to existing installations, the requirement to have a local test separator, or a second pipeline back to a test separator at the host facilities, can add an appreciable cost to the basic development. The cost of metering for reservoir management could be reduced by using a multiphase metering system that is more compact, and cheaper, than the test separator system. Details of possible systems are contained in Section 15, Multiphase Meter Systems.

14.1.5 Production Allocation

In some developments with multiple fields the cost of installation of separate processing trains for each field may be prohibitively expensive. Subject to the agreement of all partners and regulatory authorities, it may be acceptable to allocate production back to the individual fields (to determine royalties and taxes) using methods that do not require full processing trains with fiscal metering.

In some cases a remote test separator will be used (e.g. Point McIntyre to Niakuk), with adjustments made to ensure that the sum of well allocations from all contributing fields equals the fiscally metered export rates from the central processing facility.

In other cases a further cost saving may be obtained by using multiphase metering systems either remotely, or just upstream of where the production from one field is co-mingled with that from other fields (e.g. under consideration for ETAP). The multiphase metering methods currently available for this activity are described in Section 15, Multiphase Meter Systems.

Current regulatory guidelines centre on the need for each contributing field to have the same metering set-up, so that any systematic errors tend to offset for the different fields.

14.1.6 Fiscal Measurement

A further, even more difficult requirement, is to measure the phase flowrates in a pipeline sufficiently accurately to allow for custody transfer to the same level of accuracy as current single phase standards.

As yet no multiphase measurement technique or product is sufficiently accurate to achieve this to single phase flow standards. Such accuracy requirements are of the order of better than +/- 1%.

In parallel to developments aimed at improving the accuracy of multiphase metering systems will be negotiations as to what is an acceptable accuracy for partners and governments associated with an oilfield development. These two aspects (technology advance, and relaxation of accuracy) will probably both have to progress in order for 'fiscal' metering of multiphase flow to become a reality. Some precedent is already found on the North Slope of Alaska, where small fields have their production allocated on the basis of well test results (as described in Section 14.25). In this case there is no bulk flow measurement upstream of the point of co-mingling with fluids from another field with different (slightly) ownership.

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14.2 Techniques Available

14.2.1 Pressure

The analysis of a pressure measurement made by a relatively fast pressure transducer may yield some information. Differential pressure across two transducers has even more value.

It is possible to obtain an indication of flow regime from a pressure measurement. Simple observation shows the passage of slugs past a measurement location as a significant and potentially prolonged rise in pressure. Pattern recognition can be used to analyse the signal further to differentiate between stratified smooth, stratified wavy and annular flow. Such an analysis may be supplemented by flowrate or pressure loss information, which will significantly reduce the uncertainty in the deduction. However, it is not possible to get information on liquid content in a cross-section from pressure only.

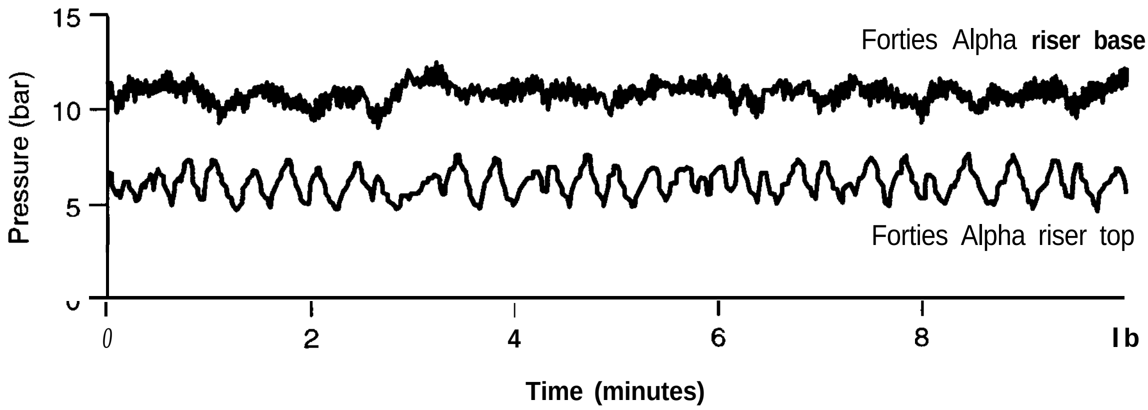


Figure 1. Pressure fluctuations in Forties Echo-Alpha flowline.

Some extension of the use of pressure measurement for flow metering has been proposed. Firstly, the variation of pressure drop across a choke can give an indication of flowrate changes. This information can be used effectively during the periods between well tests to give approximate flowrates for choke settings different from those under which well tests were conducted. Pressure loss down a long length of flowline also gives an indication of flowrate.

Secondly, some development of neural network systems to analyse data and 'train' a computer to interpret pressure signals as a flowrate measurement has gone through preliminary stages. The ability to make calibration (training) runs is required here.

14.2.2 Acoustic Monitoring

There are two basics approaches to acoustic monitoring - passive and active. In a passive system the measurement device listens to the noise generated by the flow. An active system puts sound energy into the fluid - the sound is then picked up at a receiver after passage into or through the fluid.

(a) Passive Systems

One such system has been developed within BP Exploration. Stemming from the requirements to have more easily portable and non-radiation based techniques, the idea of listening to the flow was pursued. The result is a hardware and software package that has been used successfully for flow regime identification and slug flow characterisation in a number of fields (Wytch Farm, Kuparuk, Prudhoe Bay...).

Transducers pick up flow generated noise around the 70 kHz frequency. After filtering the signals can be analysed to give wave and slug frequencies and velocities, and slug lengths.

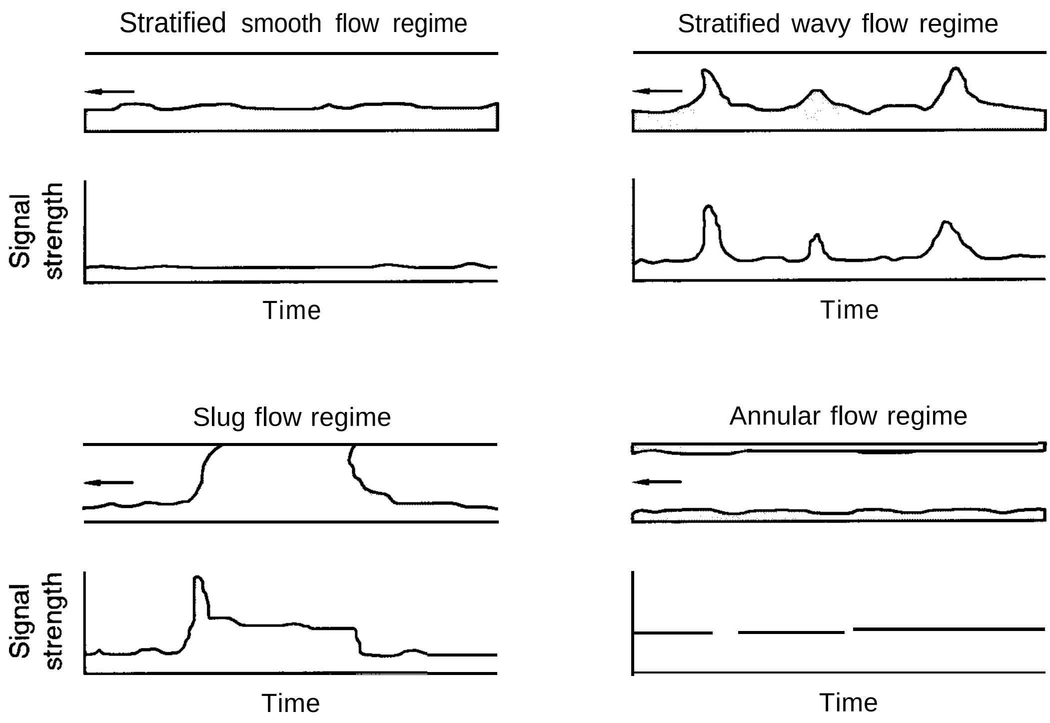


Figure 2. Title ?

(b) Active Systems

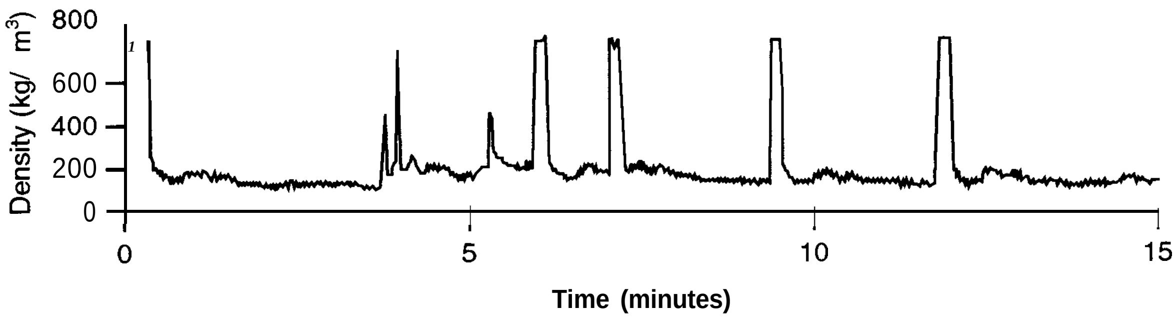
These have been used for two different aims - flowrate measurement, and gas fraction determination. Doppler frequency shift techniques are used for single phase liquid flowrate measurement. The principle has been tried in multiphase flow but with little success, because the gas/liquid interfaces break up the acoustic signal.

The fact that this break-up occurs has been used in a device to measure gas fraction near the pipe wall. By directing an acoustic signal into the flow, and measuring how much of the signal is scattered back, an estimate of gas fraction can be made (after calibration).

14.2.3 Gamma Ray Densitometry

This is the most widely used instrumentation technique for studying multiphase flow characteristics. The principle is to direct a beam of gamma rays from a sealed source (usually Caesium 137, although some dual energy systems have also been tried e.g. Caesium and Americium) across a pipe and to measure the intensity of radiation transmitted through the pipe wall, any insulation, and the fluids within the pipe. The absorption by the pipe wall and insulation can be calibrated out to give as output a measurement of the density of the fluids within the pipe. The technique will give a plot of density vs. time at the measurement location. From a trace taken from a gauge mounted with the gamma beam vertically through the pipe, the flow regime can be identified as slug, bubble, stratified smooth or wavy/annular. To resolve between annular and stratified wavy the gauge is then turned through 90° to transmit through the pipe horizontally. The existence or otherwise of a liquid film at the mid pipe location will define whether the flow is wavy or annular.

Setting two gauges on the pipe separated by a distance allows cross-correlation of the signals to determine the velocity of the slugs and waves between the measurement locations. Coupled with the measured time taken for a slug to pass a gauge, this will yield a slug length.



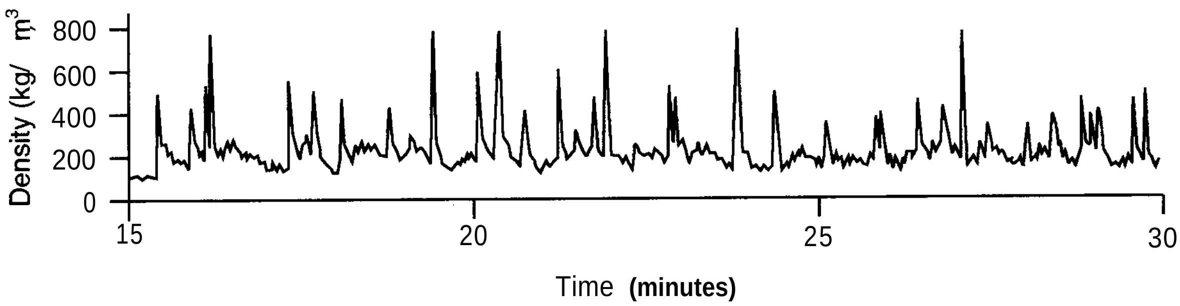


Figure 3. O-l 5 minutes period and slug flow, 15-30 minutes period and wavy flow.

14.2.4 X-ray

The expense and size of X-ray systems has, to-date, prevented their widespread use in the oilfield environment. However, there have been isolated tests in which an X-ray system has been used to identify flow pattern. Real-time displays are possible of the fluids in a pipe.

The principle of operation is the same as the gamma ray densitometer, in that the fluids in the pipe (gas, oil and water) absorb X-rays to different degrees. By directing an X-ray beam across the pipe, and mounting a detector on the opposite side of the pipe to the source, it is possible to 'see' the flow behaviour in the pipe.

14.2.5 Neutron Back Scatter

This technique depends on the concentration of hydrogen in the vicinity of the probe. A neutron generating source is used to send neutrons into the process pipe/vessel. The intensity of emitted gamma rays from the interaction between neutron and proton in the hydrogen gives the hydrogen concentration, which in turn indicates the presence of gas or liquid. This technique can give an indication of flow regime, and also of liquid film depth in a pipe.

14.2.6 Capacitance/Gamma

The principle behind this technique is the determination of dielectric constant between electrodes by measuring capacitance, coupled with a gamma ray transmission measurement to determine density. The measurement of capacitance requires great care in shielding and design of the electronics. The make-up of the overall dielectric constant is determined by modelling the contributions of gas, oil and water in the three phase flow.

The output should give a fast response indication of the gas-oil-water phase fractions in the cross-section of a pipe. This will allow flow regime determination, and slug flow characterisation if two devices are used with a suitable distance between measurement sites.

The systems are designed as full-bore in-line spools with ceramic liners. Further details may be seen in section 15, Multiphase Meter Systems.

14.2.7 Microwave Absorption

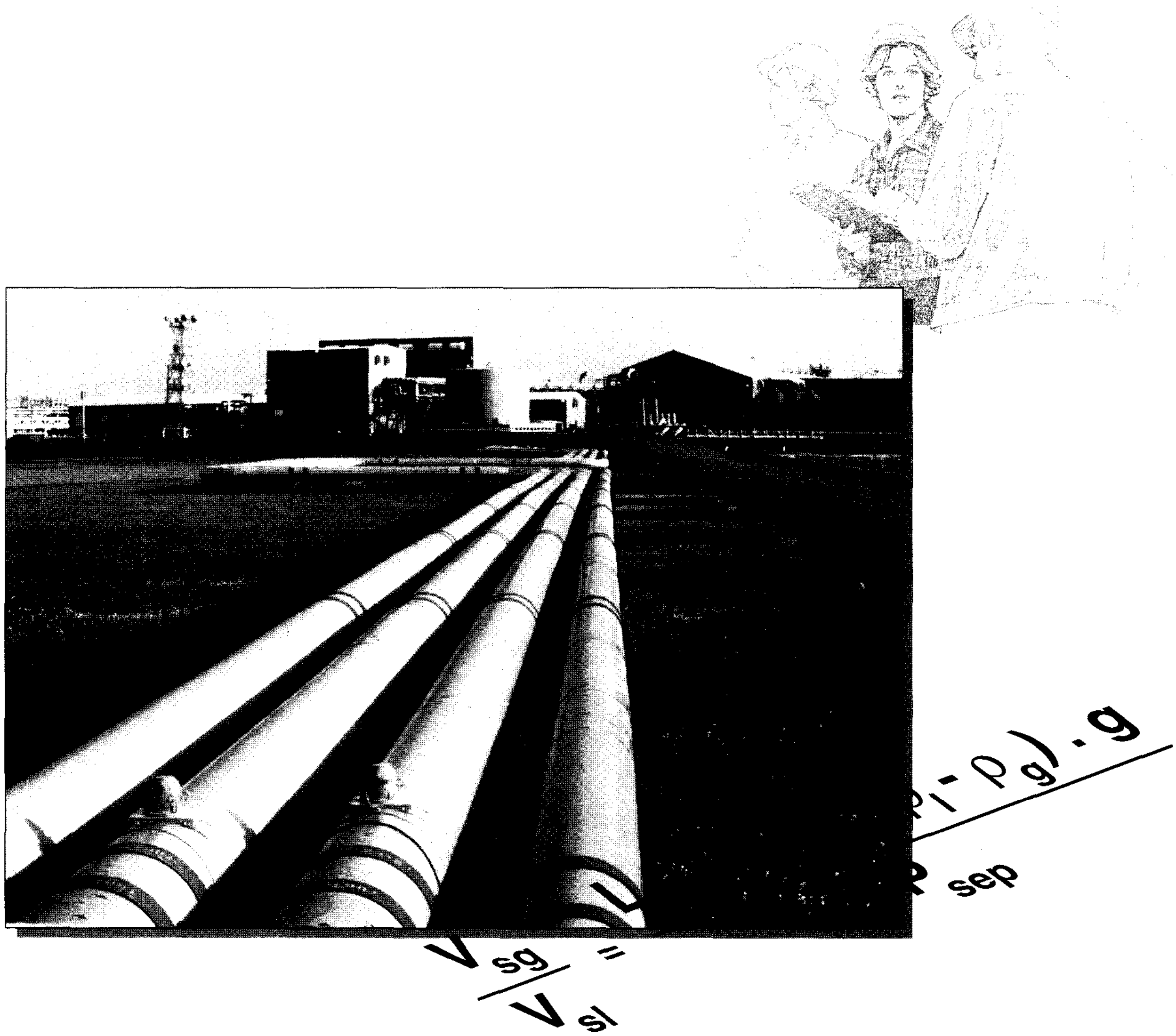
The principle behind this technique is also the determination of dielectric constant between electrodes, this time by measuring microwave absorption, also coupled with a gamma ray transmission measurement to determine density. The measurement of microwave absorption requires great care in design of the intrusive probes. To resolve the individual phase fractions, the measured dielectric constant is compared with the value predicted from modelling the contributions of gas, oil and water in the three phase flow. The calculation is iterated around the phase fractions until the calculation matches the measured data.

The output should give a fast response indication of the gas-oil-water phase fractions in the cross-section of a pipe. This will allow flow regime determination, and also of slug flow characteristics if two devices are used with a suitable distance between measurement sites.

The systems are designed with intrusive probes. Further details may be seen in Section 15, Multiphase Meter Systems.

Section 15. Metering

Multiphase Meter Systems and their Application



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15.1 Introduction

In general, oil industry effort to develop multiphase flow metering techniques has been in progress since the mid 1980s with one or two exceptions, such as various approaches by Texaco, which commenced a few years earlier. As a result of this industry-wide effort, key physical measurement techniques investigated are now emerging as commercially available field test hardware.

By comparison, relatively little attention has been paid to the problem of calibrating multiphase flow meters. Some of the development projects have resulted in the construction of oil/water/gas flow loop facilities for test purposes but little thought seems to have been given to the problem of calibration of multiphase flow meters for field service.

This section of the Multiphase Design manual presents a review of multiphase metering systems in which BP is involved or the author has technical knowledge. This encompasses what are believed to be the most advanced and soundly based technologies. The review is necessary in order to give a background perspective to the preceding statements concerning calibration. It is also intended that it provide the state of the art of emerging technologies. The review includes information on current capabilities and costs as well as describing the different measurement concepts.

The following section of the manual, Section 16, concerns calibration philosophy of multiphase flow meters. The purpose of this is to highlight the key aspects and challenges faced in establishing a calibration methodology for multiphase flow meters. Thoughts on the problem are presented, particularly, with reference to calibration for single phase flow. It is intended that the contents act as stimulus to trigger input to the problem in order to facilitate the development of practical calibration solutions.

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15.2 Multiphase Meter Development Review

This review primarily covers technology to which BP has contributed directly and in which the company has had involvement. Comment is also included concerning other techniques of which the author has knowledge and which have reached commercialisation and field test stage. It should be realised that the technology is currently evolving with new techniques, experience and test results continually emerging.

Standard terms are used in quoting accuracy or error percentages. Specifically, relative error means error or deviation expressed as a percentage of actual flow rate. Absolute error refers to an error or deviation expressed as a percentage of full scale.

All of the metering approaches described below are targeted at accuracies for well test duty for reservoir management and the possibility of allocation of production between wells. The high accuracy levels associated with custody transfer and the term "fiscal metering" in single phase flow metering have not been addressed. On the basis of research and development experience to date, this level of accuracy is not likely to be realised by multiphase metering techniques in the foreseeable future.

15.2.1 BP-Patented Positive Displacement Screw Meter

Being licensed to ISA Controls Ltd. in 1994, this flow meter is believed to be the most advanced in terms of flow loop and field test experience. Two prototypes built to ANSI class 1500 pressure rating and to NACE specifications in 1987 have been used for the test programme. A full range of flow conditions (of up to 95% gas void fraction (GVF) during trials at Prudhoe Bay) has been covered and several thousand run-hours experience gained. The meter measures total volumetric and total mass throughput of the multiphase fluid stream and uses a water cut reading by some other means to measure the flow rates of oil water and gas.

Cost to supply a BP flow meter can only be estimated at present but is expected to be of order $150,000 for worst case service specifications. Cost of a complete ISA multiphase measurement skid will be application dependent (i.e. land or subsea packaging, number of meters to give turn-down, additional instrumentation required - see below). The two field prototypes were designed for a nominal pipe size of 3" or 4" and are approximately one metre in length and three quarters of a tonne in weight. The opportunity to build the first commercial meter will permit the incorporation of many design improvements realised through the course of the prototype testing since 1987. This could have significant implications for manufacture, ease of maintenance, size and cost.

Some limited study has been conducted addressing the costs of packaging and installing a multiphase metering system for subsea duty using the BP positive displacement meter. The costings are highly dependent on the application but demonstrate how the ex-works cost of the multiphase metering system is not necessarily the only significant contributor to the overall costs associated with its implementation. The findings of these studies are available in two BP reports (1,2). Further development is required in order to package the sensors and signal processing and data transmission for remote service sub-sea, but this is not expected to be a problem.

The BP meter requires an additional device to supply a reading of water cut in order to yield the flow rates of oil, water and gas. Currently, any one of three to four composition measurement techniques under development could give the water cut measurement. The test work to date indicates that these techniques can measure water cut by dynamically tracking the phase fractions in the pipe cross-section.

The test work has shown that the total volume positive displacement measurement can achieve the target of +/- 5% of reading (relative error). Currently, the corresponding turn-down varies from a minimum of 5:1 to 15:1, dependent on flow regime behaviour. The K-factor curve to enable this in multiphase flow can be determined on the basis of a calibration in a single phase water flow loop.

The flow driven elements of the screw meter generate a pressure drop which depends on flow rate, gas volume fraction and flow regime. For high liquid loadings, pressure drops of up to 2.2 bar have been observed during tests when the meter has been run at the top of its range.

The meter incorporates a commercially available gamma densitometer to measure mixture density. This may be factory calibrated on fluids of known density (such as air and tap water). A densitometer of this type has been in use subsea, in order to monitor multiphase flow behaviour at the foot of the Forties Echo-Alpha 12" riser, since May 1988. This densitometer has required no subsequent intervention and is still functioning.

The principle behind the screw meter concept is that the positive displacement elements constrain the phases to move at a single velocity. The densitometer beam passes through a cavity between the screws mid-way along the meter axis. In principle, if slip between the phases were fully eliminated between the screws, the densitometer reading would equate to the true mixture density relating to the overall proportions of oil, water and gas flowing in the pipeline. Coupled with the water cut reading from a second instrument, the mixture density and total swept volume readings can then be used to solve out the flow rates of oil, water and gas at line conditions. (This assumes that the individual phase densities are known as a function of pressure and temperature which are also measured at the multiphase metering section).

All the test data to date have shown a tendency for the mixture density (or total mass flow measurement of the stream) to exhibit a bias error which is a function of gas void fraction, GVF. Above 40% GVF, this bias depends on the installation site i.e. it is specific to the site. Currently, in order to achieve relative errors in actual flow rate of each phase of 5% to 10 %, this bias in mixture density or total mass flow rate must be calibrated out by reference to a measurement of total mass throughput at the site installation. This is not a regular calibration requirement but would be necessary initially and subsequently following any change of operations which significantly influenced the nature (GVF) of the flow stream. Future development and design improvements are expected to reduce, and could eliminate, this requirement for a site calibration on total mass. (Note: 7/6/94: In 1993 the bias effect was eliminated in testing in Alaska).

Test data to date indicate that relative error uncertainty increases with GVF. With the bias correction, the scatter of relative error in phase flow rate increases from 5% to 10% at a maximum GVF of 95%. If the mixture density or mass bias error is not site calibrated then the meter exhibits a general uncertainty in total mass flow rate increasing with GVF to +/- 15% at 80% GVF. Without a total mass bias correction, the phase flow rate measurement errors can become large at high gas void fractions. The meter has not been field tested at the highest GVFs above 95%. Further experience would be required to confirm the GVF range of the meter at the highest levels. (Note 7/6/94: In 1993 testing at Prudhoe Bay with the bias effect eliminated, relative errors in phase flow rates were scattered ± 5% . This was observed from many tests covering several wells and choke settings giving a GVF range of 75% to 95%).

The above uncertainty levels relate to relative errors in total flow rate and individual phase flow rates. They have been determined on the basis of the scatter of measurement errors observed from all the test data (different sites and wells). It is possible that site specific calibration on total mass flow rate on an individual well basis could exhibit a superior repeatability, thus reducing the uncertainty from the above levels.

The bias error in mixture density is believed (on the basis of analysis of all the test data) to be attributable to flow pattern regime effects between the screws which do not, therefore, fully homogenise the flow and eliminate slip in the idealised manner. Further design modifications could be investigated to address this which might eliminate the current need for a site specific calibration. (See above).

The BP meter based system, including any of the separate devices being developed which will measure water cut, essentially, can be factory calibrated using single phase fluids in relatively straightforward factory test cells.

Summary remarks:

The simple rugged mechanical design of the BP meter took full account of the need for many months service without maintenance in a flow stream exposing the meter to violent mechanical shocks from slugging, abrasives and corrosive substances. The test programme so far has confirmed the design under such conditions. Experience will determine wear characteristics and the extent to which these influence the need for re-calibration, if any.

A key reason for selecting the positive displacement design and adapting it to the no-slip principle was that this would result in a meter with relative insensitivity to the generally non-reproducible flow patterns of multiphase streams. It therefore follows that the principle should lend itself to calibration.

The current design requires an initial factory calibration. The question arises- is it desirable to prove the meter once in service? Is this even a practical proposition for a multiphase flow meter? What methods of proving a multiphase meter could be used: - portable test separator system; another multiphase meter; testing by difference using the production separator measurements or where the sum of individual well flows is balanced?

152.2 Framo Multiphase Metering System (JIP - BP a partner)

As for the BP multiphase meter, this system is based on a technique which should in principle be insensitive to the non-reproducible nature of multiphase flow.

It comprises a novel mixer or flow conditioner which, unlike conventional mixers, can be designed to mix slugs and gas pockets and produce a steady exit stream from intermittent regimes (It, in effect, can mix phases segregated axially along the pipe). The flow conditioner vessel should, in practice, be compact relative to a separator and its functioning is entirely passive (no moving parts, valves or level controls).

The system is under development by Framo Engineering AS in Norway. It is at an earlier stage than the BP meter and a prototype is only just approaching a first field trial. Sensors used in the conditioned stream immediately downstream of the mixer to measure the flow rates of oil, water and gas are described below. Framo have conducted oil, water, gas flow loop tests at their Fusa Fjord dock-yard.

Total flow

A venturi is being tested. Data to date indicate questionable accuracy. It is currently unclear to what extent this is attributable to the challenge of measuring pressure differential accurately over a wide turn-down (the measurement data are particularly erroneous at high gas fraction). It may also be that, in spite of the mixing of the flow stream, localised phase slip effects still persist through the measurement section.

Phase fractions

Dual energy gamma ray attenuation measurements are used to determine the ratios of oil, water and gas. As for the total flow measurement, accuracy is currently subject to an unacceptable degree of uncertainty. Again, the extent to which phase slip is responsible is unknown. Certainly, the current basic dual energy gamma ray detection system is subject to fundamental limitations. However, new technology and steps within the JIP to improve the system could yet move the system forward.

The most recent test results show relative measurement errors in phase flow rates and total flow of 20% to 30%.

The mixing unit could, of course, ultimately be used with other measurement devices. Dielectric composition measurement techniques being developed elsewhere might do better than the dual energy gamma system. It is possible that the mixer could assist cross-correlation velocity methods to achieve a degree of consistent and reproducible measurement.

The current JIP is developing this multiphase metering package as an offshore module. Framo Engineering's expertise lies principally in this area. The cost to supply a Framo mixer based system, either as an offshore package or in terms of the off-shelf elements of the multiphase metering system, is unknown. The field prototype, designed for a nominal 3" or 4" line, is in a barrel approximately 4 m long by 0.5 m diameter. Its component parts, engineering and manufacture resulted in a total cost of approximately $900,000. Total weight is 2.5 tonne. This unit is built for top-sides use only, but some of the key sub-sea design of component parts (such as subsea housing of sensors and electronics processing and data transmission, barrier fluid system, electric coupler etc.) has been accommodated in the prototype. It is rated for class 2500 pressure service.

The complete prototype system has exhibited pressure drops reaching 2.6 bar during the flow loop test programme. This was realised at flow rates comparable to the top of the range of the prototype BP multiphase flow meter.

Calibration of the Framo system would rest with the type of measurement used. For the venturi meter, it is primarily a matter of subjecting the pressure sensors to accurately known fluid pressure. This does assume that a universal or repeatable and predictable venturi discharge coefficient can be established for well mixed multiphase flow. Flow loop progress on this so far has not been good.

The dual energy gamma system requires calibration on fluids of known density and attenuation coefficient. Once installed, some checking should be possible using a barrier fluid (pumped via umbilical) system which Framo have incorporated into the test prototype design for various functions such as sealing and pressure leg flushing. However, variations in phase properties, such as water salinity or attenuation coefficients of the phases with any change in phase composition, would result in the need for some means of re-calibration in-situ. This could result in a need to take fluid samples unless some means of separating and isolating individual phases in the measurement section can be developed*.

** Such a technique* has *been demonstrated by the author for measuring individual phase density using the BP prototype multiphase flow meter at Wytch Farm and during the later trial at Prudhoe. Bay in 1992. It relies on the installation of the multiphase meter in a by-pass and block valve arrangement. The procedures would be part of the* overall *well testing procedure.*

The question then remains as to how calibration will change with any service degradation of, for example, the critical venturi throat and gamma beam path cross-stream dimensions. Here, as for the BP meter, the desirability for regular re-calibration, possibly in-line using some means of proving, must be addressed.

A further question arises as to whether the Framo mixer can be designed in any application to accommodate the full range of flow regimes, flow rates and phase fractions expected over the life of the well pad. (The same question applies to the venturi). Framo have developed a design basis for the mixer. For example, it can be sized and dimensioned to suit a certain range of slug and bubble lengths. Framo claim, but have yet to prove, that it can be designed to cover the full range of regimes including annular. Whilst this may in principle be correct, it remains to be seen in practice whether universality can be achieved without excessive bulk and multistage configuration of the mixer.

15.2.3 The CMR (Christian Michelsen Research)/Fluenta Capacitance Cell (BP sponsorship 1985-1992)

This full-bore non-intrusive phase fraction sensor has been developed with significant assistance from BP alongside the BP multiphase flow meter. It measures in-situ phase fractions within a pipe cross-section where the phases are uniformly distributed. Absolute errors of +/-2% phase fraction have been demonstrated by the test programme for gas void fractions (GVFs) of up to 50%.

The existing cell will operate on oil-continuous liquid phase only. Water continuous liquid (or a water film around the sensor lining wall) results in a short circuit and meaningless reading. CMR (formerly CMI) are developing an additional sensor for the system which uses an inductance principle to measure phase fractions with water-continuous emulsions. Both types of sensor rely on a simultaneous gamma ray attenuation measurement of cross-section mixture density in order to resolve the three phase fractions.

The water cut may be derived from the phase fraction measurements. A class 1500 version of this device was tested downstream of the BP multiphase meter at Prudhoe Bay in 1992. These tests indicated that, by dynamically tracking the varying cross-sectional phase fractions in the multiphase stream, the cell could give accurate readings of water cut (within 1% absolute) for GVFs close to 90%. The limitation on achieving this in all cases was imposed by annular flowing well streams and flows where slugs of significant duration and liquid hold-up (instantaneous GVF well below the stream overall average level) were not in evidence.

Calibration of the phase fraction cell for the field trial was completed at the CMR laboratory. Samples of dead crude and formation water were shipped from Prudhoe Bay to enable calibration in a special purpose laboratory dielectric measurement cell. This is used to pressurise the sample oil and force into solution methane gas under varying pressures. As well as requiring as input constants the dielectric properties of the individual production fluid phases, the CMR/Fluenta system also relies on knowledge of the densities of the phases as a function of line pressure and temperature.

CMR and Fluenta have adapted the capacitance sensing section into a multi-electrode axial array in order to yield velocity measurements by cross-correlation processing of the electrode signals. The continuing project is now attempting the measurement of dual velocities by using combinations of different length electrodes in order to make some allowance for slip between the gas and liquid phases. All the drawbacks highlighted in the earlier comments, concerning the uncertainties associated with cross-correlation flow velocity measurement, apply. An accuracy figure for phase flow rates measured using the combined phase fraction and cross-correlation measurements cannot, at present, be quoted with any meaning.

The 1992 test programme, including the BP field trial and quite separate test programmes by two other operators, exposed a hitherto not experienced fundamental flaw with the use of the ceramic insulating liner material. (Development of a ceramic liner was pushed hard by CMR during earlier development of the cell and was a significant part of it). It has been discovered that the ceramic has a surface affinity to water under certain flowing conditions which results in a completely erroneous sensor response signal.

Fluenta are investigating alternative liner materials. Further extended field testing will be required to establish the suitability of any new candidate insulating materials.

Typically, Fluenta have quoted on the order of $300,000 for one of their multiphase metering systems in the past, though recent remarks from CMR suggest that this quotation will change in view of the lower price being quoted for a competitor's system (Kongsberg, see below). The sensor is a compact spool, approximately 0.5 m in length and weighing one tonne in the case of the cell tested at Prudhoe Bay. In view of the recent CMR remarks, $300,000 clearly does not relate to the basic manufacturing cost.

The CMR/Fluenta test and development programme continues in strength in spite of BP's withdrawal from the JIP in 1993. It has support from several oil companies in three separate test projects.

15.2.4 The MFI (Multi Fluid Inc.) Microwave System (JIP - BP a partner)

A competitor of the CMR/Fluenta system, this microwave instrument also measures mixture dielectric constant and gamma ray attenuation (mixture density) to resolve the in-situ cross-sectional phase fractions of oil, water and gas.

In one configuration, the instrument's microwave transmission and sensing section (which measures the phase shift of the microwave energy caused by the multiphase mixture, related to the phase proportions) functions in water continuous as well as oil continuous emulsions. This version of the MFI instrument, known as the "full range meter" is still undergoing laboratory level development of a field prototype. A first field trial is expected to take place during summer 1993 at the Elf operated Pecorade site in France.

A lower price version (so called LP version, not to be confused with low pressure) has lent itself to faster track development and has already been offshore on Statoil's Gullfaks B platform in late 1992. This configuration of the microwave sensing spool permits measurement with oil continuous emulsions only, but is more sensitive (and expected, therefore, to be more accurate) than the full range variant.

The 'low price' oil-continuous instrument is currently quoted at an estimated price of $200,000 by MFI. This includes a twin sensor arrangement which permits a single velocity measurement by cross-correlation. A price for the more expensive full range device has not yet been quoted. The sensors are intrusive, using antennae probes which protrude into the flow. The LP sensor comprises a resonant cavity, bounded by a grid of metal strips across the flow stream at each end of the sensing spool section. The prototypes of this version have been constructed to class 900 pressure rating. Further development is required in order to ruggedise the design of the probes of the full range meter for multiphase flow stream service.

The sensor spools of the prototypes built to date are 3" bore. The two parts of the meter incorporating cross-correlation measurement of velocity result in a measurement spool approximately 1 m in length. One flow composition measurement spool accounts for half this length. A further spool housing pressure and temperature transmitters is necessary (this also applies to the Fluenta/CMR system).

The Gullfaks B trials of the LP meter have resulted in only a few tens of hours exposure to production fluids so far. The phase fraction data agreed well with test separator measurements for the limited number of test cases. The GVF at the Gullfaks B test site is below 40% and the evidence suggests that the multiphase stream is well mixed and steady in nature.

Thus, these tests represent the easy end of the multiphase spectrum. Phase fraction results were within the +/- 2% absolute errors seen in laboratory flow loop tests. The few cross-correlation measurement data, in terms of both mixture velocity and phase flow rates, were more erratic. Relative errors of +/- 10% to 20% were reported.

The steady and relatively uniform nature of the low GVF multiphase stream of these field trials should have provided a flow structure and velocity profile of minimal complexity and least removed from that of the flow loop testing by MFI. This should have favoured the cross-correlation technique. However, such a flow stream also exhibits relatively weak perturbations in mixture dielectric properties on which to cross-correlate. This would tend to give a broad and relatively ill-defined correlation peak from which to deduce time of flight (to deduce velocity).

As for the CMR/Fluenta cell, calibration of the MFI phase fraction system, in principle, should rely primarily on factory or laboratory set-up prior to installation. Both systems use commercially available Cs137 gamma ray densitometers, as used for the BP multiphase meter, which can be calibrated on fluids such as air and tap water. The microwave unit requires data including density of the dead crude, produced water density and conductivity. The MFI system then relies on application specific prediction equations of the variation of these quantities and gas density as a function of line pressure and temperature.
MFI can configure their flow composition software to derive the flow stream as a split of total mass proportion of hydrocarbons and water, as opposed to volume fractions of oil, water and gas. These output readings are far less sensitive to uncertainties in the individual phase densities than the outputs of the phase fractions.

This split of output data would, of course, need to be related to total flow throughput to derive the total hydrocarbon and water mass flow rates. This could only be achieved if it could always be ensured that the phases are well mixed and flowing all at one velocity through the measurement spool (zero slip).

The question needs to be answered as to whether output of total hydrocarbon mass and water throughput is a more useful operating measure than in-situ volume flow rates of oil, water and gas. The latter usually require the application of correlations to account for shrinkage and solution gas factors in order to convert to flow rate units expressed at standard pressure and temperature.

Like the phase fraction cell developed by CMR, the fraction part of the MFI microwave technique could be used to supply a water cut reading to the BP multiphase meter system. It would be used to dynamically track the phase fractions and should yield water cut in high GVF flows as well as covering the lower part of the GVF range. Such an off-shelf option should presumably cost less than the estimate given above for the MFI LP flow meter, which includes an extra sensor and software to allow cross-correlation measurement of velocity.

It should be noted that MFI have already successfully commercialised a water cut sensor using the same microwave hardware configuration as the LP meter. This instrument is for gas-free emulsions and should not be confused with the multiphase flow meter technology described above, which incorporates a gamma densitometer and additional sophistication in order to measure oil-water-gas flows.

15.2.5 Neural Networks by EDS-Scicon and CALtec Ltd (JIP - BP a partner)

The current JIP started in 1992. The aim of the project is to develop a multiphase metering technique which uses advanced parallel processing - neural networks - to apply pattern recognition to the dynamic signals from relatively simple, low cost instrumentation such as gamma ray densitometers and pressure transmitters.

Neural networks have the capability to “train” or “teach” themselves. to recognise “features” from complex phenomena. The idea is to “train” the neural network to relate “recognised features” to the flow rates of oil, water and gas.

Activity, so far, has involved “training” different structures of neural network on flow loop multiphase flow data and some gamma ray densitometer field data (Forties Echo-Alpha pipelines) supplied by the multiphase flow group within BP. The object was to give EDS-Scicon the opportunity to work the networks on multiphase flow and demonstrate the technique. Typically, this has involved using a data set of 40 flow rates and predicting each data point (oil, water and gas flow rates) by training on the other 39.

Flow rate relative error data produced by this exercise exhibit uncertainties broadly in the range of 10% to 30%. However, the work is at an early stage. One of the difficulties facing the approach is the general lack of abundant, reliable and accurate field data containing oil, water and gas flow rates and corresponding signals from sensors such as pressure transmitters.

The programme will need to address which types of sensor are most suitable in detecting characteristic features, and where on the pipeline they should be installed to pick these out.

The method requires specialist expertise to adapt and set-up the neural network software to the application. The cost of this and of the computing capability has not yet determined.

It works best where there is an abundance of data on which to train the neural network and when interpolating between these data but not extrapolating. Given the general non-reproducibility of multiphase flow this may make calibration of the system as a universal or standalone metering solution impossible in practice.

If the neural network approach can be provided at low cost, it could lend itself to applications where the system would be trained against other expensive equipment (other multiphase meters or test separator systems) where this has limited availability and therefore impedes the well testing and reservoir management capability.

15.2.6 Texaco Starcut Meter

This water cut monitor has been developed by Texaco EPTD. Jiskoot Autocontrol Ltd. now have the exclusive world-wide licence to manufacture and market the product. It uses measurements of the phase shift and attenuation of 10 GHz microwaves caused by the flow stream to determine water cut.

The instrument will measure water cut of emulsions containing little or no free gas. It works with oil-continuous and water-continuous emulsions, i.e. water cut from 0% to 100%. Jiskoot Autocontrol can quote a number of oil industry site applications and state that gas volume fractions of up to 25% can be handled. The instrument is already developing a field track record ahead of the other phase fraction/water cut devices described above, although these instruments have, of course, already demonstrated a measurement capability on gas fractions significantly higher than 25%.

Texaco's development effort has concentrated on accumulating a vast data base of "mixing curves" for a wide range of oils and water chemistries. This is stored in the system's microprocessor, packaged with the sensor in a compact unit. The unit can continuously perform an auto-calibration in field service to ensure the correct mixing curve (the curve which traces oil/water ratio from the microwave phase shift and attenuation measurements).

The detection speed of the device is quoted as >300 samples/second It is claimed that it can cope with variations in fluid properties such as salinity. This could give it a significant advantage over the CMR/Fluenta and MFI systems which rely on operator input of salinity, which therefore must be fairly stable, and equations to allow for its variation with temperature.

The Starcut meter determines phase properties such as crude oil SG or density and water salinity and provides output signals which track these properties. The ability to track salinity is an attractive option for reservoir management. It could provide an indication of water flood progress or potential problems with premature breakthrough of injection fluids, for example.

Texaco claim the Starcut device can tolerate up to 25% GVF in the flow stream at present, though no data have been published to substantiate this. They state that development work continues in order to increase the free gas content acceptable and the capability to measure the gas fraction.

The sensing flow path is of rectangular cross-section of approximately 10 mm by 5 mm normal to the flow path. This limits the device to rely on a slip stream sample arrangement in the majority of applications since it clearly imposes a throughput limitation (quoted at 200 bpd liquid).

Should the capability to measure water cut with free gas presence (i.e. in multiphase flow) improve to a sufficiently high GVF, the Starcut cell is an attractive option for installation as a slip stream unit in combination with the BP multiphase flow meter. It would be used to provide the required water cut reading. It would readily fit via small bore piping (perhaps 20 mm to 25 mm diameter) as a compact side unit. Furthermore, the unit is available to the highest pressure ratings.

The present Jiskoot pricing for the off-shelf unit, of approximately $70,000, would also make the Starcut unit a preferable option for the BP meter relative to the MFI and Fluenta measurement systems.

The throughput limitation of the Starcut monitor and its consequent limitation as a slip stream device could be viewed as disadvantageous relative to the Fluenta and MFI full bore measurement cells. The test work with the BP screw meter has indicated that it enhances mixing of the water and oil (this has been demonstrated by taking side stream samples from the meter manually during flow loop tests). The combination of the screw meter with the Starcut monitor would be expected to provide the latter with a well mixed and representative sample from which to derive water cut.

Calibration of the Starcut monitor is essentially a factory/laboratory exercise. It subsequently performs its own auto-calibration once installed and in service.

15.2.7 Kongsberg Offshore AS - Capacitor Cross-Correlation Multiphase Flow Meter

In 1992, Kongsberg Offshore commercialised a multi-capacitor multiphase meter developed by the Shell Exploration and Production laboratory (KSEPL) in the Netherlands.

The measurement spool comprises two closely spaced (10 mm apart) parallel plates mounted across the centre of the pipe. The spool is mounted in horizontal pipeline with the plates vertical and in the plane of minimum resistance to the flow stream (i.e. aligned in the direction of flow). The plates support a column of small rectangular electrode pairs which spans the diameter of the pipe. Two more electrodes are positioned further downstream on the plate, one near the top and one close to the bottom of the pipe.

The meter has been designed to measure flow rate of oil, water and gas in slug flow. The gas-liquid interface can be sensed by the electrodes. The flow perturbations sensed at the upper and lower downstream electrodes are cross-correlated with the corresponding signals of the upstream electrodes.

Simultaneously, the capacitor outputs are used to determine the cross-sectional area occupied by gas and liquid in the pipe by measuring the liquid hold-up. This information is combined with the cross-correlation velocity measurements to determine gas and liquid flow rate. The technique processes the electrode signals in such a way as to make some allowance for the proportion of gas entrained in the main body of the liquid.

The modelling assumes a slug velocity equal to the gas pocket bulk velocity. Water cut is determined from capacitance sensed by the lower electrodes immersed in the liquid film of the slug flow. Currently, the unit only works for oil-continuous emulsions.

Shell state that units have been undergoing field trials in Oman and Gabon with “consistent” results compared to test separator equipment although exhibiting some bias error. They intend to continue funding development to extend the capability of the system to operate in non-intermittent flows and water-continuous emulsions (though they admit the latter could prove unfruitful). Low pressure flow loop results published show gross errors in flow rate measurement outside of a cluster of points lying within the intermittent region of the flow regime map.

Shell's intention has been to target this type of meter at remote land-based sites with significant pipeline between wellhead and manifold where well defined slug flow occurs. They claim to have a number of such applications and state that, in these cases, the level of accuracy achievable with the unit (quoted at between 10% to 20% of flow rate) is acceptable.

In sacrificing accuracy by the nature of the technique, Shell have attempted to evolve a relatively low cost unit, which could allow one meter per well. (Less than $100,000 unit price, including industrial PC, has been quoted with a target for Kongsberg to reduce this to $50,000 during the course of the collaboration between the two companies). Also, Shell wished to avoid deployment of nucleonic sources (common to most of the other techniques) at remote locations on land.

Currently, the units are available in 3" and 4" line sizes to ANSI class 600 rating. The 4" sensor spool weighs 90 kg and is 556 mm in length. The Kongsberg data sheets do not include a sour service specification at present.

In principle, the cross-correlation measurements to measure velocity should not rely on flow calibration. However, as noted above, Shell have reported some systematic errors during field testing. The compositional calibration of the capacitors relies on readings with the separate phases present. In this respect, the method has similar set-up requirements to the Fluenta/CMR system which also, in effect, relies on knowledge of the capacitance reading on single phase fluid.

152.8 Further Comments

The above review of multiphase metering techniques covers the most advanced equipment of which the author has technical knowledge. There are other developments of which the multiphase flow group is aware.

Schlumberger (Cambridge Research) are believed to be working with a multi-electrode capacitor technique, where rotating electrical fields are applied to multiple electrodes positioned circumferentially around the pipe. This type of system attempts to work with non-uniform spatial distribution of the phases within the pipe cross-section. It has been investigated elsewhere also.

Arco are known to have tested a system relying on separation of the gas and liquid phases. Co-operation between the multiphase flow group and Arco in conducting the Prudhoe Bay field trials of the BP multiphase metering system may reveal more details of this.

Texaco also have worked on a much publicised separator system. This, so called, "multiphase meter" comprises a 10m high structure weighing many tonnes, which houses an inclined separator system and associated pipework. Vortex shedding and variable area single phase metering systems are used in the gas and liquid legs respectively off the separator. Texaco microwave technology is used in a liquid leg to measure water cut. The projected cost of the unit when publicised in 1990 was up to $3 million. Field trials on the Tartan-Highlander subsea tie-back system in the North Sea met with difficulties such as associated with foaming of the multiphase stream. The development work by Texaco and Jiskoot Autocontrol was continuing and the package was scheduled for installation on the subsea Highlander template in 1992. The author is not aware of any further publicity since 1991.

A downhole venturi meter has been developed jointly by Aberdeen-based company Exal and BP. This device cannot tolerate the presence of any gas and presumably relies on a uniform and well mixed emulsion. A prototype was commissioned in an ESP well at Wytch Farm in January 1993. The device measures total liquid flow rate and future installations are expected also to yield a water cut reading by head measurement of the liquid column. At the time of writing the author has still to establish details of the Exal development programme and fundamental basis (for example, oil-water flow regime behaviour downhole). Such equipment could provide the most suitable means of measuring well fluid rates in certain applications. It might also be used as part of overall multiphase metering schemes in combination with other hardware under development.

Several other multiphase meter development programmes are or have been in existence. For example, the Euromatic combined mixing/side-stream separator system reached and underwent a field trial stage but seems to have faltered owing to its significant lack of sound fundamental basis and understanding of multiphase flow behaviour. The Mixmeter JIP programme launched by Imperial College, London, in 1992 and combining some elements of the basic technology already covered in the JIPs listed above, is an attempt which might bring new options to the fore in the future. The purpose of this review is to describe in some detail what is considered to be the most soundly based and advanced technology to date. Not all of the known approaches have been listed.

Most, if not all, of the techniques covered in the above review currently have a high degree of measurement uncertainty at extreme GVF (95% and above). The total volumetric measurement of the BP screw meter has been shown to be the exception to this. Annular liquid flow patterns in the rotors are believed to assist in sealing the clearances. (Also, the venturi meter is known to be used by Shell for measuring total volume flow rate in gas condensate well flows and ultrasonic wet gas meters have been developed. However, for these two cases, gas flow rate but not liquid rates in gas condensate streams are measured).

Flow patterns which are predominantly annular in nature, which can persist at lower than 90% GVF as line pressure increases, currently cannot be handled by the Fluenta/CMR and MFI phase fraction measurement systems (or, indeed, the Shell/Kongsberg meter). For these instruments, some form of flow conditioning to perturb the annular flow pattern would be needed where such conditions are expected to exist during the service life of the flow meter.

15.3 Summary

The main techniques for multiphase metering which have emerged can be broadly classified into two categories. One type of approach employs flow control or conditioning in order to establish reproducible conditions at the measurement section. Other techniques have adopted the principle of cross-correlation velocity measurement in combination with phase fraction measurements to establish the phase flow rates. This latter approach is highly susceptible to the complex and generally non-reproducible nature of multiphase flows.

The general recommendation of the philosophy is that those multiphase metering techniques employing flow pattern-independent principles, that is, the former of the above two approaches, will most readily facilitate a calibration methodology and, therefore, be of use. For certain types of these technologies, such as the BP screw meter based system, test experience is demonstrating a way forward using relatively straightforward and low cost factory calibration methods utilising single phase fluids. All measurement techniques are likely to require wide test experience to evaluate and evolve standards. The level of accuracy and its qualification required within the oil industry will dictate the extent of the testing effort. Please see section 16 for further discussion of the calibration of multiphase metering systems.

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References

(1) 'Study to Install a Multiphase Meter Subsea, BP Engineering', BP Engineering Technical Note, ESR.93.ER.036.

(2) 'Proposal to Install a Multiphase Meter Subsea on Cyrus', BP Engineering Technical Note, ESR.93.ER.035.

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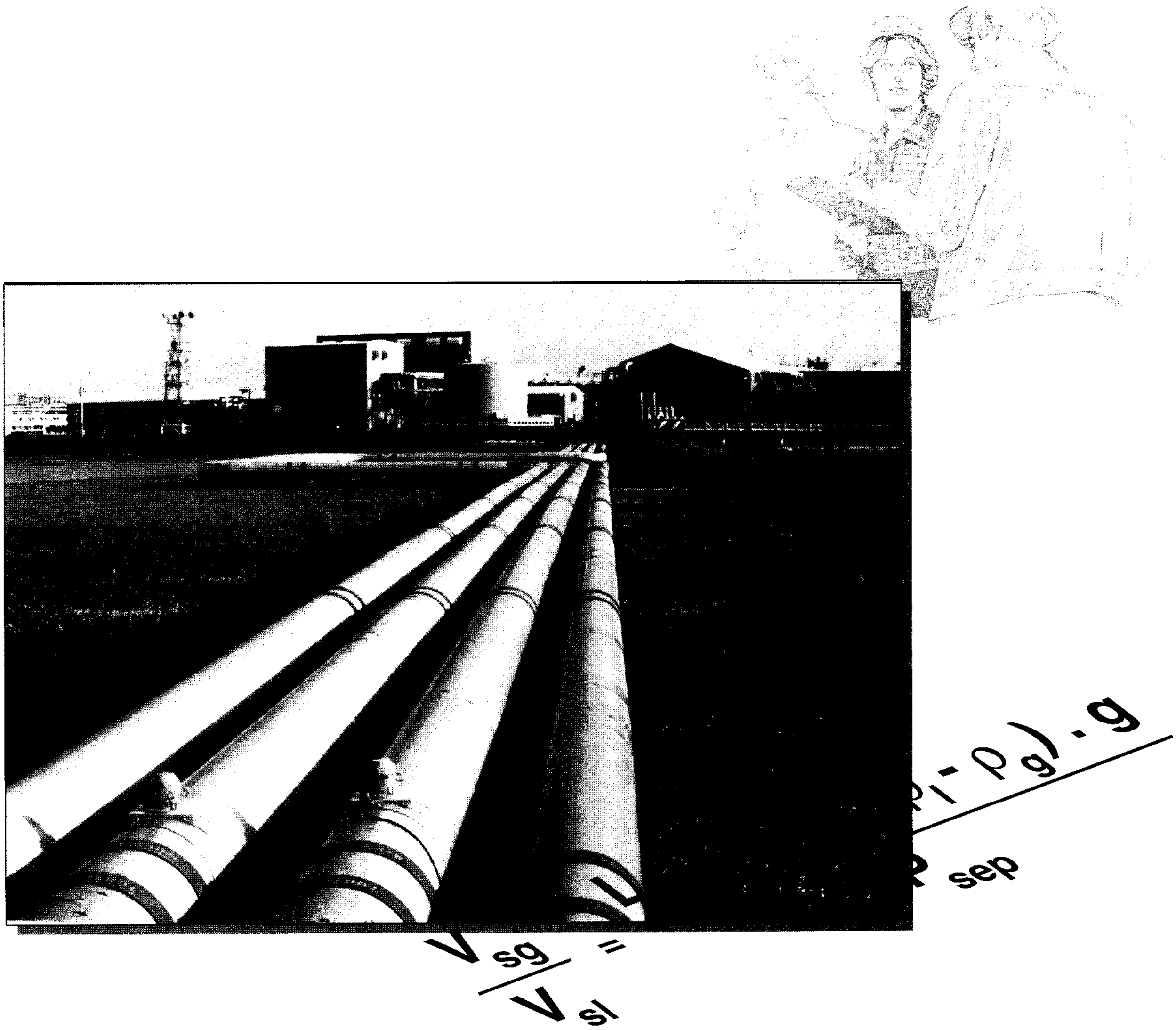
Section 16. Meter Calibration

Calibration of Multiphase Meter Systems



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16.1 Introduction

In this section, the basic aspects associated with calibration of multiphase flow meters are listed. The special challenges posed by multiphase flow - in particular, the difficulty of ensuring reproducibility, without which a calibration cannot exist - are then highlighted. In the face of these challenges, some theories, based on the BP test experience of metering multiphase flows to date, are presented as to how multiphase flow meter calibration philosophy might develop in practice.

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16.2 Aspects of Flow Calibration

16.2.1 Reasons for measurement

The multiphase metering systems currently under development have been targeted at well flow measurement for reservoir management and allocation. The significantly more stringent accuracy usually associated with custody transfer or "fiscal" flow metering has not been addressed.

Clearly, any means of calibration of a multiphase flow meter should be appropriate to the level of accuracy required. As a general rule, the reference system should be prone to significantly smaller uncertainty levels then the meter under calibration. A factor of 10 superiority of uncertainty levels of the reference meter might typically be desirable in some conventional single phase calibration scenarios.

16.2.2 Traceability

Strictly, the measurement uncertainties of the reference metering should be qualified by primary calibration against a recognised standard. This will take place at regular intervals dependent on the circumstances.

16.2.3 Proving

The specific case of a reference calibration or master meter checking system installed in bypass pipework at the metering station which can periodically be switched to measure flow and check or calibrate the flow metering system in-situ. The proving system can be a fixed installation or be a portable servicing system.

16.2.4 Repeatability

Repeatability is a pre-requisite to flow meter accuracy and is determined by means of a specitied number of repeat comparisons against a reference measurement. This procedure relies on repeatability of flow conditions (and, of course, of the reference measurement).

A calibration based on a flow loop test, such as by the manufacturer of the flow meter, relies on the concept of a repeatable or predictable and reproducible flow condition.

For example, the requirement for specified pipe straight lengths and flow straighteners in turbine meter installations ensures known calibration flow conditions are re-produced in single phase flows. Indeed, single phase pipe flow is well enough understood and its behaviour sufficiently predictable that an orifice plate can be manufactured to controlled tolerances such that flow calibration is not needed. The accuracy of the orifice plate can then be traceable by metrology

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16.3 Multiphase Flow and Calibration

In measurement, the concepts of repeatability and traceability are the foundations of calibration.

Multiphase flows are more complex and less well understood than single phase flows. They are generally not predictable in the same way as single phase pipe flows. For example, multiphase flow behaviour cannot be guaranteed in terms of a steady and defined velocity profile even if certain rules regarding pipe length are obeyed. In general, the multiphase flow behaviour depends on the entire history of flow path from source and upon influences from a significant distance downstream. Experience has shown that the nature of the multiphase flow patterns is installation dependent and therefore not necessarily repeatable or reproducible between one installation and another. A manufacturer's multiphase flow loop calibration would not, in general, necessarily represent the field installation.

Even for the case of a proving system, the change in flow path in bringing on line the prover could alter the multiphase flow behaviour at the flow meter and, therefore, its response characteristic.

Clearly, the foundation of repeatability or reproducibility is much more difficult to realise in multiphase flow.

In addition to the installation dependency of the multiphase flow behaviour, the nature of flow pattern can drastically change through the service life of the flowline. For example, a multiphase flow meter might be subjected to a stream of mainly oil flow containing dispersed gas bubbles in the initial stages of service. As gas content increases with the depletion of the reservoir, the meter could be subjected to intermittent flow behaviour, such as plug or slug flow. Finally, in the extreme case, the meter could have to cope with annular flows of high gas content and significantly higher total fluid volume throughput than in earlier service life. Additionally, the latter stages of reservoir depletion and any associated secondary and tertiary recovery schemes could result in an increasing portion of water in the flow stream.

Thus, in terms of calibration (quite apart from meter performance) multiphase flow metering can pose significant new demands of flow pattern behavioural change and flow rate range (turndown).

The above discusses calibration for multiphase flow meters by relating to the case of single phase flow. It has highlighted the difficulty in realising reproducibility in multiphase flows to determine measurement response. But, of course, measurement response in multiphase flow relates to more than one response, or signal, which must be related to more than one (three - for oil, water and gas) flow rate. The calibration process in multiphase flow, then, can be even more involved than for existing single phase methods from this consideration also.

The multiphase meter measurement can involve several simultaneous sensor responses. The combined influence of the errors in all these signals will determine the overall accuracy of the multiphase meter in measuring the phase flow rates and this must be determined by calibration procedure.

Further, most, if not all, of the multiphase metering techniques under development rely on knowledge of the individual phase properties, such as oil, water and gas densities, in order to derive the phase flow rates from the sensor inputs. The calibration and overall specification of accuracy and repeatability for any multiphase metering installation must include an account of the uncertainties associated with the measurement or prediction of these properties.

The greatest cost benefits attributable to multiphase metering are expected to be realised where the new technology is installed subsea at well pads remote from the processing facility. The requirement for remoteness subsea poses another significant challenge in terms of calibration.

How will it be ensured that a meter remains on calibration specification?

This is important not just in terms of deciding upon the level of calibration as necessary routine. The possibility for major changes in the nature and throughput of the flow stream, as discussed above, could conceivably dictate a requirement for calibration checking or proving on-line. Is proving of a remote meter in a multiphase pipeline practical? Would such a requirement go some way towards negating the advantages of a multiphase metering scheme?

16.4 Multiphase Measurement Standards in Practice

From the above, it can be concluded that, just as conventional single phase meters cannot be applied to multiphase flow measurement, calibration methods based on single phase flow understanding cannot be relied upon.

New or modified approaches of calibration need to be adopted for multiphase metering.

It could be argued that, given the general non-reproducibility of flow behaviour in multiphase installations, measurement techniques exhibiting minimal sensitivity to multiphase flow pattern behaviour will not only lend themselves to calibration, but may also provide the means to defining calibration technique and reference standards.

This would be achieved on the basis of proof by experience and physical argument. The degree of rigor involved in this, and hence the level of development and time to evolve a calibration methodology, will depend on the flow measurement uncertainty levels acceptable and the extent of qualification required by the oil companies and government agencies. These parties will need to take an involved and informed view of what the technology developers and manufacturers are able to provide.

Some multiphase metering systems under development such as the BP patented screw meter and Framo static mixer based system (see Section 15 - Multiphase Meter Systems and Their Application) involve techniques which attempt to minimise flow pattern dependency of measurement by controlling the flow behaviour. Only experience of tests in several installations of widely varying flow conditions, involving comparison against accurate single phase references downstream of separators, will develop and confirm the uncertainty levels achievable and the universality of the calibration of these approaches. The alternative to protracted testing is for all concerned (oil producers, government agencies, manufacturers and technologists) to accept argument based on sound theoretical principles with only limited flow loop and field test evidence of proof of theory. Given its physical complexity and comparative youth as a branch of fluid dynamics tackled by engineering science, multiphase flow will not readily lend itself to the latter approach

Other multiphase metering approaches have addressed measurement using sensors which primarily respond to the composition of the fluid stream. These development programmes have concentrated on ways of measuring the phase fractions or ratios but latterly some have adapted the sensing elements to yield additional cross-correlation measurements of fluid velocity.

The calibration of such sensors to measure mixture composition is a relatively straightforward proposition where it can be ensured that the phases will be well mixed or uniformly distributed within the sensing volume in service. Typically, the techniques for measuring phase fractions, described in Section 15, apply mixing models relating sensor response to mixture composition. The calibration curves by these models are fixed by measuring the response signal on the single phase fluids.

The tendency of multiphase flow to form segregated patterns of the phases in the pipe can invalidate calibrations by this approach. Where the sensors have a fast response, it is possible for these devices to track fluctuating flow stream composition provided the phases are distributed uniformly in the cross-sectional plane. This can be achieved by mounting the composition sensor in a vertical spool although some form of flow conditioner may still be required to break up annular flow (or intermittent annular flow pattern) when this establishes upstream of the

sensor.
Some highly sophisticated types of multi-sensor compositional devices are being investigated. These systems are being developed to have a rapid response which accounts for phase distribution within the sensing zone. The complexity of these approaches may make them difficult to prove in practice.

If composition can be accurately dynamically tracked, it must be combined with simultaneous total flow rate or velocity measurement to provide phase flow rates. Even in dispersed or uniformly distributed flow patterns, slip (the difference in velocity between the phases) can be significant. Thus, either some form of flow conditioning to eliminate (or minimise) slip is required or else the individual in-situ phase velocities must be measured.

Cross-correlation measurement of velocity has the potential in principle to be “absolute”. That is, flow calibration is not necessary. Accurate measurement of sensor spacing and “time of flight” will give accurate measurement of velocity.

However, the accuracy of the “time of flight” measurement as derived from a cross-correlation function, or more specifically, what it actually represents, is greatly influenced by flow velocity profile, turbulence and signal noise. In multiphase flow, as discussed already, velocity profile is generally highly complex and non-repeatable in nature and slip generally occurs between the phases. Accurate cross-correlation measurement would rely on a flat cross-sectional velocity profile or multi-velocity measurement of sufficient detail to allow for profile shape.

What is more, mixture property cross-correlation devices tend to sense particular features of the fluctuating flow pattern. Some techniques now being developed are sensitive to the interface between liquid continuous bulk and large gas pockets. Others use sensing fields which are most sensitive to certain regions of the pipe cross-section such as the near-wall. It does not, therefore, follow that these cross-correlation techniques measure bulk (average) flow velocity which can be multiplied by cross-sectional area (adding further uncertainty) to give flow rate.

At least two projects are under way to measure more than one cross-correlation feature in order to derive multiple velocity information which relates to the different phases. This is at an early stage of development but could prove impractical to qualify as universally field applicable.

The BP screw meter and Framo mixer approaches of flow pattern independence by flow control, combine a total flow measurement method with a mixture phase ratio measurement. This can allow total flow throughput measurement and phase fraction measurements to be combined to yield phase flow rates. The objective, as already stated, is to overcome the multiphase flow characteristics of generally non-reproducible flows, complex velocity pattern and phase slip. Within the BP in-house development programme, this has been achieved by combining the BP meter total flow measurements with a water cut reading derived from in-situ phase fraction measurement.

If such flow pattern-independent techniques can be developed to an acceptable degree (i.e. acceptably repeatable signal responses for given flow rates in all applications) then flow calibration on a manufacturer’s or standards institution multiphase flow loop (oil, water, gas) will be meaningful.

Indeed, if such metering techniques are sufficiently successful then certain types of total flow rate meter could be calibrated on single phase fluids, such as water, since flow pattern independent response in multiphase flow is, by its nature, an insensitivity or predictable response to the density and viscosity of the fluids being measured.

Given the general necessity of only needing to calibrate mixture phase fraction instruments on the basis of single phase fluid properties, the combined metering system could then be calibrated at the factory or laboratory site, using relatively simple and established procedures and single phase fluids.

This type of approach will result in lower cost multiphase metering solutions than calibration requiring the use of relatively complex and expensive oil, water, gas flow loops. (For example, one particular multiphase flow meter test loop built by another oil company in Europe cost on the order of $1.6 million to build in the mid 1980s. It is a low pressure, 4" nominal bore loop using gas-oil, tap water and air which are metered to traceable standards. This requires regular primary calibration of the multi-spool reference metering and six monthly checking of the proving system by the national standards authority. Additionally, full-time maintenance back-up is employed. This degree of rigor is necessary in order to qualify multiphase metering to +/-5% relative errors; - a target usually quoted for multiphase flow meters).

Of course, after initial set-up, calibration and installation, there remains the question of subsequent calibration or proving in service. Further thought on this is prompted in the review of multiphase metering techniques in Section 1.5. In that section, questions relating to the practicality of certain field proving approaches are raised. The benefit of ROC experience and input will be necessary in order to tackle the proving issue in particular. Ultimately, the proving requirement of the flow pattern-independent techniques could evolve to a basic requirement where simple static, dimensional or self diagnostic checks will ensure measurement to within specified uncertainty bands.

In summary of the above, two key approaches to multiphase metering have emerged. One approach involves conditioning or controlling the flow stream to ensure reproducible flow conditions at the measurement section. The second approach uses the cross-correlation technique in combination with flow composition sensors to derive the phase flow rates but is highly susceptible to the complex and generally non-reproducible nature of multiphase flows. It could be argued that multiphase metering systems based on flow pattern-independent principles (the first of the above approaches) will most lend themselves to calibration and, therefore, usefulness. Indeed, such techniques may well permit simple factory calibration and set-up procedures. Ultimately, these techniques could rely on only simple methods of field calibration checking which do not necessitate comparisons against flow measurements by another "master" meter or proving system. All measurement techniques are likely to require wide test experience to evaluate and evolve standards. The level of accuracy and its qualification required within the oil industry will dictate the extent of the testing effort.

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16.5 Field Calibration or Proving

Certain questions, concerning the proving or in-service calibration of multiphase meters in particular, have been raised in the preceding paragraphs. All of the emerging multiphase metering techniques are likely to need some level of in-service checking, even if this only involves an initial, and subsequently infrequent, comparison against a reference measurement.

In relation to these questions, it is recommended that consideration be given to the implications of the following approaches to the problem of field calibration for practicality and accuracy.

16.5.1 Production Train Reference

Testing using the production train single phase metering system as a reference. This could require that operations accommodate well shut-in periods, possibly flowing only one well at a time. Alternatively, it could involve testing by difference where one (or more) well is shut in at a time.

A more acceptable variation on this theme might be possible, with some types of multiphase flow meter at least. Experience with the BP multiphase flow meter at Prudhoe Bay, for example, demonstrated repeatability for all the wells tested at the well pad. These all produced within a tight band of gas volume fraction (although water cut and flow rate and behaviour varied markedly between wells). In such circumstances, the multiphase meter measurement would simply be checked by comparing the sum of the individual well measurements with the total production of the well site.

In the case of the BP multiphase meter, it can be assessed in terms of the total mass flow rate measurement of the stream flow rates or indeed to make an allowance for mass transfer between the phases and changes in phase volumes from multiphase to reference measurement station.

Such a method, based on summing the production of the wells, would have to account for any difference in the flow rate from the wells when switched between the production line and the multiphase meter by-pass, in cases where multiphase metering is not available on every individual well. Some multiphase metering hardware introduces a pressure drop (see Section 15.2 covering the BP and Framo multiphase metering systems in the review of techniques).

A further alternative, which could be considered in certain circumstances, would be to flow more than one well through the multiphase meter and compare its measurement directly with the production separator metering system receiving the combined flow from the same wells. Clearly, this approach could be constrained by the maximum flow rate or turn-down of the multiphase meter, which would dictate that some wells be shut in during the proving exercise.

16.5.2 Portable Test Separator

Although involving additional high cost, particularly off-shore, another approach to consider is the use of portable test separation equipment. This might be an option for multiphase metering equipment requiring infrequent checking in certain circumstances. In addition to reservations regarding the cost penalties of such an operation, this type of reference will be subject to uncertainty levels of an order associated with the multiphase meter itself. It would serve the purpose of a check rather than as a means to calibration.

16.5.3 Multiphase Proving

This might consider adapting existing multiphase (or single phase - such as compact swept piston devices) metering technology for portable and non-continuous use. Certain types of multiphase metering technology, designed for prolonged service in extreme conditions of flow, might be made more cheaply for this purpose. It could also be manufactured and operated for more accurate performance. This would be possible with a switch of design emphasis from robustness in favour of accuracy. It would be realisable with short service exposure intervals, coupled with regular inspection and maintenance and operational precautions not possible in the running of the installed multiphase metering system.

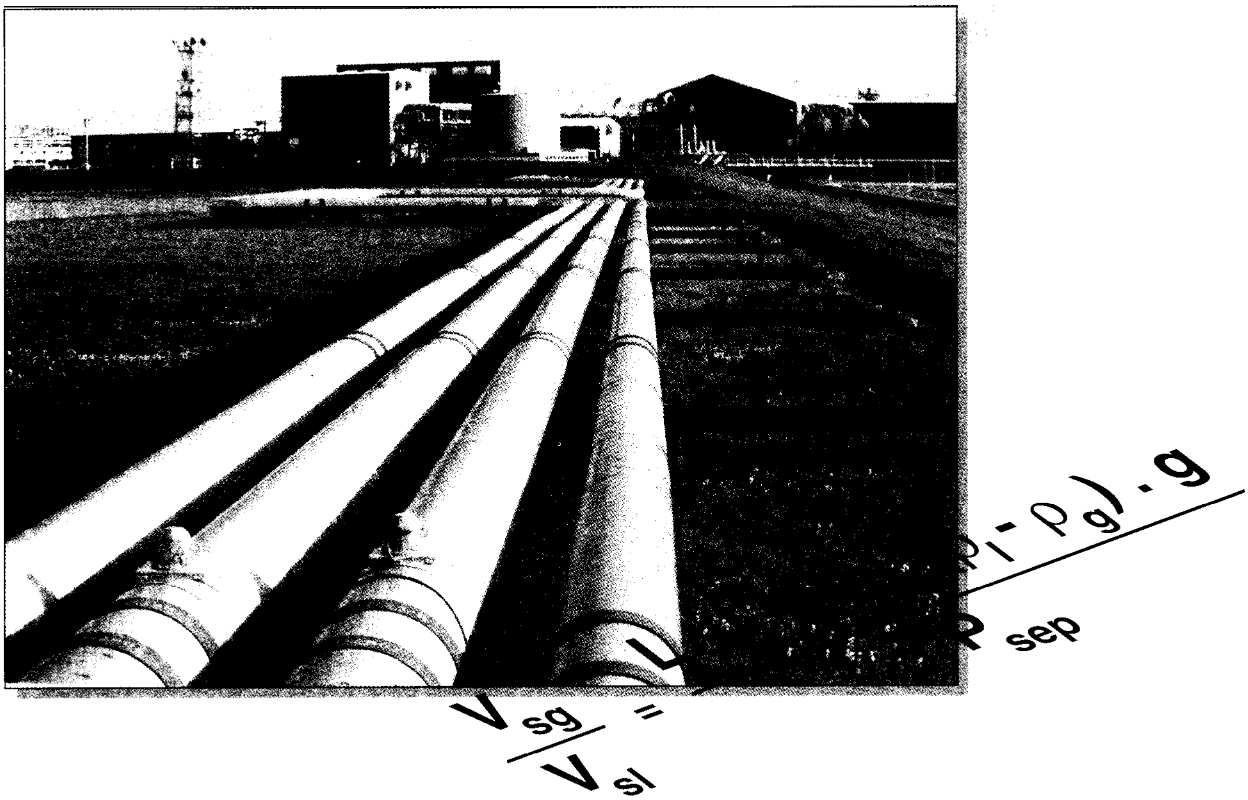
Using the BP screw meter again as an example, the clearances between the rotors and stator (approximately 1 mm) have been made sufficiently large to ensure a high degree of reliability in the presence of particulates. It would be possible to improve the current accuracy of this system by reducing the clearances. Furthermore, limited production fluid exposure, avoidance of proving coincident with certain well and reservoir workovers or operations, and the use of a serviceable fine gauze upstream of a meter in a proving skid would permit a design typical of high accuracy fiscal applications of total volume flow rate measurement in single phase liquid flows.

16.5.4 Concluding Comment

It will be recognised that the acceptability of certain approaches could differ between land or surface and subsea installations. Any further suggestions on field calibration are invited.

Section 17 Pressure Boosting

Multiphase Pump Systems and their Applications



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17.1 Introduction

If the pressure within the reservoir is inadequate to transport the hydrocarbon fluids at the required rate to a suitable place for processing them, then some form of pressure boosting will be required. This may be by enhancing the pressure of the reservoir itself, eg. by pressurised aquifer support and /or gas re-injection, pressurising the reservior from above. An alternative is the use of down-hole single phase pumping, which does not affect the reservoir itself, but raises the well head flowing pressure. Lastly boosting may be applied downstream of the well head where the flow is normally multiphase. This Section considers these options and in particular the use of multiphase pumping option.

This pressure enhancement may be required early in the field life or may only be required towards the end of field life when the reservoir pressure has fallen too low for the standard methods of transport to be adequate; never-the-less consideration of later means of pressure boosting should be considered in the 'whole-life-field' concept design. There are several forms in which this pressure boosting can be applied. Section 17.2 discusses the selection of the method, including the gas-lifting of wells (which is really a means of pressure drop reduction rather than pressure boosting). Further consideration is given to one of these means of pressure enhancement – the application of multiphase pumping systems, see Section 17.3.

Multiphase pumps generally have higher power requirements than those of single phase machines (pumps or compressors) performing similar duties. The theoretical power requirement depends both upon the duty conditions and the theory used in calculation. As a result attempts to compare performance in terms of efficiency tend to be misleading and are best avoided.

Although they are described as 'pumps' it would be more accurate to describe these multiphase pressure boosters as wet gas compressors, in that volumetrically gas is almost always the dominant phase. The volumetric throughput, that is the volume of the feed fluids at the feed conditions, is the most important factor in pump design; in addition the gas volume fraction (GVF), the feed pressure and the pressure ratio are required.

Table 1 outlines the capabilities of the pump types which presently seem to have some commercial potential.

Table 1. Multiphase pump types

Multiphase Pump Types (Only Systems with established commercial potential are included)

Pump type	Status	Subsea ?	Power source Electrical	Hydraulic	Sand Tolerance	General performance characteristics Flow	Delta P.	GVF
Rotodynamic								
Multistage Helico-axial	f	Yes	Yes	**Yes**	**Yes**	M	Lor M	H or M
	D	Yes	Yes	?	Yes	H	Lor M	H or M
Multistage Centrifugal	P/L	Yes	Yes	Yes	Yes	L	M	L
Contra-rotating axial (CRA)	Fr	?	Yes	?	No	M	M	H
Positive Displacement								
Twin Screw	f	**No**	Yes	**No**	Limited	M	H	M
Reciprocating Piston (MEPS)	Fr	Yes	Yes	?	No	M	H	H
Hydrobooster	P/L	Yes	Yes	No	?	?	?	H
Others								
Jet pump	£{4}	yes	**No**	**Yes**	**No**	L	L	H
'With Separation'								
VASPS	P/L	Yes	Yes	**No**	**Yes**	MI31	N/A{5}	H
KBS	P/L	Yes	Yes	No	Limited	H	H	H

Key:

Status:

f = Commercial
D = Commercial designs available
FT = Field tested
P/L = Prototype and/or Lab. Tested

Flow/DP/GVF	H	M	L
Flow {2}			
m3/h (MBD)	>500 (75)	500- 100	<100 (15)
Delta P. (Ratio)	>5	5 to 2	<2
GVF%	>90%	90% - 50%	<50%

Notes:

{1} >95% with liquid recirculation
{2} Total volume at Feed Conditions
{3} Limited by separation in dummy well and ESP capacity
{4} As WELLCOM system
{5} No Delta P for Gas

17.2 Pressure Boosting – Methodology of Selection 'BOAST'

The BOAST (Boosting Options Analysis Software Tool) program, which has been developed within BPX for Foinaven, can be used to analyse the production profile achievable by the application of pressure boosting techniques for a particular case. The model takes account of three main areas:

Reservoir
Curves of reservoir pressure, water cut, GOR, and PI, each expressed as a function of produced reserves, are loaded into the program. Using this data, the program can calculate the entry conditions into the well-bore throughout field life as the reserves are produced. In addition the program can restrict the maximum production into an individual well; this later restriction is to incorporate reservoir drawdown restrictions such as imposed by gas and/or water coning and sand production.

Production System
Details of the well-bore, flow lines and risers (length, diameter, number etc.) are entered into the program. Using standard multiphase flow correlations, the back pressure of the system for a range of flow rates can be calculated. Also included in the software is a description of the various boosting techniques to be analysed (generalised pump curves, maximum flow rate, maximum power demand, maximum pump gas void fraction etc.) Using input data from the reservoir description, the software is able to calculate the maximum flow rate achievable for the various boosting options.

The boosting options available are as follows:

- Well gas lift
- Electrical Submersible Pumps
- Wellhead Multiphase Pumps
- Sub-sea Separation (with single phase pressure boosting)
- Riser gas lift

Host Facility
The software takes into account the capacity constraints of the host facilities and is able to limit production rates from the field if oil, gas, water, gross fluids or power constraints are exceeded.

Families of production profiles can be generated for a given field, production system configuration, and host facility constraints for a range of boosting options, and thus a structured choice can be made. Note that the assumptions concerning multiphase pump capacities and performance is based on FRAMO information and will need to be broadened to include the greater experience now available.

For further information on BOAST refer to Multiphase Production Systems, Transportation Group, BPX. at Sunbury.

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17.3 Outline Review of Multiphase Pumping Systems

17.3.1 Introduction

Pumps may be of either rotodynamic or positive displacement type. The two types operate on different principles and it is practical to present their performance curves differently.

Rotodynamic pumps operate by imparting velocity (kinetic energy) to the fluid and then decelerating it to convert the kinetic energy to pressure. Consequently at any particular flow rate a given specific energy or head is imparted to the fluid. Thus the differential pressure generated is proportional to fluid density. The governing variable is flow rate through the pump and the characteristic is represented as shown in Figure 1 a below. If differential pressure is to be shown then the characteristic depends on density as shown in Figure 1 b.

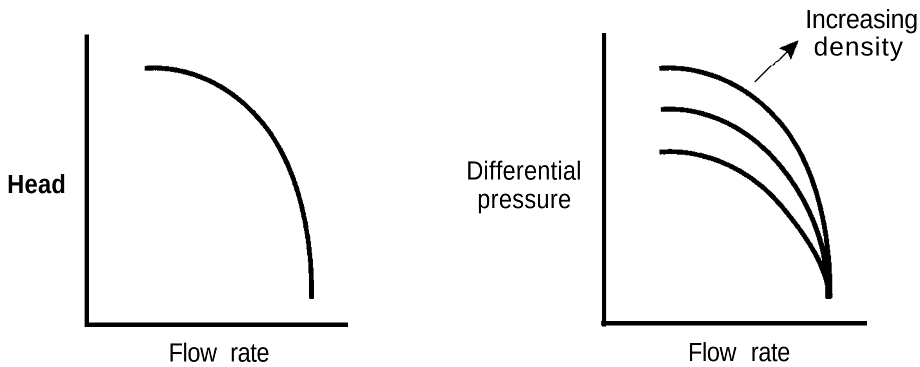

Figure 1 a and 1 b

Positive displacement pumps operate by trapping a pocket of fluid at the pump suction, carrying it through the pump and forcing it out at the discharge. The differential pressure achievable is limited by the strength of the pump to withstand pressure and the effect of the differential pressure on the internal leakage. The characteristic of such a pump is conventionally shown as in Figure 2a.

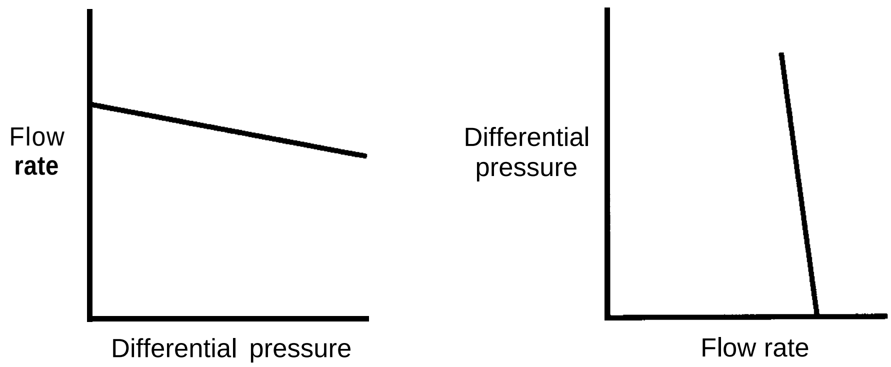

Figure 2a and 2b

If the axes are interchanged to achieve a representation similar to that for rotodynamic pumps then a performance characteristic similar to that in Figure 2b is produced. Comparison with Figure 1 b shows the difference in characteristic between the two types of pump.

17.3.2 Application to Multiphase Duties

If the two pump types are applied to Multiphase duties their characteristics remain unchanged but the effect of the presence of the gas phase must be considered.

(a) Rotodynamic Principle

If the rotodynamic pump is handling a mixture in which the two phases are well mixed it can be regarded as handling a fluid of density and compressibility between those of liquid and gas.

In order to calculate the head it is necessary to consider the work done on the compressible gas and the incompressible liquid separately. Even in cases of high GVF the mass of liquid present is much larger than the mass of gas and isothermal compression may be assumed.

This leads to:

$$H = \frac{(1-\alpha)(P_2 - P_1) + \alpha P_1 \mathrm{Ln}\,(P_2 / P_1)}{\rho_M}$$

where:

ρ_M **= Mixture density =** $(1-\alpha)\rho_L + \alpha\rho_G$

H = Head
a = Gas Volume Fraction (GVF)
P = Absolute pressure
ρ = Density

Suffices:
1 = Suction
2 = Delivery
G = Gas
L = Liquid
M = Mixture

Analysis of published data shows that for a wide range of GVF a single curve represents the pump performance at a given speed.

Conversion of performance from one speed to another can be done using the normal affinity laws:

- Capacity is proportional to speed
- Head is proportional to speed squared

The effect of speed change is shown in Figure 3a below.

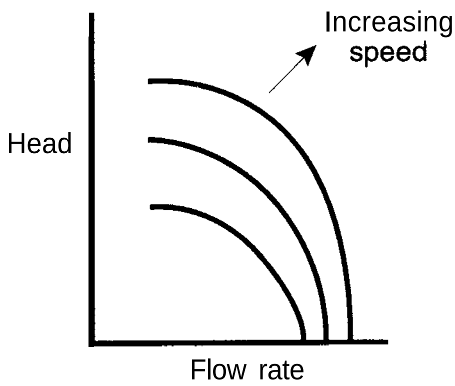

Figure 3a

(b) Positive Displacement Principle

Positive displacement machines trap a fixed volume of fluid at the suction and in general it makes little difference to the capacity whether the fluid is liquid or gas. As speed changes internal leakage remains constant while capacity varies linearly with speed. See Figure 4.

Twin screw machines tend to show increasing volumetric efficiency as GVF rises until at high GVF, typically more than 85%, volumetric efficiency starts to drop. This is thought to occur when there is no longer sufficient liquid present to fill the internal clearances and provide a seal. The lower density gas can leak through these clearances much faster than the liquid. The value of GVF at which volumetric efficiency is a maximum depends upon the number of pitches on the screw.

A typical curve is shown in Figure 5.

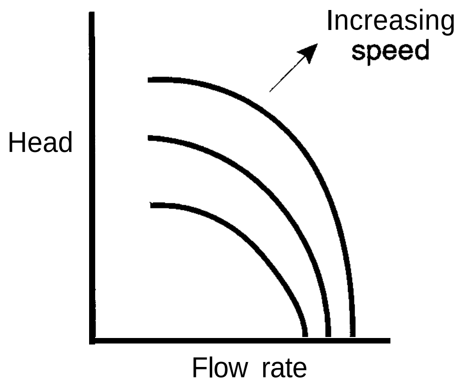

Figure 4

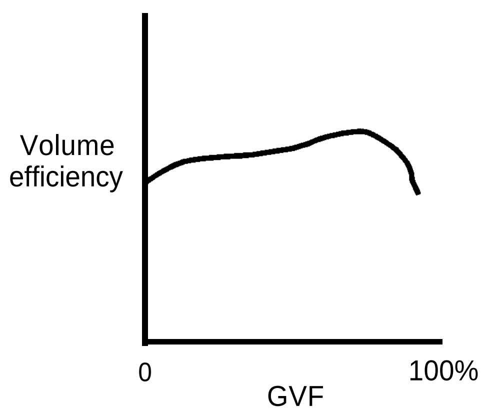

Figure 5

(c) Summary

Positive Displacement pumps tend to be of lower capacity than Rotodynamic but have a greater differential pressure capability and are less affected by changes in GVF, Suction Pressure and Differential Pressure. They are tolerant of viscous fluids but subject to wear by sand laden fluids.

Rotodynamic pumps have larger capacity but are much more affected by changes in GVF, Suction Pressure and Differential Pressure. Because head is a function of GVF, Differential Pressure and Pressure Ratio, plots of differential pressure performance can only be prepared for a given set of conditions. Viscous fluids degrade rotodynamic pump performance.

Because they do not depend on maintaining close clearances to maintain volumetric efficiency Rotodynamic pumps should be tolerant of sand laden fluids. However this has yet to be demonstrated and experience with other pumps running at similar speeds suggests that erosion may be a problem. The use of very hard materials may prove necessary.

(d) Rotodynamic Pumps

Helico-Axial

The leading rotodynamic multiphase pump is the helico-axial style of pump. The IFP designed 'Poseidon' pump and its derivatives have been the most successful. Extensive testing and production operation has been carried out on versions of this pump supplied by FRAMO (Frank Mohn) and Sulzer. It is most appropriate where high volume throughput is combined with high GVF at low pressure ratios, or where the GVF is lower and higher pressure ratios are required. The pump consisted of three basic modules:

(a) The Drive unit; this is generally an electric motor driving through a speed increasing gearbox; suitable electric motors for subsea application are under development (1 Q96) by Sulzer, the Nautilus Project, and FRAMO, their ELSMUBS Project. The SMUBS pump has a hydraulylic turbine drive, see Section 17.4.1.

(b) The Pump consists of a multistage assembly of impellers which have axial flow and high solidity. Shaft speed is high, typically in the range 4000-6000 rpm.

(c) The Homogeniser: it was found that pump performance was significantly improved if gas and liquid were well mixed; a homogeniser has been developed and is fitted immediately before the pump suction.

This form of pump is establishing a good track record of successful operation, see 17.4.1, and is considered to be 'commercial', at least in the 'medium' size range, say up to 500m^3/h.

Side Channel

Side channel pumps have high head/low flow characteristic combined with good gas handling ability. However they are of low efficiency and are used in low flow applications where the particular characteristics are valuable and the power penalty can be accepted. They also depend upon close internal clearances to control internal leakage and are therefore unsuitable for liquids containing abrasives, such as sand. It is expected that larger size pumps could tolerate larger clearances while maintaining acceptable efficiency, but no suitable pumps are presently available, and development effort has ceased. No commercial pump is in prospect.

Contra-Rotating

Two forms of multistage pump having contra-rotating impellers have been considered. One has stages arranged radially, similar to the Ljungstrom steam turbine, and the other type has stages arranged axially. The former machine is designated as Contra-Rotating Disk Compressor (CRD). Performance of a test machine was said to have been good on gas but poor when liquid was injected and development is either on hold or has been abandoned. The Contra-Rotating Axial Flow Pump (CRA) has been built by Frank Mohn and tested by Shell at their De Lier field for some 2 months, apparently successfully, although development is continuing. In both cases the purpose of employing contra-rotation is to give very high relative velocity between the two sets of blading so increasing the head rise per stage and minimising the number of stages required. Suitable application might be low/medium total volume, with high GVF and low/medium pressure ratio boosting.

Centrifugal

Conventional centrifugal pumps lose performance at GVF above about 10%. A modified impeller was developed which allowed effective operation up to about 50%. The modification involves the provision of holes in the shroud and gaps in the vanes which allow liquid to sweep gas away from the positions where it tends to accumulate.

Axial Flow

Some work on axial flow impellers showed promising results in the mid 1980s and further work has been concentrated on axial flow impellers having a high hub/tip ratio in order to minimise centrifugal separation. The impeller has been described as resembling an inducer which suggests it is broadly similar to the Poseidon (Helico-Axial) impeller. Testing at Texaco's Humble site is being progressed; claimed performance/efficiency figures should be viewed with suspicion since the basis of calculation has not been made clear.

Pumps intended for down-hole application with gas handling capability, say in the GVF range 30-60 % have been designed; the much higher suction pressure is responsible for the lower GVF and improves pump performance by increasing gas density. It is probable that the range of application is narrow, but where boosting is required down-hole at low, but significant, GVF then modified centrifugal pumping can be considered. Texaco are proposing to use hydraulically driven centrifugal pumps in the horizontal wells of the Captain Field (low GVF, heavy oil).

7.3.4. Positive Displacement

Twin Screw

Twin screw pumps were among the first to be applied for pumping multiphase mixtures. Their positive displacement action gives them an inherent gas handling capability. The use of purely rotary motion and the absence of valves and sliding components are all advantages compared to other types of positive displacement pump. In addition this type of machine has the highest capacity for a given machine size which is attractive for the flow rates involved in oil production operations. Twin screw pumps can generally handle up to about 95% GVF. Higher GVF can be handled by injecting liquid into the pump so that the liquid content within the pump is maintained to provide a sealant and also provide cooling and lubrication. In many cases GVF is so high that it is more appropriate to regard the required machine as a wet gas compressor. Since the twin screw machine in its normal form does not have internal compression the power consumption is high compared to that of a compressor performing a similar duty. The difference depends upon the GVF and the pressure ratio. Attempts have been made to incorporate internal compression but so far the increased losses have outweighed the benefits.

This type of application has led to several successful prototypes and a number of commercial and semi-commercial applications. This twin screw pump format is poorly suited to subsea application because of the problems of seals and timing gears, see Section 17.4.2. The twin screw pump is best suited to low/medium throughput applications where the pressure ratio or differential pressure is high and GVF less than say 95%. Twin screw pumps in commercial operation are noted in Section 17.4.2.

Internal Twin Screw

This development is a variant of the Mono pump (or Moineau Screw Pump) in which the stator is allowed to rotate, and known as the ‘idler’. In principle this is a twin screw pump in which one screw rotates inside the other. This has the advantage that all components run in bearings which reduces wear by reducing forces between rotor and idler; this allows much higher speeds to be used than in a conventional Mono pump, thereby reducing pump size for a given capacity and allowing the efficient use of hydraulic turbine drive. However during testing it was found that the elastomer of the idler could not stand the duty and further work has been abandoned.

Reciprocating/Linear Motor

A reciprocating pump driven by a linear electric motor has been tested for subsea application. The project is known as Multiphase Electric Pump Station (MEPS). The unit is large and heavy for a relatively low capacity, 1 30m^3/h. This machine is best suited to low flow rate, high pressure ratio, high differential pressure applications. Initial works testing was followed by testing at Texaco’s Humble facility and is said to have been successful.

Diaphragm

The machine is based on the pumping of hydraulic oil by a conventional pump. The oil is alternately pumped into and out of two pumping vessels where it acts on an elastomeric diaphragm to displace the multiphase mixture. Material selection was said to have been satisfactory but problems were experienced in detecting diaphragm position at the end of stroke. These are said to have been overcome and the pump has been on long term test. For subsea application the hydraulic pump would be immersed in the oil reservoir. The volumetric capacity is not expected to be large.

Hydrobooster

This system is based on a conventional centrifugal pump pumping separated liquid sequentially into a number of pumping vessels to alternately draw in and displace the multiphase mixture. Thus each vessel acts as the cylinder of a reciprocating compressor in which the liquid acts as the piston. A full scale prototype is said to have been successfully tested on hydrocarbon gas/liquid mixture.

Double-action Piston

Double acting piston pumps would work but have too low capacity for the application. A large number of pumps would be required to operate in parallel.

Ram Slurry Pump

There is no reason why a slurry pump should perform any better than a conventional piston pump except for its solids handling capability. The use of rams lowers capacity compared to piston pumps. Valves are generally leaky compared to those of conventional pumps.

17.3.5 Other Systems -Jet Pump

The jet pump has the potential of being a simple no-moving-part pressure booster in relatively low flow applications. Because of its very low efficiency it does need a virtually 'free' source of high pressure driving fluid, such as a high pressure well which would otherwise be choked. It has the advantage of simplicity but high fluid velocities render it susceptible to erosion by sand laden fluids; however by careful design this may be minimised and the vulnerable parts made easy to replace. The high pressure motive flow should be single phase, otherwise the already poor efficiency is further degraded. The motive flow can be gas for a gas-pressure boosting but for the multiphase flow pressure boosting the motive fluid must be liquid-only for reasonable performance. In the WELLCOM example it is proposed that the liquid from a high pressure well is separated from the vapour fraction and used to boost the pressure of a low pressure well; otherwise the power consumption is high for a given pumping duty.

17.3.6 Systems with Separation

Strictly speaking these are not 'multiphase' boosting systems but depend on the division of the flow into separate, if relatively impure, phases:

Kvaerner Booster Station (KBS)
The Kvaerner Booster Station is a separation/compressor/pump system based on a compact vertical separator with centrifugal pump below to increase the liquid phase pressure and a wet gas compressor to increase the vapour phase pressure. Although pilot trials have been satisfactorily completed and the potential is for a very flexible system of wide applicability, the complexity of the system does not bode well for its commercial success. Further development is currently (3Q-'95) on-hold.

VASPS
The Vertical Annular Separation and Pumping System (VASPS) is a way round the problem of multiphase pumping by separating the mixture and using conventional liquid pumps to handle the liquid while allowing the gas to free-flow under its remaining pressure. Only the liquid phase is pressure boosted so strictly speaking it is not a 'multiphase pump system'. The separator consists of a vertical length of conductor some 600 mm in diameter. Multiphase fluid is fed into this conductor and forced to follow a helical path. This induces accelerations of 3-4 g. The high acceleration combined with the short distance bubbles must travel through the liquid leads to good separation. The operating principle is thus that of the cyclone separator. Gas rises naturally to the top of the conductor and flows away under its own pressure. Liquid collects at the bottom and is pumped away by a conventional ESP installed at the bottom of a tube within the conductor.

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17.4 'Known' Performance Track Record

Of the many proposed multiphase pumping systems only two have any significant track record of industrial, (semi-) commercial operation, ie. the Helico-Axial and the Twin Screw type pumps. Also the Jet Pump principle has been in use in the WELLCOM system.

17.4.1 Rotodynamic – Helico-axial Pumps

The Poseidon Project was a French (IFP/Total/Statoil) led project to develop remote subsea production systems by identifying gaps in technology and developing the necessary techniques and equipment. Subsea multiphase pumping was always regarded as an essential part of such a system. The project itself was closed out in ca. 1990. A pump had been tested successfully (ca. 3000 running hours) at a land site in Tunisia. A second test in a Norwegian fjord in 1987-88 in 150 m water depth was also successful (ca. 4000 hrs).

Manufacture has been licensed to Sulzer Pumps and to FRAMO and several further successful developments have been carried out:

Pecorade Field (Elf)
A pump (Model P 302) built by Sulzer was installed in the Pecorade Field in the Southwest of France and has been running since June '94, with over 8000 hours by Feburary '96 The pump is used in normal production operations but is also the subject of a test programme so that it experiences a wider range of operating conditions than would otherwise be the case. The maximum throughput has been ca. 365 m^3/h, 55 MBD. There is an active proposal to continue the operation of this pump on a commercial basis.

Gullfaks Field (Statoil)
A pump built by FRAMO was installed on the Gullfaks A platform and started up in May 1994. It had accumulated over 5000 hours running by Mid '95, by which time the programme was considered complete. Operation is said to be satisfactory except that seal leakage is considered a bit high. Seals are manufactured by Burgmann and an in house (Burgmann) test programme is in progress aimed at improving performance.

Draugen Field (BP/Shell/Statoil)
A SMUBS (Shell Multiphase Underwater Boosting System) pump has been installed on the Draugen Field in 270 m water depth. In the SMUBS Project, FRAMO, supported by Shell, have taken the Guinard pump, which is very similar to the Poseidon Pump, and packaged it for subsea installation. Pump internals are built into a tubular casing which can be installed from the surface into a vertical housing without diver intervention. Drive is by hydraulic turbine thus avoiding subsea electrical equipment and the problems of keeping well fluids away from the motor. There are technical and economic distance limits for power transmission. Injection water is being used as power fluid, and this SMUBS pump has now been running periodically since early November '94. All sub-sea operating procedures are said to have been verified and operation is satisfactory. The use of the pump is said to have increased oil production by 39%; however the reasons for its periods of non-operation are not known for certain, although they are believed to be due to platform production limits and the process requirement for maximum water injection pressure.

17.42 Positive Displacement – Twin Screw Pumps

Twin screw pumps were among the first to be applied for pumping multiphase mixtures and several successful trials have been carried out, although the helico-axial pump is now surpassing it in terms of operational track-record.

Bokor B Sarawak
A Multiphase Systems MP 40 size pump (as on Forties Bravo) was tested on the Shell Bokor B Platform, Off-shore Sarawak, at up to 97% GVF. Production operation was regarded as successful, after initial commissioning problems, with availability approximately 96%, and an estimated 16% increase in oil production.

Wehrblec, Germany and Housemartin, Canada
On the Wehrblec Field in Germany a Bornemann Twin-Screw pump has been installed and as the result the field life has been extended by ca. 5 years. On the Housemartin Field in Canada a similar twin-screw pump has been installed to lower the well-head pressure and so avoid hydrate conditions.

Tinmar, Trinidad
On two wells in the Tinmar Field in off-shore Trinidad, operated by Mobil, Leistritz twin screw pumps (model L4) have been installed. This has resulted in the re-opening of one well and the up-rating of the other with a net increase in production and a pay-back time of less than a month.

17.4.2 Jet Pump (WELLCOM)

The jet pump principle is in use as a multiphase pressure boosting system using the high pressure flow from one source, a well or group of wells, to boost the pressure from another, adjacent source, a well or group of wells.

17.5 Outline Design

17.5.1 Required Information

The most important parameters for the selection and sizing of multiphase pumps are:

Suction pressure P,
Delivery Pressure P_2
Total Volumetric Flow at Suction Conditions Q_1
Gas Volume Fraction at Suction Conditions a

Because of its effect upon a and Q_1, P, has a marked influence on the sizing of the pump and the power requirement.

Other important considerations are the location, ie. subsea or surface, the power source and whether abrasive solids are present.

[Note that the term Gas Liquid Ratio (GLR) is often used in connection with multiphase pumps but refers to volumes at suction conditions, as is GVF, rather than the accepted definition of GLR with volumes referred to standard conditions; care must be taken to avoid this confusion. Note also that flow rates for pumps often use the units of BPD, which includes not only the liquid rate but also the gas, both at suction conditions].

Additional factors such as design temperature and pressure, and fluid composition/properties, will be relevant to detail design, pump material selection and motive power unit design. The flow regime of the feed flow may dictate a need for the pre-conditioning of the feed, particularly if sudden changes in GVF or periods of 100% GVF are possible. Furthermore the change in duty, flow rate and suction/discharge conditions during operation may mean that no single pump type, or even one form of pressure boosting, let alone a single pump size, will be ideal throughout field life.

The first stage is to calculate the maximum total flow at the suction conditions, (Q_1), the pressure ratio required (P_2/P_1) and the GVF, at suction conditions. An attempt can now be made to select a suitable pump type from Table 2. However no pump type may be completely appropriate and some accommodation may be necessary between what is ideally required and what is available.

17.5.2 Type Selection

There is as yet insufficient information or operational experience to be definitive about the selection of suitable multiphase pumping systems. The attached table, Table 2, shows the current status of the systems showing some industrial potential, but some of these may fall-by-the-wayside and others be added to the list in the medium-term future.

GVF	Suction Volume	Press. ratio High/Med./Low		Multiphase pump type	Power	Sand ?	'Efficiency'	Sub-sea ?
High 100-94%	V.High >1200m3/h	H/M/L	–	Separation – KBS etc.	E	No*	Good	Yes
	High 1200-750m^3/h	High	>5	Separation – KBS etc.	E	No*	Good	Yes
		Medium	5-2	Helico-axial	E or H	Yes	Moderate	Yes
		Low	<2	Helico-axial	E or H	Yes	Mod.	Yes
	Medium 750-250m^3/h	High	>5	Twin screw(Liq. recycle may be needed)	E	No	Mod.	?
		Medium	5-2	Helico-axial or	E or H	yes	Mod.	Yes
				Twin screw(Liq. recycle may be needed)	E	No	Mod.	?
		Low	<2	Helico-axial or	E or H	Yes	Mod.	Yes
				CRA	E or H	No	Mod.	?
	Low <250m^3/h	High	>5	Piston/MEPS	E	No	Good	Yes
		Medium	5-2	Helico-axial or	E or H	Yes	Mod.	Yes
				Piston/MEPS or	E	No	Good	Yes
				CRA	E or H	No	Mod.	?
		Low	<2	Helico-axial or	E or H	Yes	Mod.	Yes
				CRA	E or H	No	Mod.	?
Medium 94-70%	V.High	H/M/L	–	Separation – KBS etc.	E	No*	Good	Yes
	High	High	<5	Separation – KBS etc.	E	No*	Good	Yes
		M/L	<5	Helico-axial	E or H	Yes	Mod.	Yes
	Medium	High	>5	Twin screw	E	No	Mod.	?
		Medium	5-2	Helico-axial or	E or H	Yes	Mod.	Yes
				Twin screw	E	No	Mod.	?
		Low	<2	Helico-axial	E or H	Yes	Mod.	Yes
	Low	High	>5	Piston/MEPS or	E	No	Good	Yes
				Twin screw	E	No	Mod.	?
		Medium	5-2	Helico-axial or	E or H	Yes	Mod.	Yes
				Piston/MEPS	E	No	Good	Yes
		Low	<2	Jet or	H	No	Poor	Yes
				Helico-axial	E or H	Yes	Mod.	Yes
Low 70-50%	V.High	H/M/L	–	Separation – KBS etc.	E	No*	Good	Yes
	High	High	>5	Separation – KBS etc.	E	No'	Good	Yes
		M/L	<5	Helico-axial	E or H	Yes	Mod.	Yes
	Medium	High	>5	Twin screw	E	No	Mod.	?
		Medium	5-2	Helico-axial or	E or H	Yes	Mod.	Yes
				Twin screw	E	No	Mod.	?
		Low	<2	Helico-axial	E or H	Yes	Mod.	Yes
	Low	High	>5	Twin screw or	E	No	Mod.	?
				Piston/MEPS	E	No	Good	Yes
		Medium	5-2	Twin screw or	E	No	Mod.	?
				Helico-axial	E or H	Yes	Mod.	Yes
		Low	<2	Jet or	H	No	Poor	Yes
				Helico-axial	E or H	Yes	Mod.	Yes
V.low <50%	V.High	H/M/L	–	Separation – KBS etc.	E	No*	Good	Yes
	High	High	>5	Separation – KBS etc.	E	No*	Good	Yes
		M/L	<5	Modified (multistage) centrifugal pump	E or H	Yes	Mod.	Yes
	M/L	High	<5	Twin screw	E	No	Mod.	?
		Medium	5-2	Modified (multistage) centrifugal pump	E or H	Yes	Mod.	Yes
		Low	<2	Jet or	H	No	Poor	Yes
				Modified (multistage) centrifugal pump	E or H	Yes	Mod.	Yes

Notes: Sand — * Small amounts may be acceptable, but disposable may be problematic
'Efficiency' — This is a reflection of the apparent energy efficiency to transport oil; it does not imply anything about the thermodynamic efficiency, see 17.1
Power — E = Electrical H = Hydraulic

Table 2. Multiphase pump selection chart

17.5.3 'Feasibility' Design

The feasibility of using a rotodynamic pump can be checked as follows:

1 Calculate maximum total flow at suction conditions. Figure. 5 shows the range of pump sizes available. The maximum size for a single pump is presently quoted (by Sulzer Pumps) as about 1 200m^3/h, (180 MBD = MPP6. in Figure 5.); however a limit of 500m^3/h (75 MBD = MPP3/4) is shown in Table1 . Above this flow multiple pumps in parallel are presently recommended

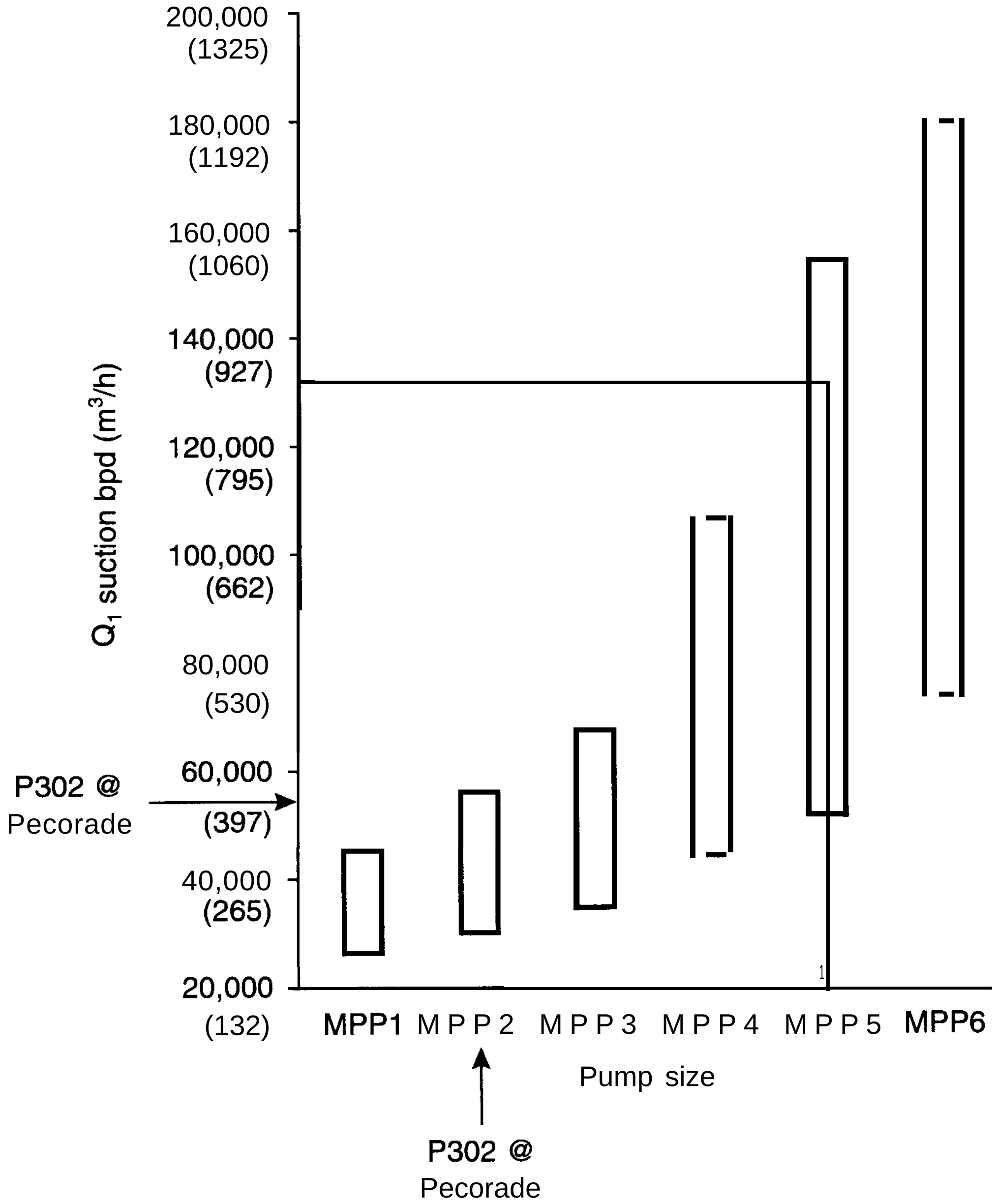


Figure 5. Maximum volumetric flowrate at suction

2 Calculate the pressure ratio required (P_2/P_1) and verify from Fig. 6 that it is achievable.

P_1 = absolute suction pressure
P_2 = discharge pressure

$$\Delta P = P_1 \left(\frac{P_1}{P_2} - 1 \right)$$

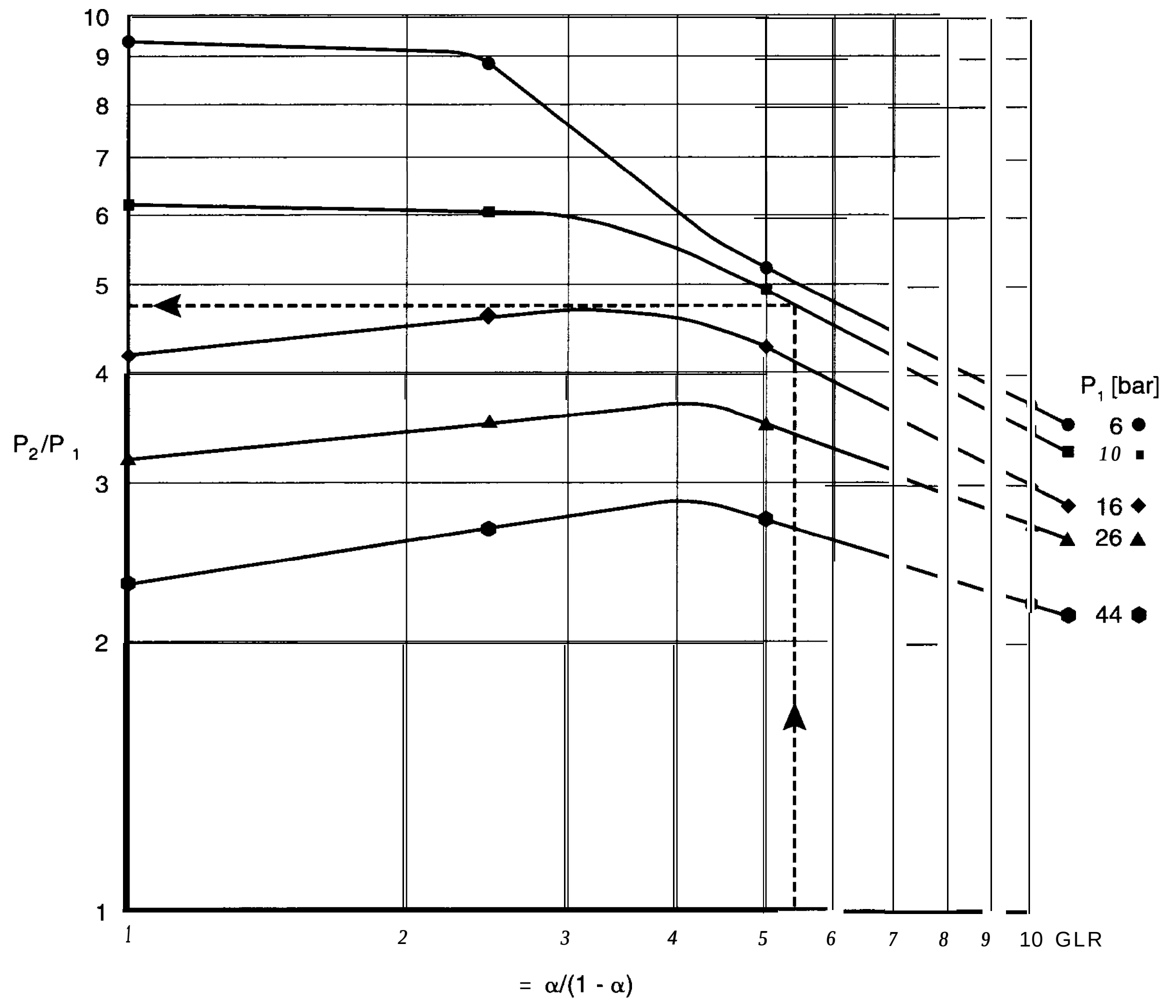


Figure 6. Maximum achievable pressure ratio P_1/P_2

3 Assuming that these steps have shown the use of a rotodynamic pump to be feasible the power requirement can be estimated from Figure 7.

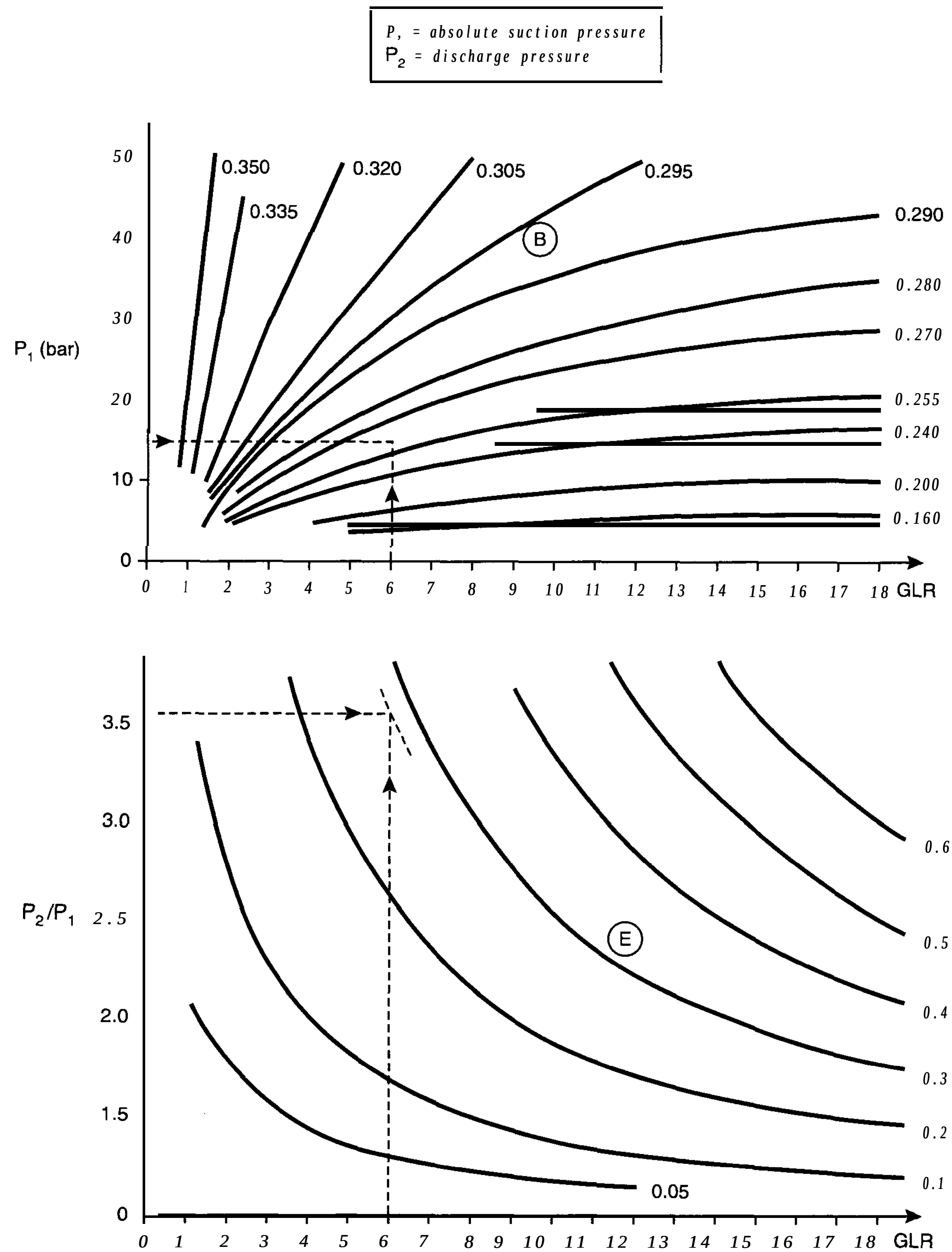


Figure 7. Power coefficients

Read:

B = f(P_1,GLR) from the upper graph
E = f(P_2 / P, ,GLR) from the lower graph

$$\text{Power} = \frac{E \times Q_L \times P_1}{B} \text{ kw}$$

Note in this equation:

Q_L = Liquid flow rate, m^3/h ($Q_L = Q_1$ (1 -a))
P, = Suction Pressure, bar
GLR = Gas/Liquid Ratio = al (1-a)

Note that published data appears to show a marked improvement in the performance of roto-dynamic pumps as suction pressure increases. This is because such data is normally presented as graphs of Differential Pressure vs. Flow Rate. As the suction pressure increases the differential pressure can increase while keeping the head constant.

The feasibility of using a twin screw pump can be checked as follows:

1 Calculate the total suction flow rate.
Twin screw pumps are obtainable up to about 2000m^3/h, (300,000 bpd)

2 Check that the pressure requirements are within the capability of the pump being considered, ie. pressure rating of the casing, seals etc.

3 Estimate the power required from

$$\text{Power} = \frac{(P_2 - P_1) \times Q_1}{36 \times 0.6} \text{ kW}$$

Note:

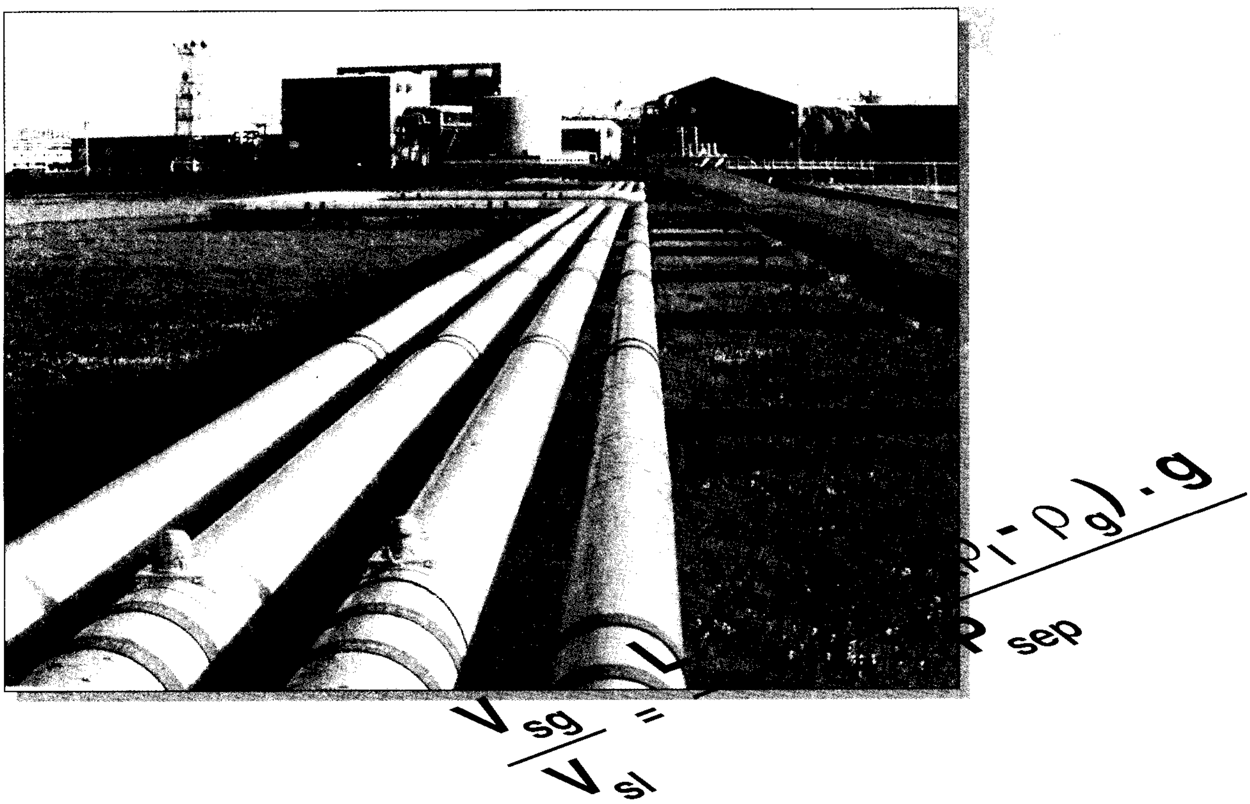
1 In this equation pressures are measured in bars and flow rate in m^3/hr.

2 The 0.6 factor assumes an efficiency of 60%

3 Twin screw pump vendors' machines vary widely in size and pressure rating.
It would be best to check possible applications with one or more vendors.

It is possible that the power transmission limitations may restrict the available pump performance and will almost certainly affect the pressure boosting economics.
It is necessary to consult electrical and control engineers on this aspect.

Section 18 Slug-catcher Design



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18.1 Introduction

The transportation of hydrocarbon fluids down a pipeline of any significant length often results in separation of the vapour and liquid phases such that the fluids travel along the pipeline in partially separated packages, known as 'slugs', which are predominantly liquid, and 'bubbles', which are predominantly vapour. This is usually known as 'slug flow', and generally occurs in lines with a high liquid content. Alternatively with a predominately vapour flow, liquids may separate and settle in the low points in the pipeline; these pools of liquid may subsequently be projected forward as slugs. This can happen either during an increase in vapour flow rate, known as 'bean-up' slugs, or when the line is pigged or sphered for any reason, often to prevent the excessive build-up of liquids in the line, thus producing 'pigged' slugs.

In normal slug flow (see Section 3.3 and 3.4) regular or random slugs occur virtually continuously and although the size of an individual slug is not predictable the statistical size distribution can be calculated. In the vapour flow-lines the sizes of the pigged slugs, and to a lesser extent the bean-up slugs, can be calculated; in particular the size of the pigged slug may determine the frequency of the pigging operation, see Section 4.4. Slugs produced by transient effects are covered in Section 7.

The phenomenon of slugging has both mechanical and process implications. Mechanically the weight and momentum of the liquid slug will put additional strains on the pipe work support structures, in particular because of the rapid changes in forces when the slug/bubble boundary passes. The process problems are caused by the rapid changes in duty/rate required by the down stream processing facilities. The solution to these problems is a slug-catcher; this is large vessel placed at the end of a multiphase line to enhance the vapour/liquid separation and to catch and store the liquid slug so that it can be passed forward for processing at a more steady rate.

Even if a multiphase pipeline running under its normal operating conditions does not suffer slug flow, it should be remembered that start-up, up-set, low-flow or shut-down conditions may result in the productions of slugs, which need to be addressed at the design stage.

Note that, correctly designed and operated, a slug-catcher system will yield a nearly steady forward feed of liquid but that the vapour phase will normally still be relatively unsteady. However it is possible to design the vessel with 'over-pressure' capacity to catch the gas bubbles, see Section 8.5.

Almost as important as the correct sizing of a slug-catcher is the operational control of the system; this is described in Sections 18.5 and 18.6.

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18.2 Selection and Design

Once the occurrence of slug flow, or other unsteady flow phenomena requiring the use of a slug-catcher, has been identified the following procedure should be followed:

Step 1
Estimate/Calculate Volume of the Slug, ie. the liquid volume of the slug; it can be assumed that the slug-catcher will carry out a basic vapour/liquid separation and although the vapour may contain small amounts of liquid (say up to 10 USgal/MMSCFD or 0.001 kg liquid per kg vapour), it can be assumed that the liquid product is essentially gas-free, ie. contains no free vapour phase. It is initially assumed that no liquid/liquid (oil/water) separation takes place within the slug-catcher. The vapour stream can generally be sent forward to a scrubber for the removal of the remaining liquid before further processing, eg. dehydration, dew-point control and/or compression.

Step 2
Decide on the type of Slug-Catching Vessel required; this may either be a **pipe-finger** or a vessel type of slug-catcher. See the attached Decision Tree Chart, Figure 1.

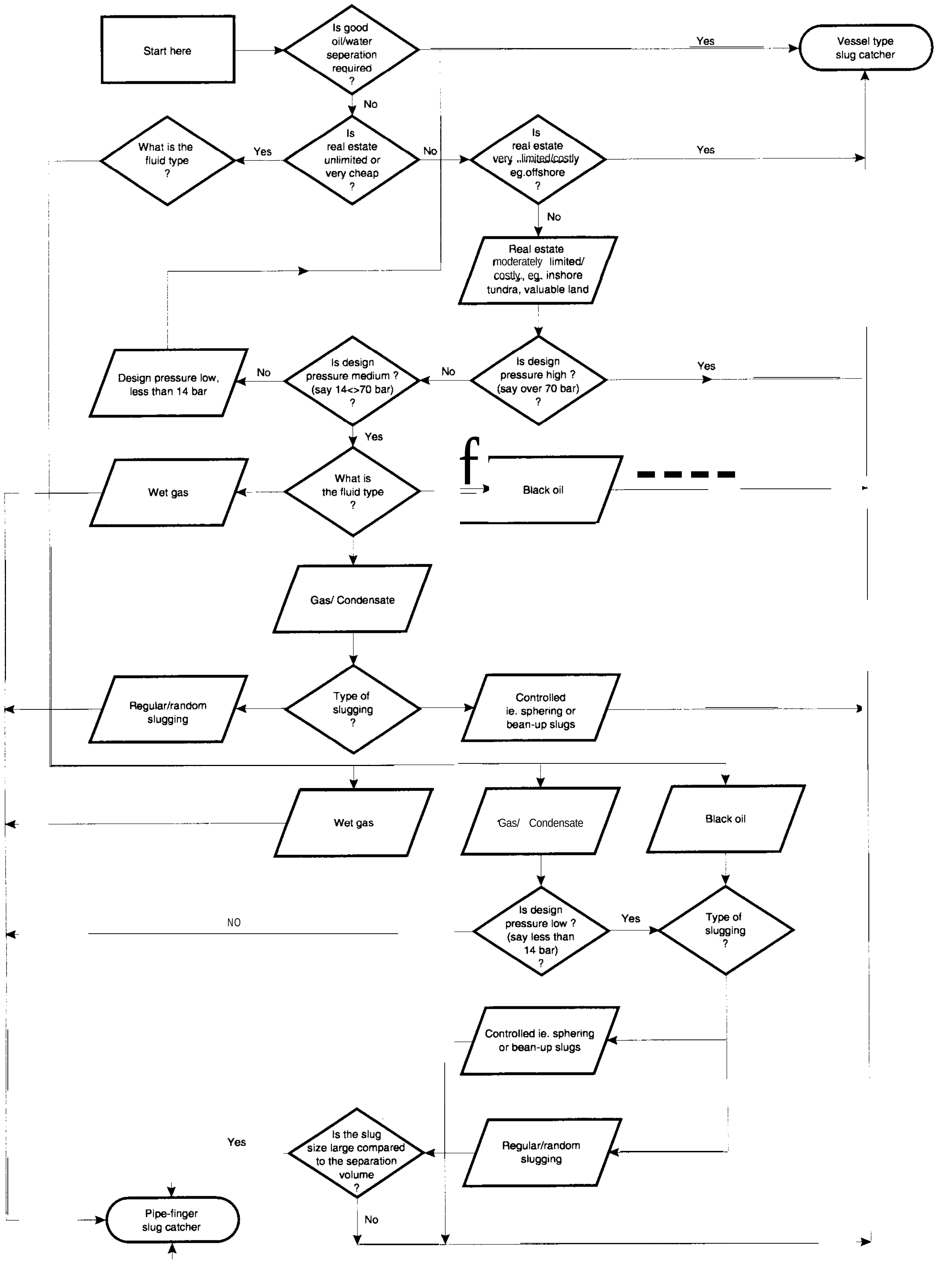

Figure 1 Decision Tree Chart

Step 3

The total liquid retention volume of the vessel has next to be calculated by adding to the liquid slug size calculated (Step 1) additional volumes for control and operational safety: These will vary depending on the form of the down-stream processing of the liquid phases; the processing may either be continuous, at as near as possible to a steady rate, matching the pipeline average liquid rate, or it may be batch-wise, with periods of no liquid processing between the slug arrivals. The former would be typical of a black-oil pipeline; the later typical of a wet-gas pipeline.

In a slug-catcher the slug size determines the volume between the High Liquid Level (HLL) and the Low Liquid Level (LLL). The HLL is the level normally used as the design liquid level for the determination of the vapour handling capacity, defining the normal feed to the down-stream vapour handling equipment, and is the level at which an alarm will sound to warn the operator of the high level for him to take action as appropriate, but no automatic executive action is taken. However a High-High Liquid Level (HHLL), a level at which executive action is taken to protect down-stream equipment, needs to be determined. At this HHLL the down-stream equipment may not function to specification but the operation should not be hazardous. The volume between HLL and HHLL should be the maximum liquid flow multiplied by the time required to start-up the batch processing plus, say, one minute safety margin, or the time to close the feed flow to the slug-catcher in a safe manner, whichever is the greater. If the processing is continuous, then it should be the maximum liquid flow rate multiplied by the time required to close the feed flow to the slug-catcher in a safe manner. The volume between the LLL, at which an alarm sounds but no executive action is taken, and the bottom of the vessel is composed of three parts: the ullage volume, the volume from there to the Low-Low Liquid Level (LLLL), where executive action is taken to prevent gas breakthrough, and LLLL to LLL. The ullage volume is to choice but is a volume to hold sand, solids etc. and is assumed to take no part in the retention or processing of the liquid phase(s). The LLLL to LLL volume depends on the form of the liquid down-stream processing; if the processing is batch-wise then it should be the maximum liquid processing rate multiplied by the time taken to shut-down that processing in a safe manner; if the processing is continuous then it should be the volume associated with the time taken to turn-down, in a safe manner, from the maximum to the minimum liquid processing rate, or it should be the maximum liquid processing rate multiplied by the time taken to shut-down that processing in a safe manner, which ever is the greater. The total liquid holding volume is now known.

Step 4

The first pass at designing the slug-catcher can now be made. The first pass assumption is that the vapour nominal velocity is 1 m/sec, in the free part of the vessel, initially assumed to be 50% of the cross-sectional area. The relevant vapour flow rate is the maximum flow, ie. the highest vapour velocity in the bubble flow in between the slugs and at the minimum operating pressure.

Notes on the Decision Tree Chart:

Oil/Water Separation

Do you require immediate oil/water separation within the slug-catcher vessel ? Note that this is generally to be avoided in that good oil/water separation requires the maintenance of steady levels whereas the essence of slug-catching is that the level(s) are inevitably varying.

Real Estate

Is the land or platform upon which the slug-catcher is to be installed of high value and **very limited** in extent, eg. on an offshore platform, or is it of value and/or somewhat **limited eg.** in shallow water or an onshore location with significant land values, or of more-or-less **unlimited** extent, eg. an onshore location with low land values.

Fluid Type

What is the general nature of the fluid ? A **Wet Gas,** with low liquid content in a predominantly vapour flow; this will generally only produce liquid slugs during pigging or sphering, but may also do so during increases in flow rate ('bean-up' slugs). A **Gas-Condensate,** or very high GOR oil flow, will generally produce pigging or bean-up slugs and also irregular terrain induced slugs. A Black Oil with medium or low GOR is likely to produce regular/random slugging at all times.

Design Pressure

What is the design pressure of the vessel ? Pipe-finger type slug-catchers can be made relatively economically (and can be site-built) for high pressure application. **High** pressure applications are say above 70 bar, **Medium** pressure between 70 and 14 bar and **Low** pressure application below say 14 bar. Note that any requirement to damp out the vapour phase flow rate variations may dictate a higher operating and design pressure than might otherwise be required.

Type of Slugging:

If the slugs are only produced by **pigging** or flow increase **(bean-up)** their arrival can be anticipated and their size can be to some extent controlled, either by pigging frequency changes or limiting the size of flow rate changes. Thus the size of the slug can be limited by operational constraints and the slug-catcher size reduced. If the slugging is produced **randomly** by a slug flow regime then the size distribution can be predicted but the size of an individual slug can not be controlled within that distribution.

In some border-line cases it may be necessary to carry out a preliminary sizing of both the vessel and pipe-finger type slug-catchers to see which will be more suitable.

Note: If it is necessary to effect oil/water separation in the slug-catcher then this should be referred to XFE. XTP. Sunbury.

18.3 Vessel Type

If a vessel-type slug-catcher has been selected then, from the flow area calculated above, the diameter of the vessel is given, initially assumed to be 50% full and assuming a length/diameter ratio of say 5:1. The liquid volume can be then calculated. If the liquid volume is inadequate for the known slug volume and the control functions as outlined in Step 3, then either the design level can be increased to a maximum of 65% and/or the vessel lengthened up to maximum of say 10:1 . If the liquid volume still dominates the design consideration should be given to limiting the slug size or increasing the liquid processing rate; otherwise the vessel should be designed only on the basis of the liquid holding capacity. Conversely if the liquid volume available is greater than needed the design liquid levels may be lowered to say 30% but it must be remembered that the differences between the various levels must be a minimum of 1 00mm in order that the level instruments can function adequately; this may dictate a minimum liquid depth. The vessel length/diameter ratio can be reduced to a minimum of 4:1, if the liquid depth is limiting.

The vessel design for slug-catching should be as simple and robust as possible in view of the hammering which internals are likely to suffer as a result of the liquid loads and slugs at high velocities. The vapour/liquid separation performance can be checked by using the BP software PROSEPARATOR. Note that this should be used on the basis of a vessel with no internals, the liquid at a suitable elevated level, and the maximum vapour velocity.

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18.4 Pipe-finger Type

If a pipe-finger slug-catcher has been selected the number of standard large diameter tubes to handle the vapour flow, as described above, is calculated eg. 36", 42" and 48" tubes, (larger sizes may be used, but they are not generally readily available). A smaller number of large bore tubes is generally the more economic solution. The number of tubes is then doubled (to allow for the tubes being 50% full). However the result must be a power of 2, ie. 2, 4, 8, etc. so that appropriate dead-end tees may be used to sub-divide the flow as uniformly as possible. See Section 11 for more information on dividing multiphase flows. A typical Pipe-finger Slug-catcher is shown in Figure 2.

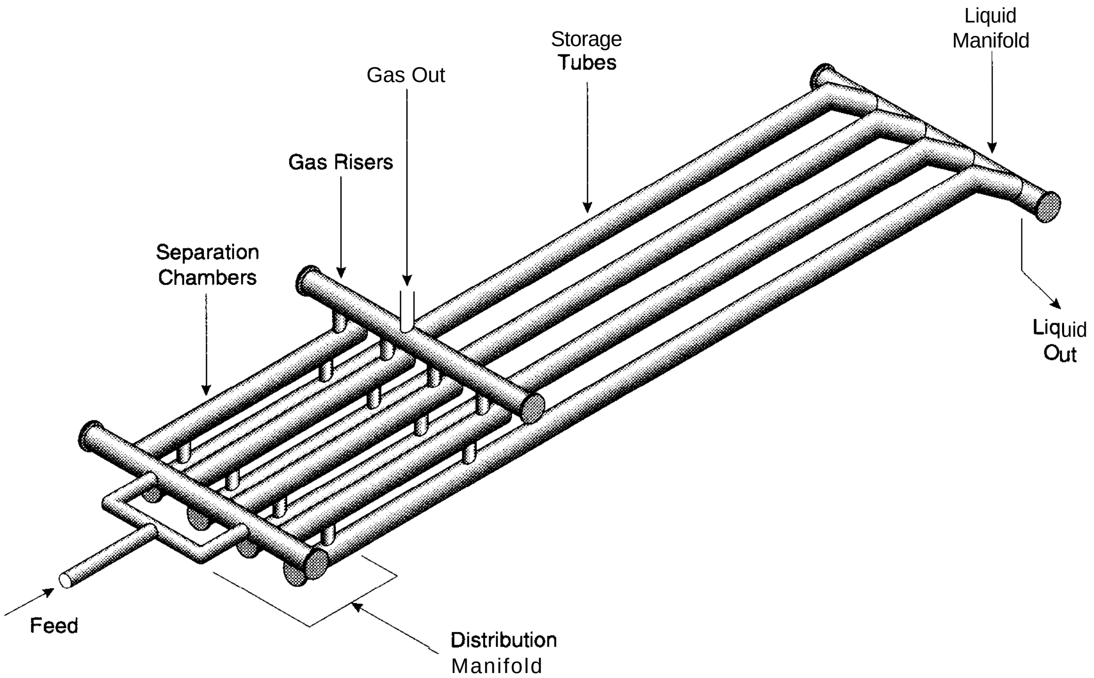

Figure 2 Pipe-finger Type Slug-catcher

The initial length of the tubes should be taken as a minimum of 15 times the diameter. If the result is a reasonable number then an initial design can be produced by linking the ends of the tubes with a transverse tube, usually of the same pipe-size, to act as a collecting vessel; the volume of this is ignored for the first pass. The lower half of the main tubes is used for the liquid storage but if this is not adequate then additional tubing length may be added either to the separation tubes or as an extension to the transverse tube. If drainage into lower tubes is required to hold the liquid volume then large bore, free drainage facilities into the transverse collection tube and the storage tubes must be provided.

If the number of tubes (calculated for the vapour handling duty) of the maximum size available is very large, ie. more than say 8, then the assumed design liquid level in the tubes can be reduced to say 25% of the cross-sectional area, and the number of tubes re-calculated. Full bore drainage from the main tubes to the liquid storage tubes must then be provided to drain the liquid slug from the vapour separation tubes as quickly as possible.

The design of the vapour off-take pipes should be large enough to give a low initial vertical velocity to maximise droplet separation. The preferable size is an equal Tee, ie. the same size as the separation / storage tubs, but should be no smaller than 0.7D, ie. area equal to the flow area of the separation / storage tubes at 50% full.

Slug-catcher pipe-finger design should be kept relatively simple. One of the advantages of the pipe-finger type is that it is classified as 'pipe-work' for the purposes of design and inspection, with less onerous requirements; however this may be lost and the system re-classified as a vessel if the design comes too complex, particularly if internals are added.

A possible hybrid design can be used with a 'vessel' designed for the vapour/liquid separation and pipe-work as the storage medium. Although more compact than the pipe-finger type and perhaps with a higher, more predictable separation performance than the pure pipe-finger type, it does not enjoy the full advantages of either the pipe-finger or vessel types. See Figure 3.

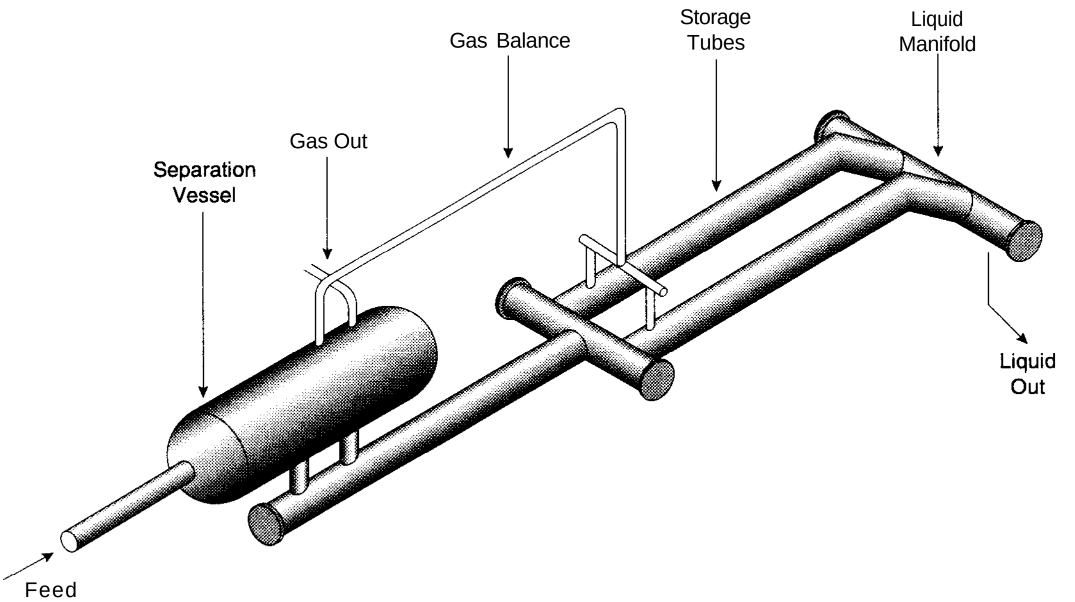

Figure 3 Vessel Type Slug-catcher

Where a slug-catcher is fed from a manifold or collection header, gathering the well-fluids from a number of wells and/or flow-lines, then care should be taken to arrange the pipe work such that the individual flow-lines are hydraulically disconnected from each other and from liquid surges in the slug-catcher. This can be achieved by connecting the flow lines into the top of the header so that there is no direct liquid communication between the individual flow-lines and the slug-catcher.

18.5 Bubble-catching

When slugging flow occurs the flow consists of trains of liquid slugs and vapour bubbles. While the slugs of liquid can be caught by level variation in the slug-catcher vessel there is no such volumetric 'catch' of the vapour phase bubble. The best that can be achieved is damping of the vapour flow by pressure variation in the vessel, but the required additional pressure rating of the vessel can be expensive to provide.

In its simplest form smoothing the forward flow of vapour can be achieved by flow control of the vapour stream, rather than using the vapour release rate to determine the pressure within the slug-catcher. However there will be an upper limit to the pressure variation allowed within the slug-catcher according to its design, so that the flow controller set point must be re-set or over-ridden by the measured pressure.

Allowing the pressure in the slug-catcher has implictions on the other parts of the system which may be undesirable: a higher pressure will cause an increase in feed flow back-pressure and might ultimately cause production restrictions. Changing the pressure has an effect on the quality (RVP, Bubble Point, etc.) of the liquid phase, which in some cases may cause other processing difficulties. The extent to which bubble-catching by pressure variation might be useful or detrimental can best be determined by carrying out a dynamic analysis on the slug-catcher operation, see below.

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18.6 Operation and Control

The control of a slug-catcher is fundamentally very simple but in practice rather more complex due to unsteady nature of the operating parameters.

The operational principle is to carry out a rudimentary vapour/liquid separation and to accumulate the liquid within the vessel to allow a steady forward flow of the liquid phase to the down-stream process. This means that the liquid exit flow should be by flow control, rather than level control; this flow control will need to be set at the time-average liquid make, which can only be estimated from the historical records and or the choke positions/well test data. The liquid level will need to be monitored to avoid over-filling (carryover) or emptying (gas blow-by) and may be used as re-set value for the flow control if, but only if, the level moves outside the acceptable range. A slug-catcher does not by its very nature have a level control system – the level is variable or it is not acting as a slug-catcher. A historical plot of liquid level and forward flow against time, possibly with a computed figure of inflow, calculated from choke positions and well-test data (in the same units and at the same conditions as the measured flow), would be useful aid to operator control.

The pressure in the slug-catcher can be controlled by regulation of the vapour off-take rate, in the same way as a standard V/L separator. However if the slug-catcher is also required to smooth out the vapour flow this will not be adequate. The vapour exit flow should then also be flow controlled, reset by the vessel pressure.

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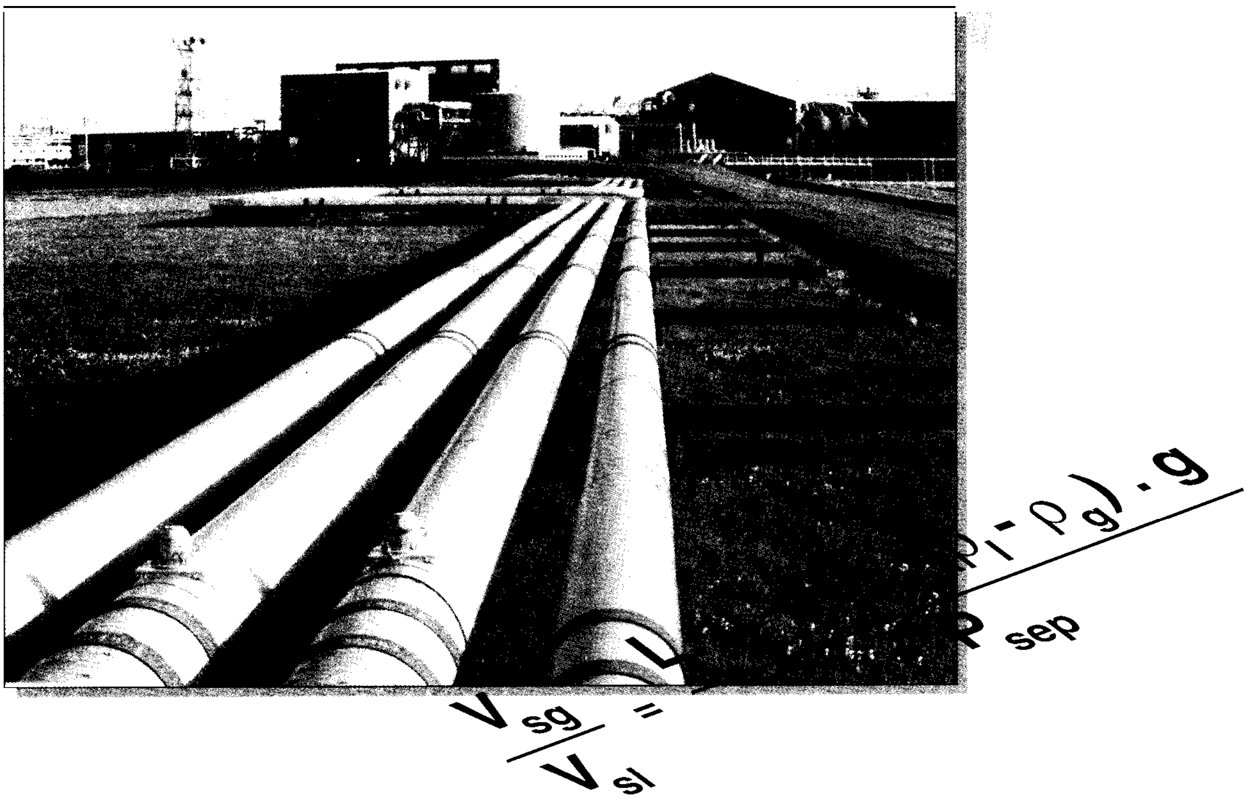
Section 19 Sandwashing

Design of Vessel De-sanding Systems



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19.1 Introduction

Solids are often co-produced with hydrocarbon fluids, such as sand, from unconsolidated sand reservoirs such as Forties, and Proppant, which is injected into the formation in the Southern North Sea Gas Fields to keep open fissures. The solids are often associated with increased water production. This section describes a design method for a de-sanding system for horizontal three phase separation vessels while on-line, i.e. without emptying the vessel or even stopping oil production.

It is based on work carried out at Sheffield University, Department of Mechanical & Process Engineering, and Flow Systems Design Ltd. (J R Tippetts & G H Priestman), and the BP software subsequently written based on that technology. The concept is to inject wash-water to a create a given momentum flux through a number of flat fan-jets carefully distributed through the lower part of the vessel, subdivided as necessary, followed by gravity drainage through a small number of sand-wash drains.

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19.2 Method Outline

Basic Information

Before the design of a de-sanding system can be attempted certain basic information is required concerning the vessel to be sand-washed, the solids to be removed and the water available to carry out the process. These data are listed in Table 1. The sand-wash method is based on the use of high pressure water injected through 45” fan flat jet nozzles (“jets”) of the type supplied by Lechler Ltd. of Sheffield, or equivalent, see Table 2.

Sand-Wash Sections and Tiers

Unless the vessel is small i.e. less than ca. 2m diameter with an L/D ratio of less than 3:1, then the vessel will need to be subdivided into sand-wash sections and will need more than one tier of sand-wash jets each side. The general arrangement is usually dictated by the design/size of the vessel, particularly the presence of weirs or baffles in the liquid part of the vessel, although in some cases it may be advisable to reconsider the choices first made in the light of later results.

FF Equation

The key tool in the design of the sand-wash system is the “Fluidisation Factor” or FF equation. Having determined the length of the sand-wash section and the number of tiers, a preliminary figure for the jet spacing and number of jets can be determined. Using the FF Equation, a jet size is selected which determines the jet pressure drop required to give a Fluidisation Factor (FF). The jet size and number and the pressure drop give the total wash-water flow. By selecting different jet sizes and numbers/spacing, the balance between jet pressure and water flow rate can by changed as required. However, if there is a fundamental mismatch it may be necessary to change the number of tiers or sand-wash section length.

Ancillary Calculations

Having determined the wash-water flow rate for a section, the header, wash-pipes and the sand drain can be calculated. If the quantity of sand in the feed fluids is known then an estimate of the required interval of de-sanding can be made. This may influence decisions about the degree of automation in the control system.

The above functions are partially automated in an Excel Program SANDWASH.XLS,

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19.3 Design Steps

Wash Sections

The length of a wash section, that is the length of the vessel serviced by a set of wash pipes and a single drain nozzle, can be up to three times the vessel diameter; this assumes that the required volume and pressure of wash-water are available, that the drain can be placed in the centre of the section and that the floor of the vessel is un-obstructed. One and a half vessel diameters is the maximum distance of the end of a wash-section from the drain, hence the maximum wash section length of three diameters.

This generally will mean that no more than two or possibly three sand-wash sections, each with wash-pipes and drain nozzle, are required in a standard horizontal separator, unless process internals dictate otherwise, or the wash-water feed rate or sand-wash drainage rate is very restricted. The existence of weirs or oil/water baffles for process reasons may already divide the vessel into sections and it is generally advisable to make use of these sub-divisions.

The end of each section, unless bounded by an existing feature such as the vessel end or process baffle, is bounded by a solid segmental baffle, or sand-dam, in the bottom of the vessel, about 0.085D high. The purpose of the sand-dam is to restrict the flow of sand out of the section being de-sanded and so settling in a neighbouring in-active section. The height of this sand-dam is not critical and a lower baffle may be necessary for other process reasons, such as oil/water interface level, but a minimum height of 150mm is recommended, (see Fig.1 (a)).

The number of wash-sections is determined by inspection; the longest section is the most critical and should be designed first. It is this section which will need the largest water flow and/or jetting pressure and will thus determine the size of the wash-water supply pumps and the effluent handling system.

Tiers of Wash-Pipes

Depending on the internal diameter of the vessel, more than one tier of wash-pipes and jets may be needed. Unsupported sand will settle in the vessel on internals and on the walls up to the point where the angle of the wall is the angle of repose, (see Fig.1 (b)). It is assumed that the initial depth of sand is not excessive, say less than 10% of the vessel volume. If the angle of repose is not known then the default figure of 34deg. should be used; this is typical of sand-like solids.

The wash-pipes are fitted with the flat fan jets and should be positioned so that they sweep the bottom of the vessel from the 34” line along the side of the vessel down into the centre where the drain nozzle is located. The upper tier of jets should be on the longitudinal line at 34”. The lateral “throw” of a jet is 1.67 times the longitudinal spacing; the maximum longitudinal spacing recommended is 450mm so that the maximum lateral separation is 750mm. This correspond to a maximum vessel size for a single tier of jets of 2.5m. If the jet spacing is 300mm the largest diameter vessel which can be de-sanded by one tier is 1.67m, for two tiers the maximum diameter is 4.5m. or 3.4m for 300mm jet spacing. If a large number of jets is required, giving a very close jet spacing along the wash-pipes, then it may be advisable to increase the number of tiers above the minimum; the aim is to obtain a fairly even pattern of sand-wash jets.

Wash-Jet Nozzles

The sets of jets on either side of the centre line should be identical; therefore the number of jets in each section must be divisible by two. The jets should be at an even spacing along the wash-pipe running the length of the sand-wash section, with half spacing at either end, (see Fig.3). If the numbers of jets in different tiers are not identical, i.e. because the number of jets in the section is not divisible by four, (or six in the case of three tiers), then the smaller number should be in the upper tier, with full spacing between the jet and the end of the section, (see Fig.4). The maximum practical jet spacing is ca. 450mm, otherwise significant piles of sand will be left between the jets. There is no minimum spacing but if the jets are closer than say 150mm, larger jets, more widely spaced, or an additional wash-pipe tier might give a better spatial arrangement of jets, and thus a more even momentum flux.

The 45" flat fan jets should be positioned so that the plane of the fan is parallel with the axis of the vessel and at 45" dip to the horizontal; the jets should be positioned 0.028D away from the vessel wall, (see Fig.1 (b)). This positioning is to avoid direct impingement and abrasion of the vessel wall or any lining. The fixing of the jets is critical -they must not be allowed to rotate so that the fan impinges on the vessel wall, nor allowed to fall out, thus leaving a very large hole in the wash-water pipe with a disastrous effect on the de-sanding efficiency and probable rapid and severe damage to the vessel walls.

The size of the jets is required in terms of the Effective Diameter, deff, i.e. that diameter which will gives the pressure drop/flow relationship with a discharge coefficient of unity. Lechler quote the sizes in terms of the Equivalent Bore (EB); a list of EB v. deff for the relevant Lechler Jets is given in Table 2. The size of the jets selected should give the required FF considering the wash-water pressure and flow rate available, (see below). They should also be large enough to resist blockage by solid contamination in the wash-water and small enough not to suffer blockage by the ingress into the jets when submerged in the sand. A size (EB) range of 4mm to 12mm is recommended.

F-Factor, Wash-Water Flow Rate and Pressure

The fundamental parameter for successful de-sanding is momentum flux; this is described by the Fluidisation Factor "F" in the F equation. For sand a FF value of 0.001 is required; for proppant, preliminary indications are that a value of 0.0025 is required. It is probable that the value of F is proportional to the size of the particles, but until this is confirmed an experimental value of F must be determined for solids differing significantly from sand.

The value of FF assumes a depth of sand which is much larger than would normally be expected, i.e. up to 10% of the vessel volume, (a depth of 15% D, see Fig.1 (a)).

Slightly lower values of F than those recommended will result in rather slower sand-wash rates, but significantly lower values will give incomplete solids removal, whatever the duration of the sand-washing. Values of FF greater than the recommended value will merely waste energy and will not result in faster or more effective de-sanding, but will of course do little harm.

The FF equation is as follows:

FF = (2.Pj.n.deff*2) / (Ls.D*2.(rhos-rhol).g)

where:

FF = Fluidisation Factor
Pj = Pressure Drop across Jets
n = Number of Nozzles in Section
deff = Effective Diameter of Jets
Ls = Length of Section
D = Internal Diameter of Vessel
rhos = Density of Solids
rhol = Density of Wash Liquid
Q = Wash-Water Flow
Po = Operating Pressure
Pd = Wash-Water Delivery Pressure

A preliminary number of nozzles suitable for the sand-wash section has already been determined from the geometry of the vessel. Entering the value of FF, the number of nozzles, and a first guess at the jet effective diameter in the FF equation gives the required pressure drop across the nozzles; this also defines the wash-water flow rate (Q) by virtue of the known jet size and number:

Q = (π.deff*2/4).n.(2.Pj/(rhos-rhol))*0.5

To obtain the necessary wash-water delivery pressure the operating pressure within the vessel plus an allowance for the sand-wash pipe-work must be added to the calculated jetting pressure drop. The vessel pressure when sand-washing is generally the normal operating pressure, and pressure drop loss might be say 10% of the jetting pressure; any higher and the mal-distribution between the jets becomes significant. Thus:

Pd = Po + 1.1.Pj

The first estimate of the wash-water flow rate (Q) and pressure (Pd) may well not be satisfactory and adjustments can be made to the size and number of jets. It should be remembered that a high wash-water rate also requires equivalent disposal facilities to treat the effluent from the vessel, whereas higher jetting pressures only require a higher energy in the wash-water feed system. The various courses of action are suggested below:

P & Q slightly too high:
Reduce FT – this is only acceptable for a small reduction of FF and/or if the sand loading is light and/or the sand is fine, as in a second stage separator.

P & Q too high:
Reduce the size (length) of the wash section.

P & Q too low:
Increase the size (length) of the wash section.

P too high and/or Q too low:
Increase size and/or number of jets, check the space distribution.

P too low and/or Q too high:
Reduce size and/or number of jets, check the space distribution.

Having determined the maximum jetting pressure and wash-water flow rate for this section of the vessel, the wash-water delivery system, pump(s), supply pipe-work and internal headers can be calculated. The sand drains can also be sized, (see below). It should be remembered that the sand-wash-water has to be disposed of, and the size of the system is determined by the greatest flow. It is therefore advantageous to equalise the wash-water flows between the various sections as far as possible.

Drain Nozzles

The number of sand drain nozzles is given by the number of sand-wash sections. The nozzle should be placed as near as possible to the centre of each sand-wash section but no further than 1.5D from the end of the section.

The nozzles should be a minimum of 3"NB to avoid blockage by stones etc. which are found in some formations and should be fitted with sand caps; the design of these are shown in Fig. 2.

The size of the drain nozzles can be determined once the wash-water flow is known; the exit velocity should not exceed 3m/sec. to limit the erosion. Larger nozzles may be required in some cases if the driving pressure differential is low.

19.4 Control and Monitoring

The de-sanding operation is very simple. The section to be sand-washed is selected and the wash-water is opened: flow and pressure of the wash-water flow are noted, see below. After a couple of minutes the appropriate sand drain is opened and the sand and wash-water drained from the vessel either for a preset time based on experience and expected sand content or, preferably, until a monitor on the effluent flow indicates that the sand content of the effluent flow is low -the wash-water and sand drain are then closed.

Once the sand-wash system has been designed correctly, the amount of control of the system for efficient de-sanding operation is limited to monitoring. Measurements of the delivery pressure and flow rate of the wash-water can be used to monitor the status of system and give warning of jet blockages or wear. This is very important as severe wear or even the loss of jets has a disastrous effect on the de-sanding efficiency and can result in severe damage to the vessel walls.

The flow/pressure relationship for the system should be measured during commissioning and this relationship can be used during subsequent operation to monitor the mechanical health of the system. During commisioning of the system with the pipework and jets in a good mechanical condition a series of measurements of wash-water flow rate against delivery pressure, preferably with the vessel filled and pressurized to standard operating condition, should be taken. With this data a graph of flow rate v. pressure can be drawn for comparison with subsequent operations. A higher flow rate for a given pressure or a lower pressure for a given flow rate indicate either worn jets (gradual change) or even a missing nozzle (sudden change); the reverse symptoms indicate probable nozzle blockage(s). Depending on the severity of the deviation some remedial action may be required, probably involving vessel entry.

A measurement of the solids content of the effluent stream is also highly desirable in that it can be used to determine the end of the useful sand-wash period and thus minimise the sand-wash time and costs and the disturbance to the oil/water interface in the vessel. While there is bulk solids in the vessel to be removed, the momentum of the jets is largely absorbed by the solids and the disturbance to the oil/water interface is minimised and thus potential cross contamination of the streams is minimised. Once the bulk sand has been removed, the concentration in the effluent flow drops markedly to the sand content of the “stirred-up” water. An ICI Tracerco Densitometer fitted on the effluent pipe-work has been shown to give this indication at the experimental facility, but has yet to be proven industrially.

Input Data Required.
Non-essential but useful information is shown {...}

Vessel Dimensions:

Diameter (ID)	= D (m)
Length (T/T)	= L (m)
Position of Internals in Liquid Phase	
{Length of sand-wash sections	= Ls (m)}

Operation:

Operating Pressure	= P (bara)

Sand or similar solid:

{Sand Feed Rate	= S (m^3/sec.)}
Angle of Repose	= a (if known, default =34°)
Density of solid	= rhos (default for sand = 2600 kg/m^3)

Wash Water:

Restrictions (if any) on wash-water flow rate and delivery pressure.	
Density of wash water	= rhol (default = 1050 kg/m^3

Table 1. linput data

Lechler Fiat Fan Jet Nozzles 45"

Lechler Code Nos.	Equivalent Bore "EB" (mm)	Effective Diameter deff (mm)
670643	2.50	2.06
670723	3.00	2.59
670673	3.50	2.91
670803	4.00	3.26
670843	4.50	3.64
670883	5.00	4.12
670923	5.50	4.61
670963	6.00	5.15
670993	6.90	5.64
671043	8.00	6.52
67 1083	9.00	7.28
671123	10.00	8.18
671163	11.00	9.21
671203	12.00	10.30

Table 2. Lechler jet data

References

"Fundamentals and Design of Sand-wash Systems for Separator Vessels, Final Report, December 1990", by J.R. Tippetts & G.H. Priestman, Flow Systems Design Ltd. and Sheffield University, Department of Mechanical & Process Engineering.

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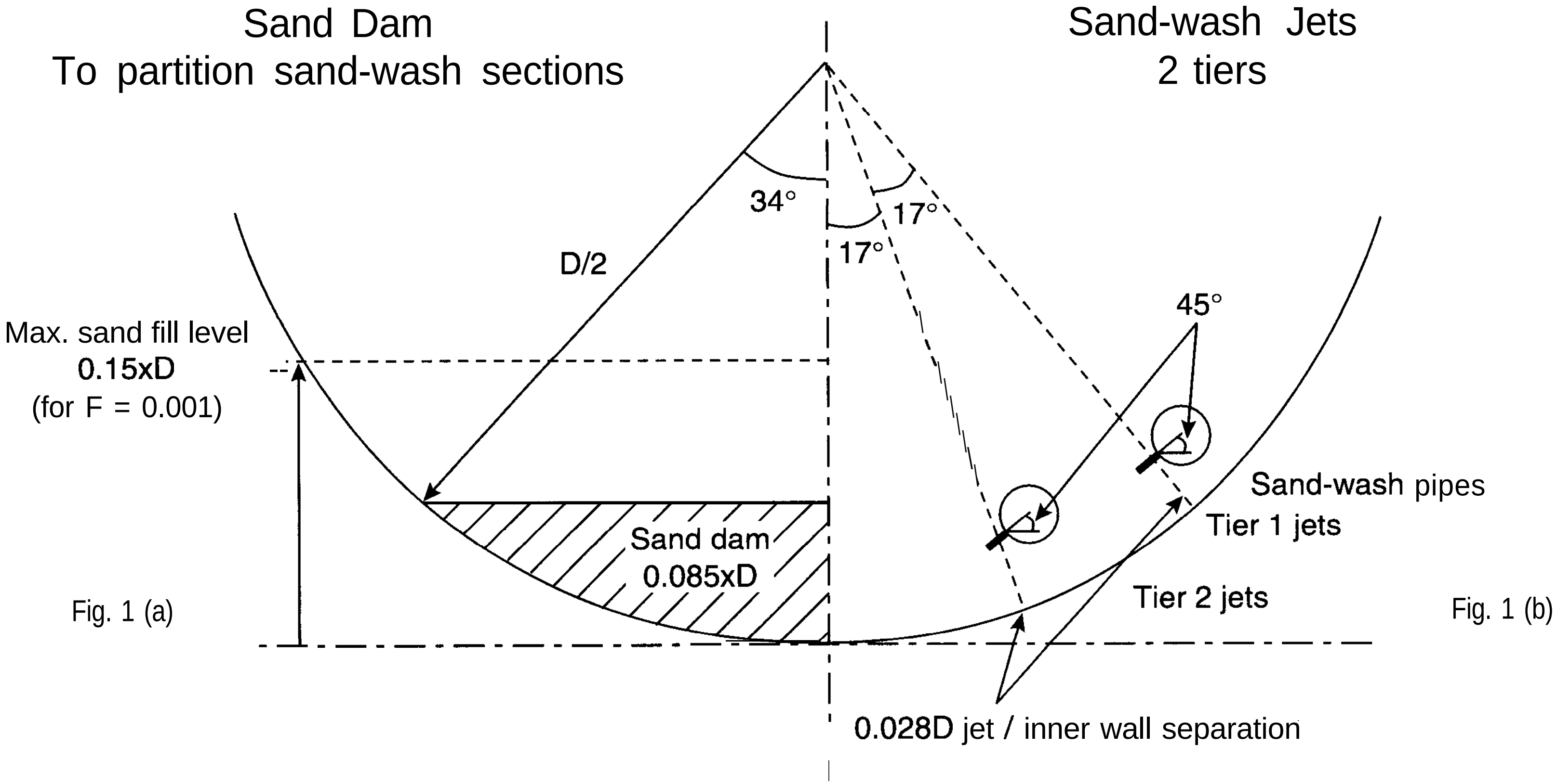


Figure 1. Sand-wash intervals

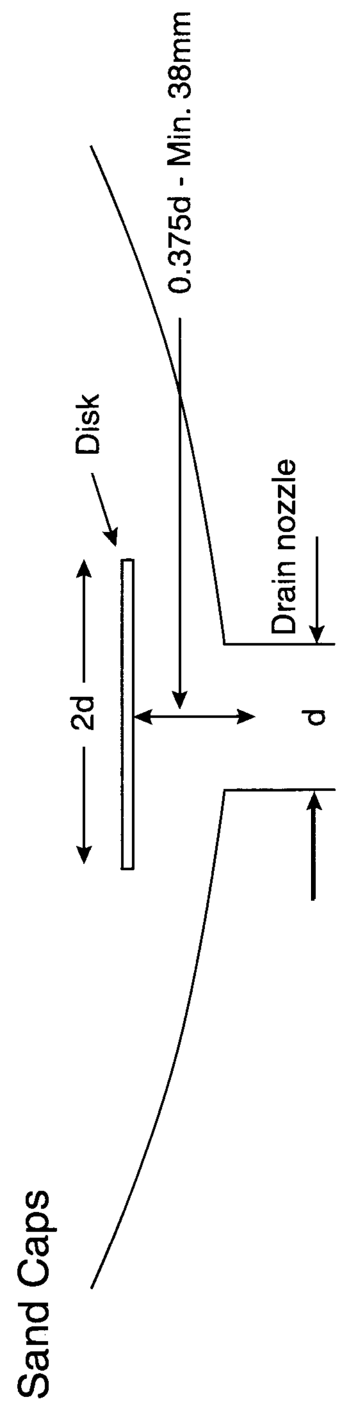
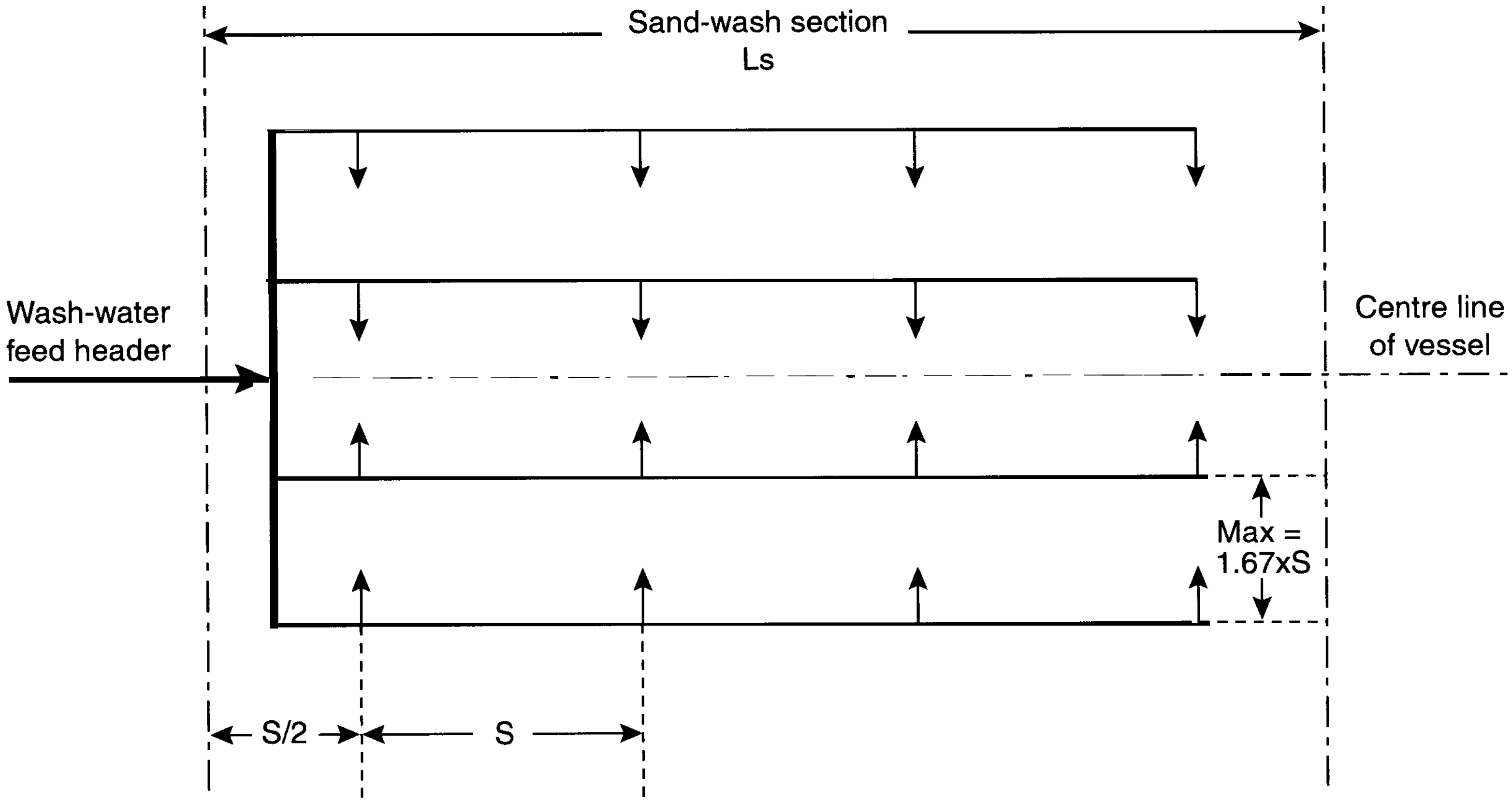
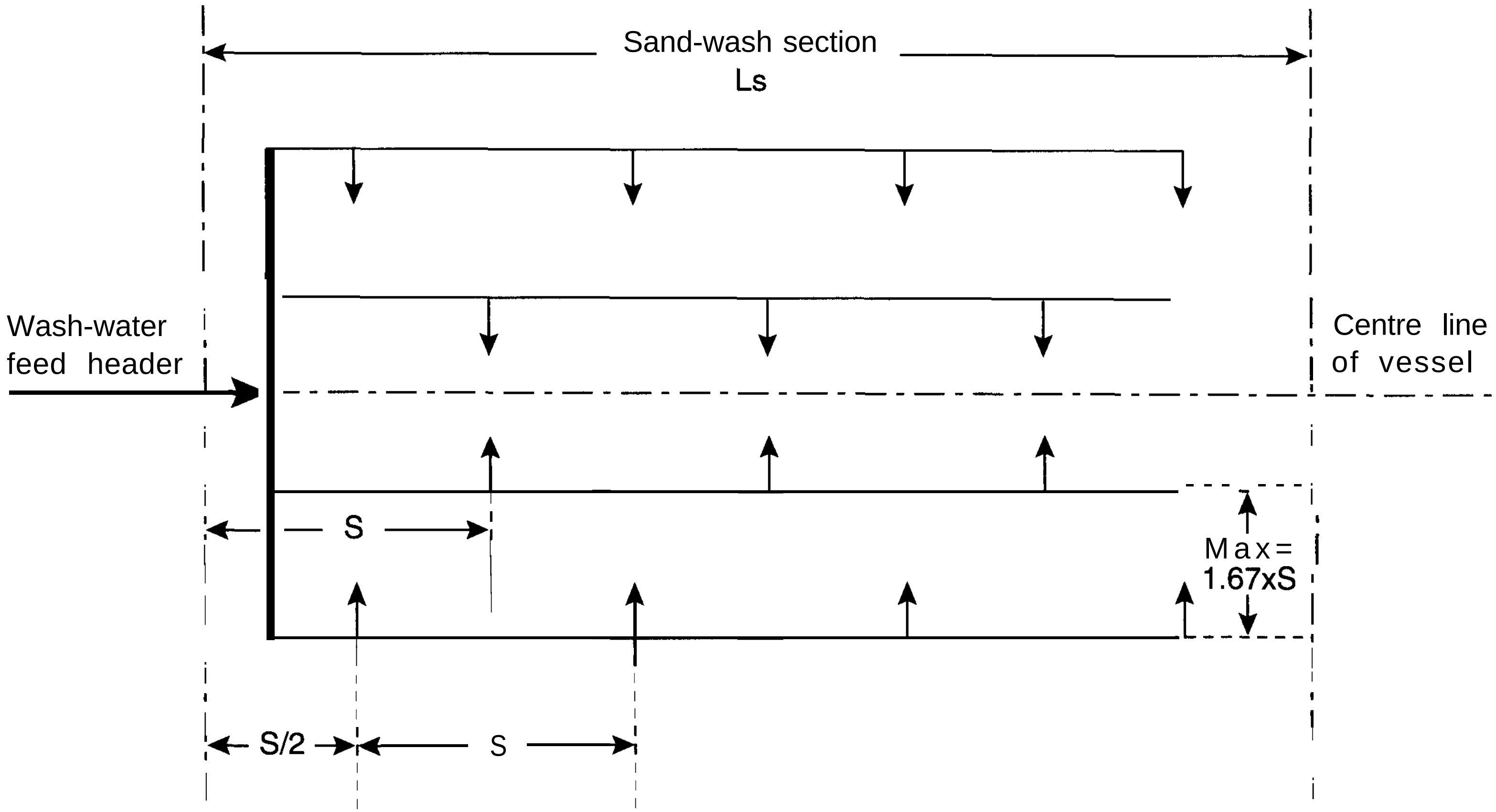


Figure 2. Sand drains.

Figure 3. Sand-wash jet spacing (1).



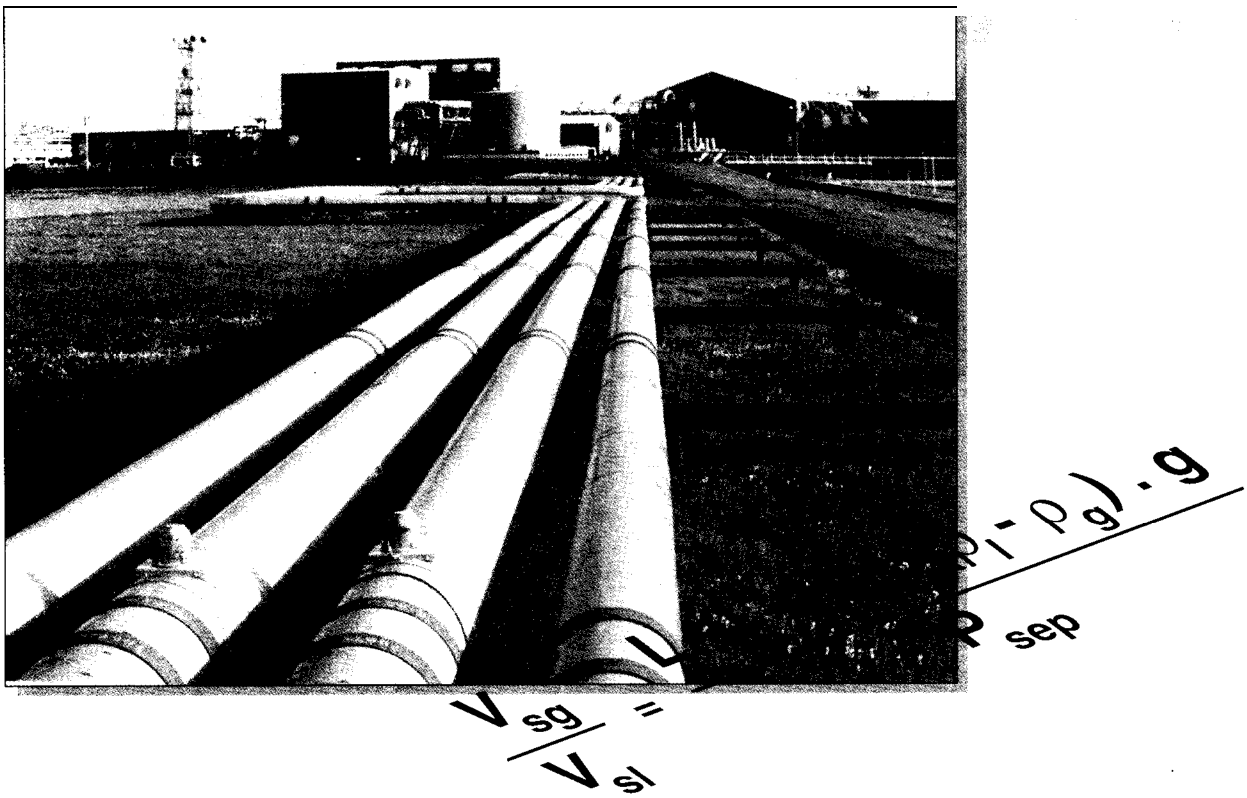
Arrangement of sand-wash pipes/jets (Number of jets divisible by four)
Jet spacing = S



Arrangement of sand-wash pipes/jets (Number of jets not divisible by four)
Jet spacing = S

Figure 4. Sand-wash jet spacing (2).

Section 20 Multiphase flow through deepwater systems



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20.1 Introduction

A large proportion of BP's future oil production is likely to be provided by deepwater offshore developments. A DeepStar study estimated that 95% of presently unexplored offshore acreage is in water depths of 3,000 ft or more. The main areas of activity at present are in the Gulf of Mexico (GOM) where BP has a 100% stake in 267 licence blocks in water depths of over 1,000 ft and is presently assessing developments in over 6,000 ft of water with current leases extending to 9,250 ft. West of Shetland the water depth is not as great (around 1,600 ft at present with potential developments at depths up to 4,000 ft). However the environmental conditions are very severe and have had a large impact on the design of the present Foinaven and future Schiehallion projects, and potentially for the latest discovery Suilven. The other main deepwater development areas are offshore Norway, where BP is drilling a well in the Voring basin at a water depth of 4,200 ft, and West Africa (600-7,000 ft) with present studies of the 4,500 ft WD Girassol development. Although these are the main deepwater provinces for BP there are also other potential future prospects in the Black Sea, Caspian Sea, and offshore Indonesia, China and Vietnam. Each of these areas have their own unique combination of conditions relating to the available oil and gas infrastructure, environmental conditions, and political climate etc.. but share in common the deepwater nature of the province. Figure 1 illustrates the environmental conditions characteristic of these areas of activity.

One of the critical issues is whether a tie-back to shallower water depths is practical as this may eliminate the need for ultra-deepwater surface piercing structures. Hence we need to investigate the hydraulic limitations on the step-out distance, but how far do we need to go ? At present the Troika subsea development requires 14 mile long flowlines from a seabed template in 2,670 ft of water to the existing Bullwinkle platform in 1,350 ft of water. One of the controlling features of the design is to maintain a high level of insulation on the flowlines in order to provide high flowing temperatures and to avoid hydrate formation. The GOM Neptune discovery is in 6,000-7,500 ft of water and could conceivably involve tie-back distances in the range 25 to 30 miles. However this may not represent the ultimate challenge.

Drilling has already taken place in the GOM at water depths of 7,800 ft (BAJA), and the deepest lease in the April 1996 GOM Central lease sale was in 9,289 ft of water. Hence a realistic design challenge is 10,000 it water depth. The Shell 'Mensa' development is in 5,300 ft of water and involves a 68 mile, 12" tieback to a host platform. Looking at the existing GOM infrastructure and the potential future plays indicates that a tie-back distance of 100 miles is a reasonable target. Hence we need to understand the limitations of pushing the design operating envelope to water depths of 10,000 ft with tie-back distances of 100 miles.

If we can design for such long tie-backs this may eliminate the need for ultra-deepwater surface piercing structures. However it is likely that conventional fixed platform designs will be uneconomic and impractical at such water depths and hence some floating structure will be required. Although there are a large array of potential host platform configurations ranging from floating tankers, tension leg platforms, spars, compliant towers, and semi-subs etc.. they share in common the need for riser systems to connect the subsea flowlines to the facilities. These risers can be rigid vertical conventional or hybrid designs, steel catenaries, or may be constructed of flexible pipe in a variety of configurations including catenary, 'S' and potentially double 'S'. As the water depth increases the pressure drop over the riser increases and has a larger impact on the hydraulic characteristics of the system. However there may be severe limitations in the ability of present design methods to accurately predict the hydraulic performance of such deepwater riser systems, and there is no field scale data on which to test and evaluate the performance of the models. The present limiting size of flexible risers is around 12", although at increased water depths sizes of up to 20" may be required. This large diameter brings into question the accuracy of the hydraulic models which have historically been validated on relatively small bore well tubing.

This section of the Multiphase Design Manual is based on the results of a study into the hydraulic design issues relevant to deepwater developments which was undertaken in 1996 under the sponsorship of the Deepwater MTL (Reference 1).

20.2 Key technical issues

The key technical issues can be summarised under the headings of energy, integrity and delivery:

20.2.1 Energy

'The provision of sufficient pressure to transport the required flowrates of fluids from the reservoir to the processing facilities.'

The system pressure drop increases with water depth and step-out distance as illustrated in Figures 2 and 3. This may be a limiting factor if the natural well head flowing pressure is insufficient. Most of the pressure drop is expected over the deepwater riser, although the proportion of the total pressure drop over the flowline will increase with the tie-back distance. A true assessment of design limits requires accurate pressure loss prediction methods for which there are two main concerns. Firstly, the shape of the flexible and steel risers utilised in deepwater is expected to influence the hydraulic characteristics and there is a need for design tools and guidance on the effect of the different riser shapes, to enable accurate predictions to be made and to assess the issues and limitations pertaining to the various configurations. One potential problem lies in the basis of the steady state multiphase design programs, such as Multiflo, which describe pipelines and risers by a number of discrete segments to capture the features of the topography. This may require many sections for a true representation of a catenary or 'S' shaped riser. However each segment is treated in isolation and hence the flow regime and holdup can change significantly at the segment boundaries. In practice the conditions in one segment can influence the downstream conditions, for example, slug flow generated in the flowline may persist in a riser, even though the riser is predicted to operate in another flow regime.

The second potential inaccuracy is in the fact that the hydraulic models developed for vertical and highly inclined flows have historically been developed from relatively small bore well tubing data and have not been validated at the increased diameters that may be required for deepwater developments.

In deepwater developments the reservoir can fail to produce naturally at much higher pressures than conventional shallow water developments. Hence it is possible that a great benefit can be obtained by providing some kind of pressure boosting or pressure drop reduction. Pressure boosting can be obtained by using down hole ESP's, multiphase pumps or subsea separation and boosting. Pressure drop reduction can be provided by using gas lift, drag reduction agents (DRA), using foaming agents, or removing the water at the wellhead. The optimisation of these methods requires that many factors be taken into consideration and can benefit greatly from integration with the reservoir performance, the flowline and riser hydraulics, and the process equipment performance. Steady state programs such as Multiflo are ideal for generating the hydraulic data provided that the accuracy is sufficient when applied to deepwater systems. Future system integrated models offer promise if the complexity does not hamper the simulation speed. This may require that a simple description is used for the reservoir performance as has been adopted by the BOAST program developed for AFP. Figure 4 illustrates the main energy related issues.

Transient considerations may also play a part in the energy related issues in that unacceptable conditions may arise due to unstable operation under normal flowing conditions or during shutdowns, restarts and flowrate increases. For example sufficient pressure may not be available to start-up wells against the high back pressure generated by stationary fluids in the flowline and riser. Temporary riser gas lift may be a solution to unloading the riser and kicking-off the wells which may flow naturally after start-up.

20.2.1.1 System pressure drop components

The production capacity of a deepwater development is principally determined by the reservoir pressure available and the pressure drop over the entire system from the sand face to the facilities. If there is sufficient driving energy from the reservoir to overcome the total pressure drop to the facilities the desired production capacity will be realised provided that the system integrity is maintained and the facilities can handle the delivered fluid flowrates. Figure 5 illustrates an integrated approach to the overall system pressure drop and indicates the various components that contribute to the system pressure drop.

(a) Reservoir

Starting at the reservoir the pressure is normally high initially and declines with time as the reservoir is depleted, although the pressure decline can in normal cases be moderated by pressure maintenance via water or gas injection for example. A reservoir that is normally pressured has an initial pressure approximately equal to the hydrostatic pressure for a column of water equal to the depth of the reservoir. For example a 15,000 ft reservoir from the seabed located at a water depth of 10,000 ft would have an initial pressure of approximately 10,800 psi. If adequate pressure support is not provided the pressure could drop to the bubble point of say 5000 psi. This will have a great influence on the energy available for transporting the fluids.

The range of reservoir characteristics and flowing conditions are great and it is beyond the scope of this manual to consider all options. This is best done when analysing specific developments. The aim here is to illustrate the effect of various parameters pertinent to deepwater systems and hence to illustrate the significance.

(b) Formation pressure drop

As the fluids are produced from the reservoir to the well bore a pressure drop is experienced over the formation. The relationship between the formation pressure drop and the flowrate is called the inflow performance and in its simplest form is expressed as a linear relationship between the oil flowrate and the draw-down, which may be typically 10 to 100 bpd/psi. Hence a productivity index of 10 with an initial reservoir pressure of 10,800 psi would yield a bottom hole flowing pressure of 8,800 psi for a oil flowrate of 20 mbopd.

(c) Well tubing pressure loss

The next component of the system pressure drop is the tubing loss which in shallow water systems is usually the biggest loss. However in deepwater the greater riser depth substantially increases the significance of the riser pressure drop. In the early production life of an oil field the flow into the well bore may still be above the bubble point, but the flow usually becomes two phase due to the pressure drop in the well tubing. Hence we need two-phase flow models to predict the pressure drop. For gas wells retrograde condensation usually also gives rise to a two-phase mixture.

Typical vertical two-phase flow patterns are illustrated in Section 1 where the flow may be bubbly just above the bubble point in an oil well but may change to slug, churn, and finally annular if there is sufficient gas evolution. The flow is generally symmetrical about the centre line due to axial gravity effects.

There are three main components to the pressure drop across the well tubing and these are due to friction, hydrostatic effects, and acceleration. At low flowrates the hydrostatic term usually dominates and requires accurate prediction of the liquid content or hold-up. At high flowrates friction dominates. The acceleration term is usually only of significance at very high flowrates which may be impractical for other reasons such as erosion. The competing hydrostatic and frictional losses usually result in a 'U' shape to the plot of tubing pressure loss vs flowrate.

Original two-phase pressure drop methods were based on simple correlations but they were found to be limited when applied to conditions outside the original data range on which they were developed. New methods based on mechanistic models for the flow regimes provide a higher level of confidence. The Ansari mechanistic model has been shown to give pressure drops of within 3% of the Foinaven EWT data. Typical tubing sizes of interest in subsea wells are in the range 4" to 7".

(d) Choke and manifold pressure loss

Usually the pressure loss associated with the subsea wellhead, manifold and jumpers etc.. is relatively low, typically tens of psi compared to the control chokes with hundreds of psi pressure drop unless full open for maximum production rate. The pressure drop generated by the production control chokes is usually determined by a calibrated correlation for the particular type of choke and its duty. A representative relationship is usually provided for design and during operation the performance of the actual correlation used can be monitored through well testing. It is often the case in steady state analysis that the solution involves calculation from the reservoir to the well head and from the facilities to the manifold, with required choke pressure drop inferred from the difference in pressures between the well head and the manifold.

(e) Flowlines and risers

The flow regimes in horizontal and slightly inclined flowlines are analogous to the vertical flow patterns but are complicated by the asymmetry caused by gravity. These have also been discussed in Section 1. The pressure drop across the flowline is mainly due to friction effects and hence usually increases with flowrate, although if the flowline is inclined uphill hydrostatic effects can be important. Again the mechanistically based models are thought to provide the best results when applied to real systems and for this purpose we recommend using the BP mechanistic model developed by GRE in 1992. This model reverts to the Beggs and Brill correlation under normal slug flow, but has a special model for terrain slug flow and has a modified stratified flow and annular flow model.

As the water depth increases the effect of the riser pressure loss becomes more significant and the shape of the combined flowline and riser characteristic becomes more 'U' shaped due to the increased hydrostatic effects, similar to a well bore. The largest uncertainties are expected to be in the accuracy of the riser pressure drop prediction because of the potentially large diameter (up to 20") compared with the small well bore sizes on which the models have been validated, and due to the possible complex riser geometry. The larger diameter and inclination effects are expected to lead to increased slip between the phases which will increase the liquid hold-up and the hydrostatic pressure loss. This is particularly true for three-phase flows where a much greater proportion of water is possible than in the homogeneous assumptions used in two-phase flow models.

Another potentially large inaccuracy can arise from the hysterisis nature of multiphase flow in which the flow regime developed in one section of the geometry may persist for sometime in downstream sections. This is particularly true for large slugs and is a major limitation of general steady state two-phase design programs in which the topographical segments are treated as independent.

(f) Arrival pressure

Historically a range of arrival pressures may occur during the lifetime of a development as the oil production rate declines, This can be accommodated by delivering to the lower pressure separators. For example the initial Pompano fluids are produced into the HP separator at 1250 psig, with the initial WHFP being high enough for the flowlines to run at a back pressure of 2000 psig at the platform upstream of a choke giving rise to bubble flow in the flowlines. As the available pressure subsea reduces the production can be successively switched to the MP separator operating at 450 psig and finally the LP separation stage operating at 150 psig. There is hence a large variation in the arrival pressure that can be used to provide for flexibility of operation.

The arrival pressure is usually fixed by the processing requirements although there will inevitably be some trade-off with the production capacity of the whole system if an integrated approach is used.

20.2.1.2 Network integration

An integrated approach to the system pressure drops is important if an optimised system is to result from the design. This approach is likely to overcome past failures in the communication between the various disciplines involved. For example, the subsurface analysis is likely to provide a minimum WHFP in order to give them a design margin, whereas the process design may adopt a high delivery pressure to provide them with a better margin for compression for example. The result can be that there is an unrealistically low pressure drop available for the interconnecting flowline and riser system. In addition, real systems can involve multiple wells, flowlines, and risers where it can be difficult to optimise production without an integrated model to determine which wells should be choked to what manifold pressure and produced to which separator pressure. Figure 6 shows a simple multiple well, single flowline example where the hydraulic characteristics of the wells are different and may be updated in a database using the results from well tests to adjust for changing operating conditions. It can be imagined that the approach becomes even more complicated if other operational constraints are involved such as well bore gas lift. This complication may give rise to the need for computer assistance in making operational decisions for optimised performance and for future field development planning, and can be followed-up further in Reference 2.

By consideration of the various components in the production system it can be seen that the natural oil flowrate may be maintained or increased by supporting the reservoir pressure, reducing pressure drop from the reservoir to the separator, and by reducing the delivery pressure. Maintaining the reservoir pressure can involve gas re-injection, voidage replacement, or water injection, usually with associated CAPEX and OPEX cost penalties. The pressure loss between the reservoir and the tubing can be reduced through improved completion design and stimulation to reduce the pressure draw down across the formation. The pressure loss across the well tubing, flowline and riser may be reduced by increasing the diameter (although too low a velocity may lead to hydraulic instabilities causing unmanageable transient flowrates and possibly high re-start pressures). Tubing and riser pressure losses may be reduced by using foams, gas lift or DRA and boosting may be provided by using ESP's, multiphase pumping, or seabed separation and pumping. Flowline pressure drops may be reduced by using DRA or removing the water prior to transportation. Reducing the arrival pressure may increase compression requirements and worsen oil-gas separation efficiency. The optimum solution will involve a trade-off between the competing economics.

In some **cases** pressure boosting or pressure drop reduction may provide economic alternatives to maximise production as a solution to the problems faced by natural flow.

20.2.2 Integrity

'Designing to ensure that the fluids are contained and that the flow is not impeded.'

One of the biggest potential deepwater problems here is related to cold temperature operation and the impact on production chemistry issues such as wax, hydrate, and emulsion formation. Lower ambient temperatures and longer transportation distances are likely to lead to increased cooling of the fluids, although this may be partly offset by higher wellhead flowing temperatures which may give rise to material limitations.

Two modes of operation are possible:

- **Normally operate at temperatures above the wax, hydrate, or emulsion formation conditions and provide measures for abnormal operation such as during shutdowns and restarts.**

- **Operate in the wax, hydrate, or emulsion formation region but use chemicals to suppress formation or the occurrence of problematic conditions.**

The DeepStar project has recognised the importance of assuring flowing conditions and has estimated that 1 0-30% CAPEX reductions could be realised if chemicals could be developed to avoid flowline blockages. The cost savings are related to moving from expensive bundle pipeline designs to bare pipe options ($ millions per mile saving), removing insulation requirements ($150,000 per mile saving), plus $ millions per mile if a pigging loop is avoided, with additional OPEX reduction of $1 million for a 36 hr shutdown at 30 mbopd.

Figures 7 and 8 illustrate flowline temperature profiles for a concrete coated pipeline and a bundle with a high value of insulation and indicates that high insulation and flowrates are required to maintain high temperatures over long distances.

Solutions to the thermal problems revolve around developing inhibiting chemicals, preferably lower cost alternatives to conventional chemicals, and insulation or heating.

The higher pressure operation leads to increased hydrate formation tendency and corrosion. Higher pressure also leads to more expensive pipeline designs and could lead to the use of High Integrity Pressure Protection Systems (HIPPS) if it is not possible or economic to design for full shut-in wellhead pressures.

Deepwater flowlines are expected to operate at a higher pressure due to the greater pressure drop across the riser. To provide acceptable velocities at the riser top within erosion or corrosion inhibitor shear stripping limits, the velocity in the flowline may be relatively low leading to the potential for surging, and solids or water deposition, which can have an impact on the flowing capacity, pigging, and corrosion.

At a first glance the integrity issues related to the topsides plant are not expected to be any worse than conventional shallow water developments because the arrival pressures and steady velocities are expected to be similar. Hence, the slug mechanical loads imposed on the pipework and the erosional velocities are expected to be similar. However, there is concern over the potential dynamic effects caused by the large hydrostatic pressure drops possible at increased water depths, giving rise to the stalling of large slugs as the gas in the flowline packs to provided the increasing hydrostatic pressure. When such slugs are produced into the facilities the extra hydrostatic back pressure is no longer needed and is converted to friction as the slug tail is accelerated giving rise to potentially high transient velocities and mechanical loads. Figure 9 illustrates the acceleration of the slug tail. Section 11 of this manual discusses further the estimation of slug forces. Figures 10 and 11 illustrate the potential integrity related issues.

A more detailed description of physical chemistry related issues is provided below:

(a) Low temperature limits and effect on physical chemistry issues

Low operating temperatures occur if the fluid is exposed to a cold environment for a significant period of time. The greatest effects that deeper water and longer step-out distances will have are on the lower ambient temperature in deeper water and longer residence time that the fluid will have in the transportation system, and riser auto-refrigeration effects. If the pressure drop available is relatively low such that a large diameter flowline and riser is required, the residence time will increase for the same production rate, and this will give rise to lower temperatures. It is apparent that for typical systems the operating temperature may be below 90°F even with high levels of insulation. As rule of thumb 90°F is a minimum arrival temperature for good separation efficiency.

Each of the physical chemistry issues will be introduced below and considered in the context of deeper water developments.

(b) Hydrates

Hydrate formation conditions are typified by low temperatures and high pressures, hence operations in deeper water with longer tie-backs are expected to increase the likelihood of hydrate formation which can reduce the flowing capacity, or in the extreme, complete flowline blockage can occur which can give rise to safety and containment problems as well as the potential high cost of remediation and production loss.

Hydrates are part of a chemical class of compounds called clathrates and can form at temperatures as high as 70°F in hydrocarbon/water systems operating at high pressures. Hydrates are solids composed of water and gas and are in some ways similar to ice with a frost like or ice appearance. The density of hydrates is similar to ice. However, whereas a pressure increase over water reduces the tendancy to form ice as the temperature is decreased, a pressure increase in a hydrocarbon gas/water system increases the tendancy to form hydrates as temperature is decreased. Figure 12 shows a typical hydrate decomposition curve for a oil/water/gas fluid on which is superimposed a pressure and temperature profile for a 50 mile subsea flowline from deepwater wells to a shallow water platform where the ambient temperature increases. It is seen that most of the flowline operates in the hydrate formation region in this case.

The classical approach to avoiding hydrate problems under normal operation is to insulate the flowlines or to use inhibitors such as methanol or MEG. Insulation or heating the flowline is employed to move the operating line to the right of the hydrate region, whereas chemical inhibition is used to move the hydrate curve to the left of the operating line. Insulation can significantly add to the CAPEX cost of a flowline ($150,000 per mile) whereas classical inhibition can require large quantities (1 0-40% of the water rate) of relatively expensive chemical which can add significantly to the OPEX cost. Even if the normal operation remains out of hydrate forming conditions, shut-down and subsequent cool-down can cause hydrate formation. If de-pressurisation is used to prevent hydrates during extended shut-down periods, chemicals may still be required during start-up since the flowline pressure often increases faster than the temperature, taking the operating line into the hydrate formation region. In this case there can be a trade-off between the methanol storage requirements and the start-up production profile.

By comparison with Figure 8 it is seen that the flowing fluids cool to 70°F in around 30 miles for a production rate of 5 mbd. At higher rates this distance is increased, hence turn-down to low flowrates needs to be considered and the lowest expected production rate can define the design case, depending on the effects of water cut. Figures 7 and 8 indicate that there are significant operating regions where the arrival temperature can be below 70°F. However the pressure may also be lower at the facilities, hence the determination of hydrate free conditions requires consideration of the flowline temperature and pressure profile, since as illustrated in Figure 12, it is possible for each end of the flowline to be free of hydrates, yet they may form part way along the length.

The high cost of classical hydrate inhibitors has recently led to a search for lower cost alternatives which are required in smaller quantities. The three main types of low dosage additives are kinetic inhibitors (THI), growth modifiers (HGI), and emulsion additives. Each have their own operational implications as discussed below:

Kinetic inhibitors

These are added to suppress the formation of hydrates like classical inhibitors but will not work at high sub-cooling (more than 1 0°C) or for extended time periods. This may have implications for three-phase flow wet/gas systems operating at low velocities where the water phase can have a very long residence time in the transportation system. The active agents in these inhibitors are polymers which are typically transported by a carrier solvent and are thought to be effective for gas, oil or condensate fluids.

Growth modifiers

Growth modifiers are active in the water phase and are designed to modify the growth of the hydrate and to encourage the crystals into the hydrocarbon phase. They are effective up to around 30% water cut and require that the hydrate crystals are transported as solids in the hydrocarbon liquid.

Emulsion additives

These inhibitors work in the oil phase by creating stable water-in-oil emulsions which do not have a high viscosity and which can be transported as a slurry, hence a hydrocarbon liquid phase is required. These are the newest and least tested low dosage inhibitors.

Each of the hydrate inhibitors have specific implications for processing and these are outlined in Table 1. The required dosage rate of the new inhibitors can be as low as 0.25 wt% giving treatment costs of around £4/bbl of water compared to MEG and methanol inhibition costs of El 1 and f7 respectively. Given the potential cost saving of avoiding insulation and possibly also of burying subsea flowlines, the new breed of hydrate inhibitors may have a large impact in the future since it will be likely that, even with the best insulation, transport over long distances will not be possible at temperatures above those at which hydrate can form. There is, however, a concern that the relatively small sub-cooling possible with these inhibitors will be insufficient for long step-outs and water depths.

The hydrate formation can sometimes be detected by a reduction in the produced water flowrate as water is being absorbed to form hydrate crystals. In other cases an increased pressure drop may be detected. Complete blockage is usually detected by reduced gas flowrates and pressure abnormalities. Some systems are being developed to detect hydrate and wax deposits on pipewalls and are based on ultrasonic techniques. These may be useful to monitor trouble spots such as uninsulated jumpers for example.

Complete blockages can be difficult to locate precisely and may take sometime to melt using inhibitors. A preferred method for melting hydrate plugs is to de-pressure the flowline. However this generally requires that the pressure is reduced on both sides of the blockage simultaneously as high differential pressures can cause the plug to move rapidly on melting, and can cause structural loads due to impact damage if the flowline changes direction, such as at the riser base for example. To de-pressure a subsea flowline blockage on both sides generally requires a flowline loop to the platform, which is a set-up typically used in the GOM as it also assists in turndown and allows for flexibility if producing good and bad producers to different separator pressures. However, the total subsea production is lost during the de-pressurisation. The Shell Mensa development only has one flowline and depressuring from both ends will be facilitated using a surface vessel to tie-in to the subsea manifold and lower the pressure on this side.

Melting hydrates using the application of external heat is possible. However care is required if localised heating is employed as very large pressures can be generated as the hydrate vaporises if it is contained within the plug.

Although we have discussed hydrate formation problems here it should not be forgotten that most production operations can suffer from a combination of simultaneous problems which can be compounded. For example, hydrate formation followed by or concurrent with wax formation can form particularly persistent blockages which have in the past lead to the abandonment of flowlines.

(c) Wax

Problems with wax or paraffin deposition have been identified by DeepStar as the top priority problem for deepwater developments. Typical cloud points or wax appearance temperatures are usually in the range 50 to 100 °F and are not very sensitive to the water cut or the operating pressure, but are dependent on modelling the operating conditions by using ‘live’, rather than ‘dead’ oil samples. Best estimates of the wax formation and deposition characteristics can be determined using real fluids in laboratory or field deposition tests.

As flowing temperatures fall below the cloud point small particles of wax may form which are carried along with the fluid. A temperature gradient is required for wax deposition on the pipewall to occur and this is usually the case with subsea flowlines where the wall temperature is lower than the fluid temperature. Because of these temperature gradients wax can deposit before the bulk fluid reaches the wax appearance temperature. Further downstream when the fluids have cooled to the temperature of the wall the temperature gradient and wax deposition is reduced, hence a wax deposition profile can exhibit a hump for a long flowline as illustrated in Figure 13. When the fluid and pipe wall has cooled to the same temperature the wax forms by diffusion in much lower quantities and can be transported by the fluids. One way of tackling the wax problem could hence be to cool the fluids produced fluids to the ambient temperature at the wellhead.

Potential problems with wax deposition can depend on the amount and type of wax deposited. Soft wax deposits may be carried along with the fluids and may not cause a significant problem, whereas hard deposits may require mechanical removal using pigging or melting by solvents or by flushing with high temperature fluids. There will likely be a trade-off between the lost production due to the reduced capacity resulting from paraffin deposition and the frequency of remediation, which will incur operating costs and deferred production

For example, the Foinaven crude has a wax content of 7.5-8.5 wt%, which begins to deposit at pipewall temperatures below 38°C (100°F). It is expected that 0.5-2 mm of wax could be deposited in the flowline and 1 .5-4 mm in the riser during 1 month of operation. The increased pressure loss is estimated to reduce the flowing capacity to 97%. Hot oil flushing requires a temperature of 10°C above the wax appearance temperature to be effective and it is estimated that a flushing period of 12 hours will be required, which if carried-out once a month would give a average capacity of 98.5%.

In practice optimised operation is likely to involve careful monitoring of the wax deposition and the evaluation of gains from wax control procedures. If operating conditions change rapidly it may be difficult to determine whether paraffin deposition is responsible for the reduced system capacity or an increase in the water cut, for example.

If pigging is used to remove wax deposits care must be taken to ensure that the pig does not become stuck. Some operations have benefited from using soft pigs so that some wax is left on the wall, which acts as an insulator and reduces wax deposition rates, whereas complete wax removal could give rise to high deposition rate and high pigging frequencies, not to mention potential wax handling problems at the facilities.

Deeper water operations and longer step-out distances are likely to giver rise to more severe paraffin problems where conventional remediation methods may have problems. For example, it may be difficult to maintain the temperature of flushing fluids over longer distances and pigging large quantities of wax may not be possible due to plugging. For these reasons there is considerable interest in the development of wax inhibitors which modify the wax crystals and prevent deposition on the pipewall. These chemicals may not prevent the formation of wax but may make them transportable.

As with hydrates, the interaction with other solids and inhibitors may be significant in practical operations. For example the BP West of Shetlands developments have the potential for paraffin deposition and relatively high sand production rates of up to 10 Ibs per thousand barrels of oil. The effects of the combined wax and sand production are little understood.

(d) High viscosity oils and emulsion formation

The viscosity of oils increases rapidly at low operating temperatures, although this may be partially offset at high operating pressures due to the effect of solution gas which reduces the viscosity. Figure 14 shows typical oil viscosity vs. temperature relationships. Measurements on field samples at several temperatures can be used to tune physical property correlations in multiphase design software, and hence to improve the accuracy of predictions. The increased oil viscosity leads to greater frictional pressure loss which may have implications for long flow-lines.

Some oils are very viscous and can be non-Newtonian giving rise to high pressure drops in normal operation and high re-start pressure requirements, Some of these problems have been overcome by using chemicals to form stable low viscosity emulsions or by ensuring that a lower viscosity fluid is in contact with the pipewall with the high viscosity fluid flowing down the middle, such as in core-annular flow.

The determination of the liquid viscosity is generally more difficult if oil/water emulsions are present since the viscosity can be many times higher than the individual phase viscosities. The viscosity of an emulsion depends to a large extent on the particle size distribution, which is a function of the flow history and the shear generated during transport through chokes, pumps, and the flowline itself, for example. High shear creates smaller particle sizes and tighter dispersions which gives rise to increased fluid viscosity as illustrated in Figure 15. Peak viscosities (over 30 times the oil phase value) typically occur near the inversion point where the emulsion reverts from a water-in-oil dispersion to a oil-in-water dispersion. The inversion point usually prevails at water cuts in the range 40 to 60 %.

Emulsions generally exhibit non-Newtonian shear thinning flow behaviour so that the viscosity is dependent on the shear rate. Problems can arise following shutdowns due to increased restart yield stresses and the effects of ageing. Longer flowlines with emulsion formation may require excessive pressures to restart the flow.

It is hence seen that emulsion viscosities depend on the temperature, water cut, and the flowing history. Typical assumptions of a weighted average viscosity of the oil and water are likely to grossly underestimate the liquid viscosity if emulsions are present. More realistic models for the viscosity of emulsions relate the ratio of the emulsion/continuous phase viscosity to the concentration of the dispersed phase modified by some power. This relationship can be tuned with measured data to improve the prediction accuracy.

A final point on the effect of high fluid viscosities is the impact on multiphase flow regimes and slug characteristics where it has been seen during tests with air and water/glycerol mixtures at Sunbury that high viscosities lead to a larger slug flow region. However the slugs are generally smaller and more frequent.

20.2.3 Delivery

'Ensuring that the process facilities can handle the fluid delivery from the flowlines and risers.'

The two main issues here are related to the arrival temperature, which if too low can give rise to separation problems and if too high may require cooling, and to the variations in the flowrates of the fluids delivered to the facilities as a result of normal hydrodynamic slugging, terrain induced slug flow, or resulting from transients caused by changing operating conditions and pigging.

The arrival temperature problems are related to the low temperature issues discussed in the integrity section which can lead to wax, hydrate or emulsion formation. In addition, good separation is also assisted by higher temperatures where around 30 °C (90 °F) is a good rule of thumb and may require the use of heating plant upstream of the separation facilities. A too high an arrival temperature resulting from HT well production can also give rise to problems and may require coolers necessary for compression or dehydration plant.

Some of the most severe potential operating limitations may be imposed by transient flowrates resulting from normal or abnormal operation. The greater flowline length and riser depth leads to a larger liquid inventory with potentially increased transient surge problems. In addition, the flowline topography and the shape of the riser can lead to severe slugging at relatively high production rates. This would make processing the fluids almost impossible and could require the use of control techniques such as gas injection, for example, to alleviate the condition. Figure 16 illustrates the delivery related issues.

Further discussion of the effects of deepwater riser configuration on severe slugging and the limitations of design methods is provided below:

(a) Effect of riser shape on severe slugging

The effect of the deepwater flowline/riser configuration on the hydraulic stability of a system was investigated by performing PLAC transient two-phase flow simulations for various flowline and riser geometries. Three riser configurations were used for the dynamic simulations and are shown in Figure 17. The configurations are all based on a 10,000 ft water depth and consider vertical, catenary, and 'S' shapes. The base case flowline configuration consists of a 20 mile horizontal flowline and sensitivities have investigated the addition of a 10,000 ft, 1° downhill sloping section to the riser base. All simulations are based on a 12" inside diameter system.

The PLAC simulations are based on a Pompano fluid composition with a GOR of 1200 scf/stb. A constant inlet mass flow rate is assumed, and two cases equivalent to 10 and 20 mbopd have been modelled using a fixed inlet temperature of 40 °F. The outlet pressure is fixed at 1000 psia.

Table 2 shows the results of the simulations and indicates that inclining the flowline down hill 1° at the base of the riser has a large effect of the stability of the system. For all cases with a horizontal flowline the stability limit is between 10 and 20 mbd, but the dip shape of the 'S' leads to much more severe surges with a horizontal flowline. With the inclined line the surges produced by the 'S' riser were not as severe as with the other shapes due to the trapped gas bubble preventing the riser completely filling with liquid.

(b) Conclusions on the limitations of transient codes to simulate deepwater flowline/riser hydraulic instabilities

Although general purpose transient two-phase flow simulators offer the most potential for assessing the hydrodynamic stability limits of deepwater production systems there are presently some concerns over the limitations of the codes. This is highlighted in Figure 18 which shows flow pattern map for the BHRg lazy 'S' riser tests (Reference 3) compared to the results of the OLGA code. It is seen that OLGA predicts a much smaller severe slugging region. Present limitations of the transient codes are thought to be:

Cannot inherently model slug flow

Flowline slugging can have a large impact on the severe slugging region. However general purpose transient two-phase flow computer programs such as PLAC and OLGA usually employ averaged equations when slug flow is predicted, so they are insensitive to the flow and pressure variations that result from normal hydrodynamic slug flow. OLGA has been modified to incorporate a 'slug tracking' model, but this is a simple model that initiates slugs of a pre-determined length when formation conditions occur. In some cases this can 'force' the result and while it may be a useful tool, it does not necessarily lead to a valid simulation. The implementation of the slug tracking model also slows the code down significantly. Work is in progress on both PLAC and OLGA to improve the modelling of hydrodynamic slugging.

Poor boundary conditions

Normally when transient codes are used to model severe slugging in flowlines and risers simple fixed inlet and outlet boundary conditions are usually specified. In most cases the inlet mass flowrate supplied to the flowline is specified as constant, as is the platform arrival pressure boundary.

In practice the flowrates at the start of the pipeline may vary as a result of slugging or heading in the wells, or even as a result of flowline back pressure variations affecting the well inflow rate. If there are interactions between the wells and the flowline this may require that the wells are modelled also, although if the pressure drop across the choke is high this may de-couple the well and flowline dynamics.

At the other end of the system variations in the arrival pressure may result in practice due to the interaction with the process plant and its control action. Even relatively small outlet pressure variations have been shown in the past to completely dominate the transient response of flowlines. There have been various projects aimed at coupling dynamic process plant simulators with transient pipeline codes such as the D-Spice/OLGA interface. This goes some way to simulate the interaction between the flowline/riser and the facilities, but the resulting tools are relatively cumbersome and may not be appropriate for simulations early in a project where there is insufficient definition of the facilities. Under these circumstances a sensitivity to typical boundary condition variations may be a sensible way to proceed.

Three-phase flow effects

If water is present in the fluids there will be a tendency for slip to take place between the oil and water phases as well as the gas and liquid phases. This will be more pronounced at low flowrates typical of severe slugging operating conditions and will give rise to a higher liquid holdup and density compared to the homogeneous liquid phase assumptions usually applied to handle water cut effects in two-phase flow steady state and transient simulators. In addition, the homogeneous assumption cannot model the actual oil and water arrival rates and this may affect the operation of the process plant if, for example, transient sweep-out produces a large water slug. It is expected that proper accounting of the three phase flow may result in an increased region of flow instability.

A simple three-phase flow model has been implemented in OLGA and there are plans to include one in PLAC also. BP also developed an in-house code TRANFLO which compared favourably with dip slugging experiments and may be developed further in the future.

Single composition

Both PLAC and OLGA use a single fluid description for the transient simulation which is used to generate look-up tables of the physical properties. This is adequate for most cases, but if modelling riser base gas injection, for example, the injection gas is forced to have the feed gas composition. In practice the injection gas will have been processed and will have a different solubility in the oil which will effect the free gas content and the overall effect on the system stability. Compositional tracking has been attempted with both PLAC and OLGA with limited success due to the much longer computing times involved. The TACITE code developed by IFP has compositional tracking but is limited to two components at present.

Simple one dimensional flow models employed

The hydraulic models employed in the transient simulators rely on simplified one dimensional descriptions of the flow regimes with closure laws derived from correlations. In some cases this approach has limitations due to the three dimensional nature of the flow and the inherent complexity in real systems, In addition, the closure laws are usually based on laboratory scale data that may not be valid under field conditions.

Field trials are proposed on Foinaven in 1997/8 that should provide valuable operating data and experience for the testing of the steady state and transient design tools.

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20.3 Deepwater operating limitations and solutions

We shall now combine these operating limits in order to assess those which are a direct result of the deepwater nature, and those which are more generic, in an attempt to identify the critical issues for deepwater developments. The most promising solutions are discussed and the implications for future deepwater system design and operation are outlined. It is not possible to consider limitations simply as low throughput and high throughput limits as the dynamic nature of two-phase flow leads to some obscure results. For example, low steady flowrates may lead to riser slug stalling and accelerations that produce higher loads than steady production at high throughputs.

A large proportion of the operating limitations are related to high or low flowing velocities which are partly determined by the characteristics of the reservoir fluid, and are principally beyond our control, for example the GOR and water cut. However down-hole gas and water separation may change this in the future. The operating pressure and the flowline size are also major variables.

It can be difficult to assess the relative importance of the various operational constraints in practice as it may not be possible to provide a large enough operating envelope which avoids all limitations, and it may be required to asses the grey areas of operability which mitigate against the worst problems. Such an assessment requires that all the relevant considerations are understood and weighed-up. This often requires input from various disciplines in a true integrated design.

Figure 19 shows a typical operating envelope that was produced for the Foinaven development and considers the capacity limit due to the available pressure drop, the unstable hydraulic operating limit, the erosion rate, sand deposition limit, and the minimum arrival temperature. This only represents a few of the possible operating limitations, and the Figure shows a clear operating envelope. If the water depth and tie-back distance were to increase the flowline pressure would increase and the arrival temperature would decrease, with the result that some of the limit lines may move to reduce the operating region, or may cross-over to produce a grey operating region.

The relative importance of the various operating limits depends to a large extent on the specific development and hence it is not possible to reliably generalise. It is recommended that the results of this study be applied to a number of potential future deepwater developments in order that the research efforts can be better focused.

20.3.1 Critical factors for deepwater hydraulic design

The most critical factors for the successful design and operation of Deepwater developments are judged to be as follows:

(a) Provision of required system pressure drop

The differential pressure across the transportation system will increase with water depth and tie-back distance, hence requiring higher wellhead flowing pressures and larger flowline diameters. The flowline inlet pressure may be increased by:

- **High reservoir pressure**
- **High well PI**
- **Reduced tubing loss**
- **Low choke pressure drop**

The tubing loss may be counteracted by adding energy in the form of down hole ESP's or reduced by using DRA/foams or gas lift. Gas lift has the disadvantage of potentially increasing the flowline pressure drop.

The capacity of the flowline/riser system may be increased by:

- **Boosting at the wellhead or riser base**
- **Reducing the delivery pressure**
- **Increasing the flowline size**
- **Using DRA or foams**
- **Water removal at the wellhead or down-hole**

The effect of each mitigating measure should be judged using accurate pressure drop prediction methods in an integrated system model. Bearing in mind the practicality and cost of some of the boosting measures, the most promising technologies for deepwater are considered to be down-hole ESP's, water removal systems, the use of drag reducing agents, and accurate integrated model predictions. The addition of energy using ESP's is probably the most efficient and reliable means, but may still have problems due to long distance power transmission. Other means of supplying power using injection water, for example, may be worthwhile considering.

(b) Low temperature related problems

Low operating temperatures are likely to result from the lower ambient temperature and longer fluid residence time in the transportation system. The low temperature related problems may be overcome by:

- **Reduced flowline diameter**
- **Insulation**
- **Use inhibitors**
- **Chemical or electrical heating**
- **Pigging**
- **Hot oil and solvent flushing**
- **Water separation at wellhead**
- **Don't insulate risers**

Because of the difficulty and cost of maintaining the temperature of long tie-backs the most promising control method is the use of low dosage inhibitors but these may have limited application due to the relatively low subcoolings that can be tolerated. Cooling at the wellhead may be a potential for wax problems since if the fluids are cooled to the temperature of the pipe wall there is no temperature gradient for wax deposition and the small amount of wax that is formed by diffusion can be transported as a slurry. The auto-refrigeration in the deepwater risers was previously not anticipated and should be the subject of further study.

(c) High temperature problems

High temperature related problems are only likely in the vicinity of the wellheads and are mainly associated with increased corrosion and the strength of coatings and flexible pipes. These issues are being addressed by HP/HT developments and are not considered as critical here.

(d) High velocity limits

Steady state velocities at the receiving facilities are not likely to be any higher than with conventional developments, but the transient velocities caused by slugging could be much higher as the water depth and the riser diameter increases. This can give rise to intermittent erosion and corrosion inhibitor film stripping. The biggest problem for deepwater is considered to be in the prediction of slug accelerations and the pipeline load and fatigue implications. If accurate predictions cannot be made then in-service monitoring of the material wall thickness and fatigue life could be necessary. Loads may be reduced by aerating the slugs or moderating the slug accelerations.

(e) Low velocity related problems

The high back pressure generated by Deepwater risers gives rise to lower operating velocities in the flowline which may lead to solids deposition and water accumulation. Pigging can be used to control these problems to some degree, but is costly and time consuming. There is little that can be done about the solids problem apart from separation at the wellhead. Remote water removal is also a possibility as is the use of emulsion forming agents to maintain the water in suspension in the oil phase, and hence avoid water accumulation and corrosion.

(e) High pressure operation

High pressure operation is likely to be unavoidable. There is still a potential that there will be a considerable difference between the wellhead shut-in pressure and the maximum required operating pressure of the flowline. HIPPS systems can be used to reduce the flowline pressure rating and cost if the system is able to monitor the pressures and control effectively. There are a number of circumstances that can arise where the measurement of the pressure at the ends of the flowline is insufficient to infer the pressures throughout, such as during pigging, or if blockages occur. Here dynamic hydraulic models, tuned to field measurements, may improve the accuracy of the shutdown system.

(g) Problems with gas and liquid flowrate variations at the facilities

Larger diameter flowlines and risers are likely to generate bigger normal hydrodynamic slugs and the longer flowline length combined with the shape of the riser may increase the likelihood of severe slugging. The larger system inventory can also lead to more severe surges during transient operation. Deepwater systems are hence likely to experience more severe flowrate variations at the facilities than conventional developments. The flowrate variations resulting from severe slugging are so large that this mode of operation is to be avoided, so the use of control measures such as riser base gas lift or the use of foams is essential, as are design tools required to predict the occurrence of severe slugging. Methods for limiting the effects of transient operation, such as reducing the liquid inventory and using operating procedures, are also important. Normal slug flow is more difficult to avoid, and methods for active control and detection subsea may be worthwhile pursuing.

The critical areas discussed above are illustrated on a deepwater schematic in Figure 20.

20.3.2 Assessment of limitations

The real life assessment of critical design limitations is likely to be complicated by the wide range of factors involved, the variation of operating conditions over the life of the field, and the large changes in the flowing conditions from the reservoir to the processing facilities. There is also a compromise to be made regarding the relative importance of various issues which may reduce one problem while worsening another.

Table 3 shows an assessment of the significance of a number of variables related to deepwater developments and may be used to rank the configuration of a development with regard to potential hydraulic limitations. The higher the overall score, the more challenging the hydraulic design is likely to be. This is an initial attempt at this approach and may be elaborated on in the future if considered to be worthwhile.

The practical application of hydraulic design limitations can be communicated using design charts such as illustrated in Figure 19. However the number of limits may be many and such charts may become complicated. In addition the operating envelope may change considerably as the field is depleted and the system characteristics change. For these reasons it is anticipated that in the future the operating limitations should be computerised and regularly updated to take account of changing conditions, Because of the many variables it may not be possible by intuition to determine optimum operations, and intelligent software tools may be required. These could be linked to the monitoring and control system and warn the operator if corrective action is required. An example of this could be if the measurements of the fluids into and out of the flowline are used to determine the liquid inventory and warn the operator when to send a pig to control the build-up of liquid. This type of approach could be important as operations become more complex and more marginal.

References

1. Fairhurst, C.P,"Hydraulic Design issues for Deepwater Developments", SPR report number SPR/TRT/101/97, June 1997.

2 Fairhurst, C.P,"The Virtual Operator-The Concept of an Advanced Monitoring, Control, and Optimisation System for Multiphase Production Operations", BPX Facilities Engineering, November 1996.

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Type of additive	Kinetic inhibitor	Growth inhibitor	Emulsion additive	Methanol
Application:				
Multiphase	*Yes*	*Yes*	Yes	*Yes*
Gas	Yes	Unlikely	No	*Yes*
Condensate	Yes	Yes	Yes	*Yes*
Oil	Yes	Yes	Yes	*Yes*
Dose rate	0.1250.25% of water rate	0.25% of water rate	2% of water rate	10-40% of water rate
Operating limits:				
Temperature	*10°C* subcooling	15°C subcooling	15°C subcooling	15°C subcooling (dependant on dose)
Water cut	No limit	30-40% WOR	30-40% WOR*	No limit
Effectiveness against ice	Not typically effective	Not known	Probably effective	Very effective
Additional equipment	No	Heating / demulsification	Heating / demulsification	Recovery
Additional chemicals	No	Demulsifiers	Demulsifiers	No
Known deployment problems	Emulsion formation viscosity of blend	Not	Not	Material compatibility
Regenerable	No	Unlikely	No	80% typically
Downstream impact	None known	Possible product contamination potential cause of haze in heavier fractions	Possible product contamination potential cause of haze in heavier fractions	Occasional product contamination - loss of LNG specification
Disposal stream	In water	Typically in oil	In oil	In all phases
Environmental consideration	Water quality hydrocarbon limit	Not known	Not known	Onshore disposal limits
UK regulatory rating for disposal	D/E	A	Unknown	E

** For these additives, the amount of hydrate formed and so the effective operating limit could be a complicated function of flow rate, flow regime, gas amount and water cut. At this stage in development, practical limits are best determined from field trials, or better, from loop tests*

Table 1. Characteristics of hydrate inhibitors

Table 2. Effect of riser shape

Riser	Flowline	Oil rate (mbopd)	Static analysis	Dynamic analysis
Vertical	20 mile	10	Unstable	Small fluctuations – 10bbl surge
Vertical	20 mile	20	Stable	Stable
Vertical	20 mile with 1° dip	10	Unstable	Severe slugging – 1069bbl surge
Catenary	20 mile	10	Unstable	Small fluctuations – 10bbl surge
Catenary	20 mile	20	Stable	Stable
Catenary	20 mile with 1° dip	10	Unstable	Severe slugging – 1432bbl surge
‘S’	None	10	Unstable	Stable
‘S’	20 mile	10	Unstable	Slugging – 306bbl surge
‘S’	20 mile	20	Stable	Stable
‘S’	20 mile with 1° dip	10	Unstable	Severe slugging – 866bbl surge

Table 3. Nature of extremes/effect

Variable	**Extremes / Effect** ← **Worst** ←	→ **Best** →	**Significance** (out of 10)
Water depth	Very deep (10,000ft)	Shallow (Oft)	1 o-o
Flowline/riser diameter	Large 28”	Small 3”	10-l
Upstream flowline length	Long 100 miles	Short 0 miles	*8* Lowflow - 0 2 highflow
Inclination of upstream flowline	*O-20”* down	*7-20”* up	IO Lowflow - 0 2 Highflow
Geometry of riser	Double ‘S’ ‘S’ Catenary Vertical		IO Lowflow - 2 Lowflow 2 Highflow - 2 Highflow
% of production up the riser	100%	small %	1 o-3

Multiply to get comparisons of different combinations

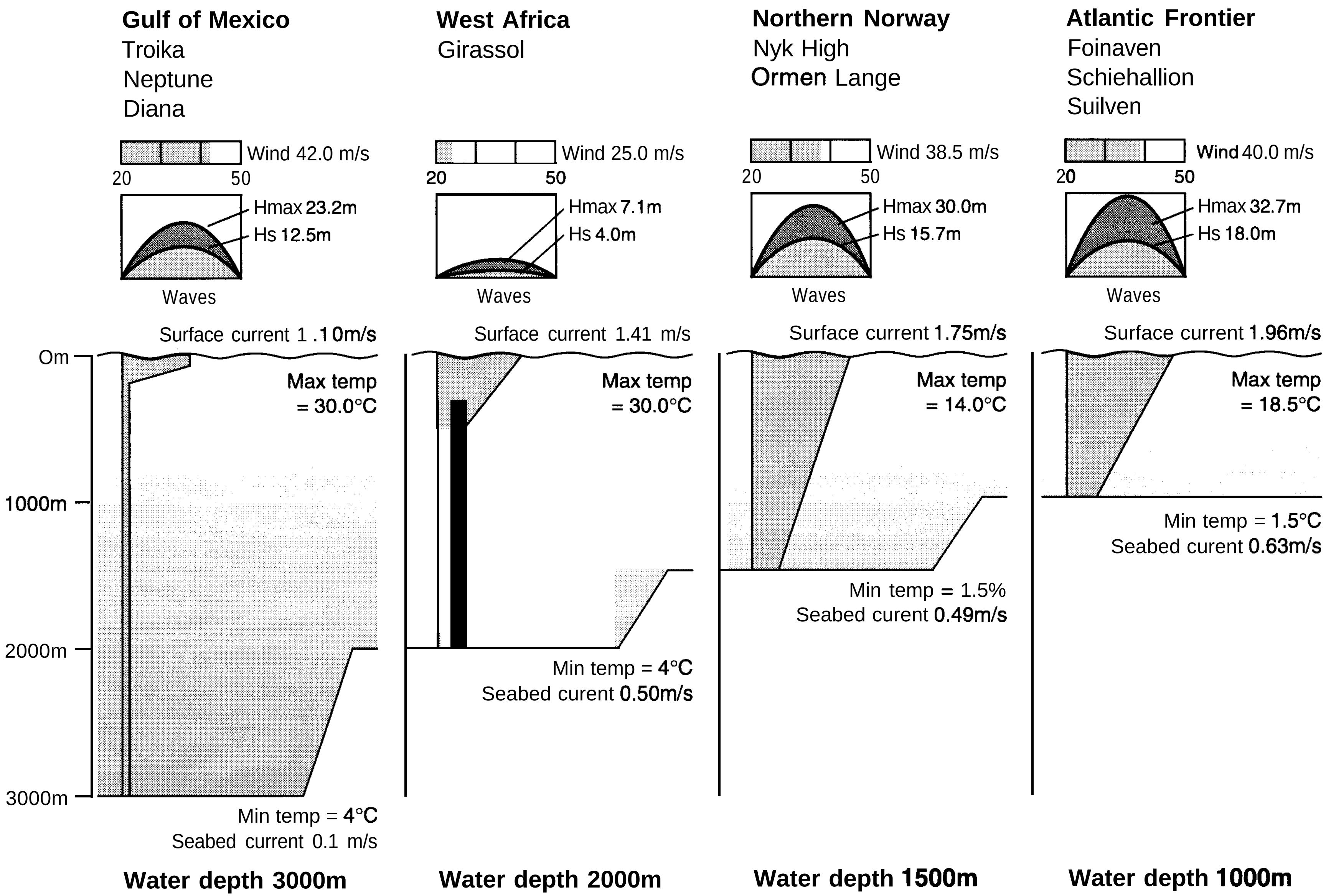


Figure 1. Areas of activity

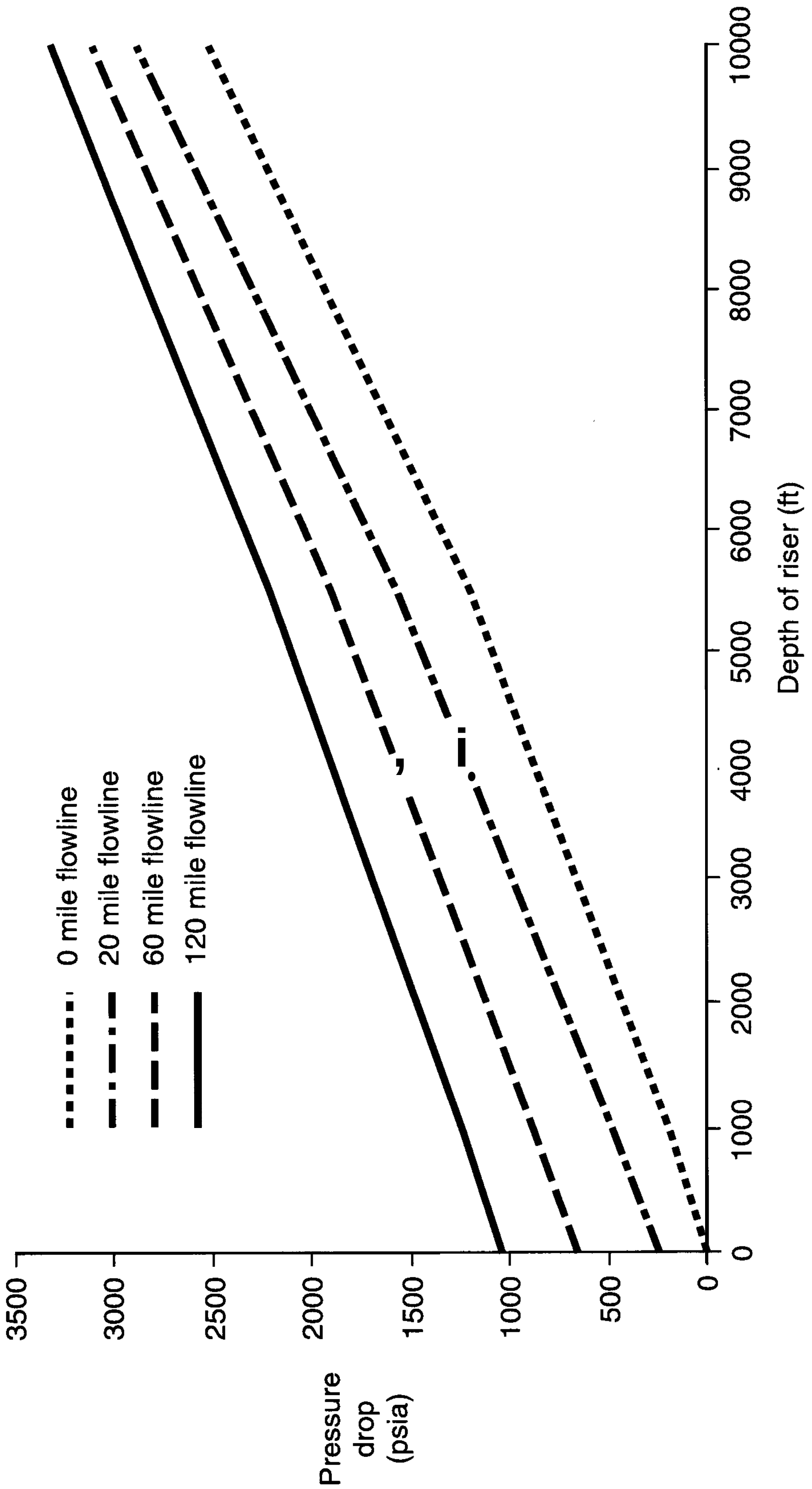


Figure 2. **Effect of depth on pressure drop**
– 12” ID, 1200 GOR, 0% WC, 20 MBD, 1000 psia delivery

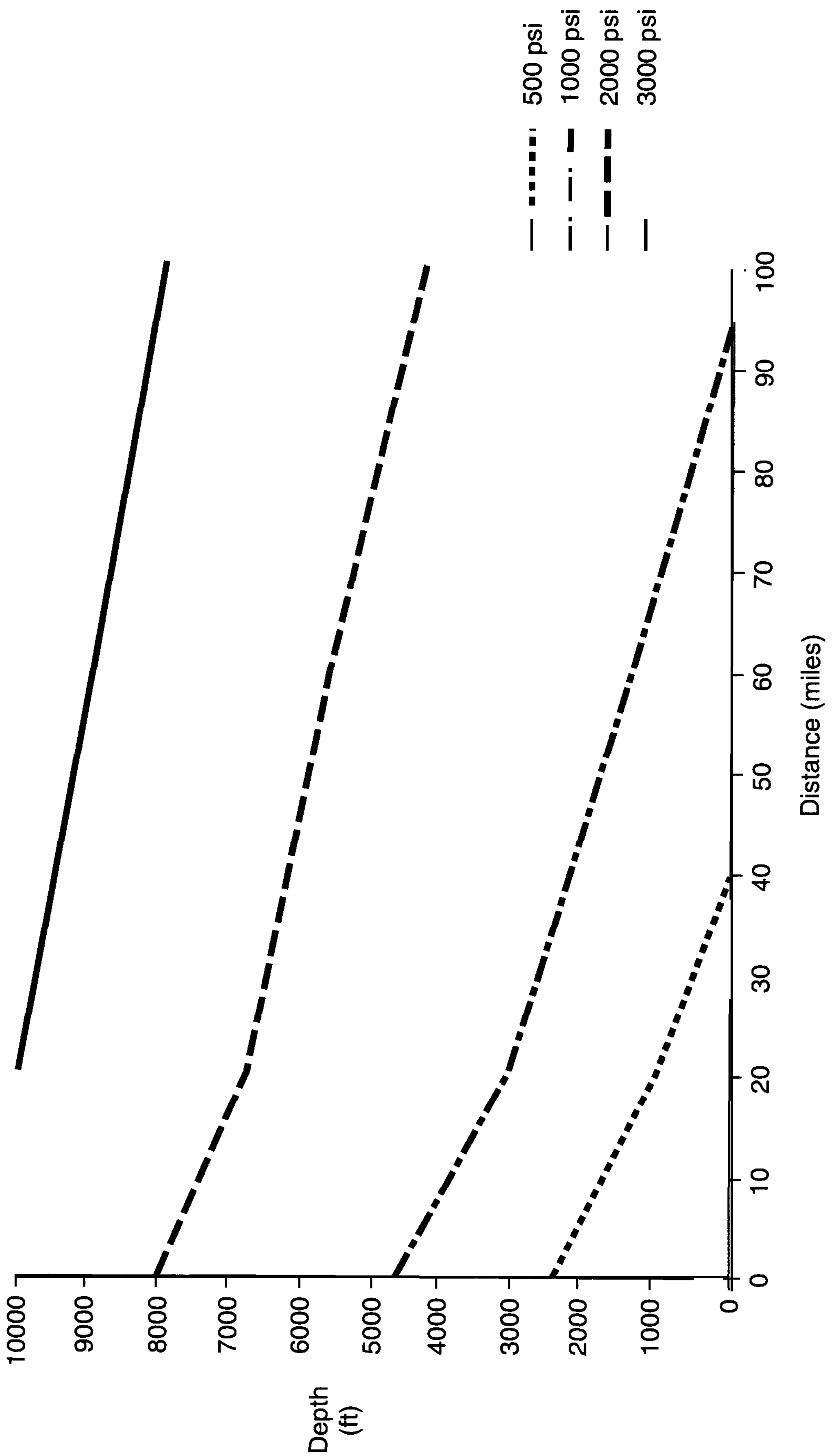


Figure 3. **Effect of distance on pressure drop**
– 12" ID, 1200 GOR, 0% WC, 20 MBD, 1000 psia delivery

Limitation	Impact of deepwater	Solutions
Insufficient flowing pressure	• Pressure drop increases with water depth and step-out distance	• Accurate prediction models • Proper integration modelling • Multiphase pumping • Gas lift • Separation and boosting • Drag reducers • Foaming agents • Lower arrival pressure • Subsea water removal

Figure 4. Energy related issues – Provide pressure for transport

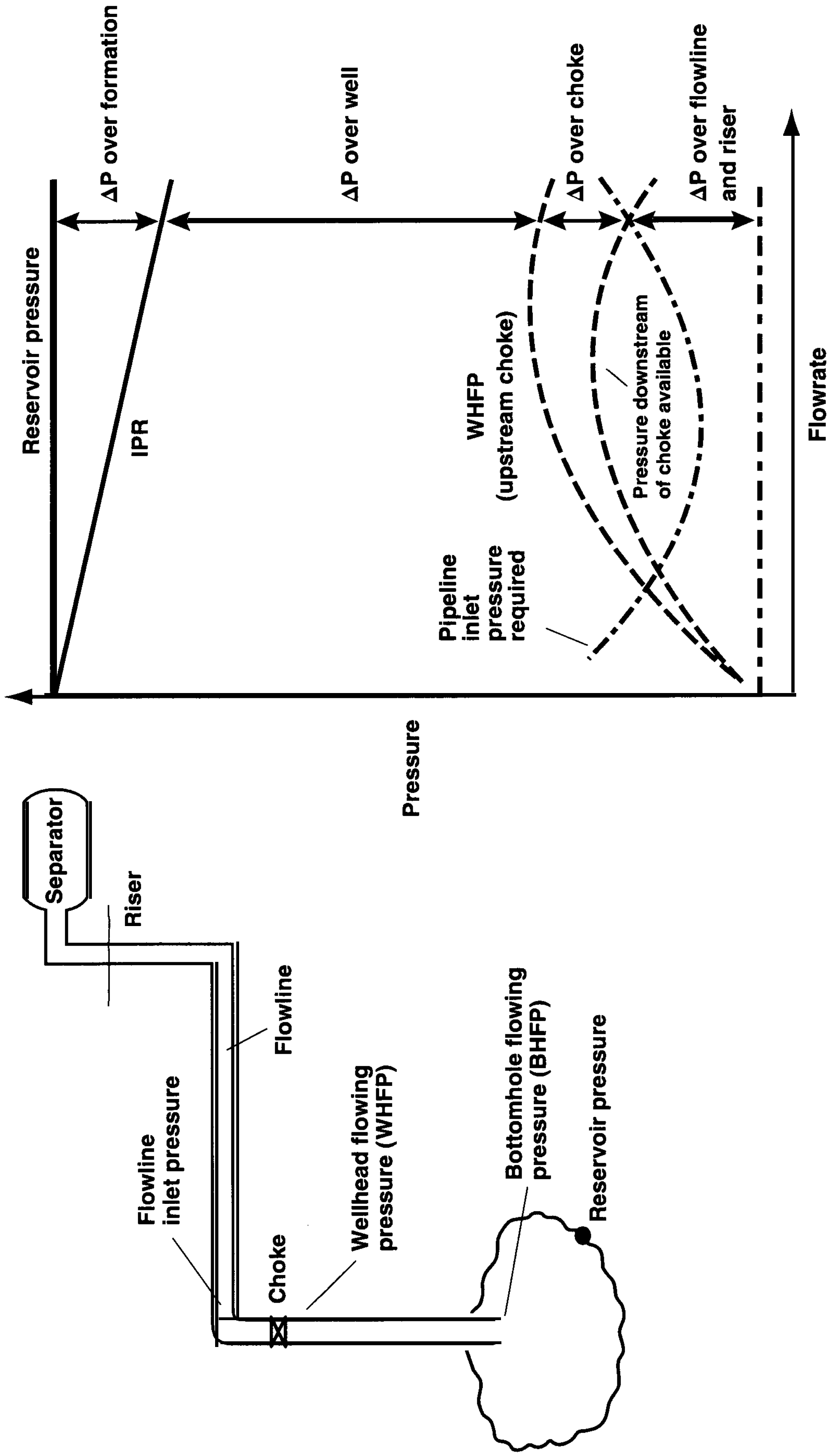


Figure 5. **Wells as part of the production system**

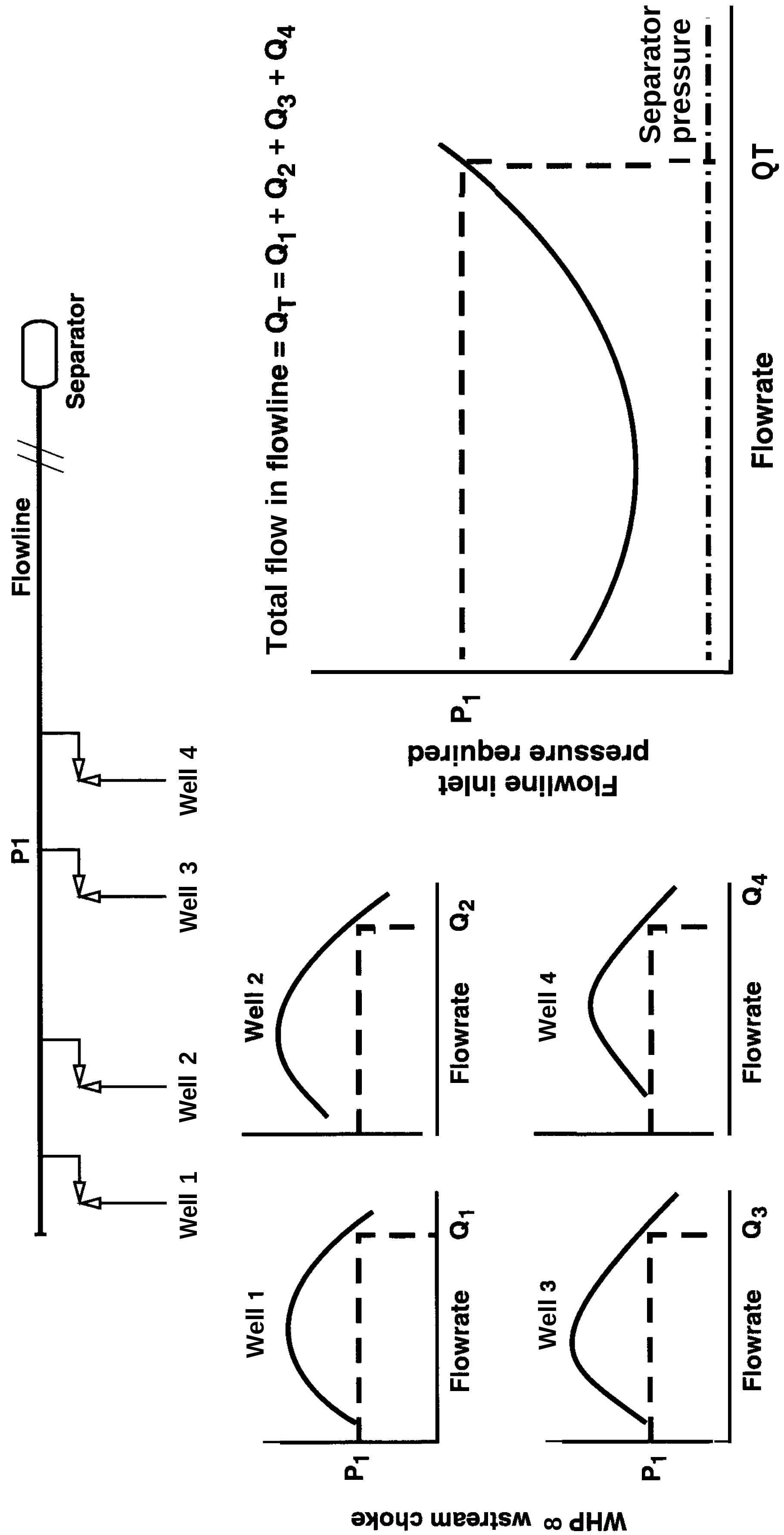


Figure 6. **Multiple wells to one flowline to separator**

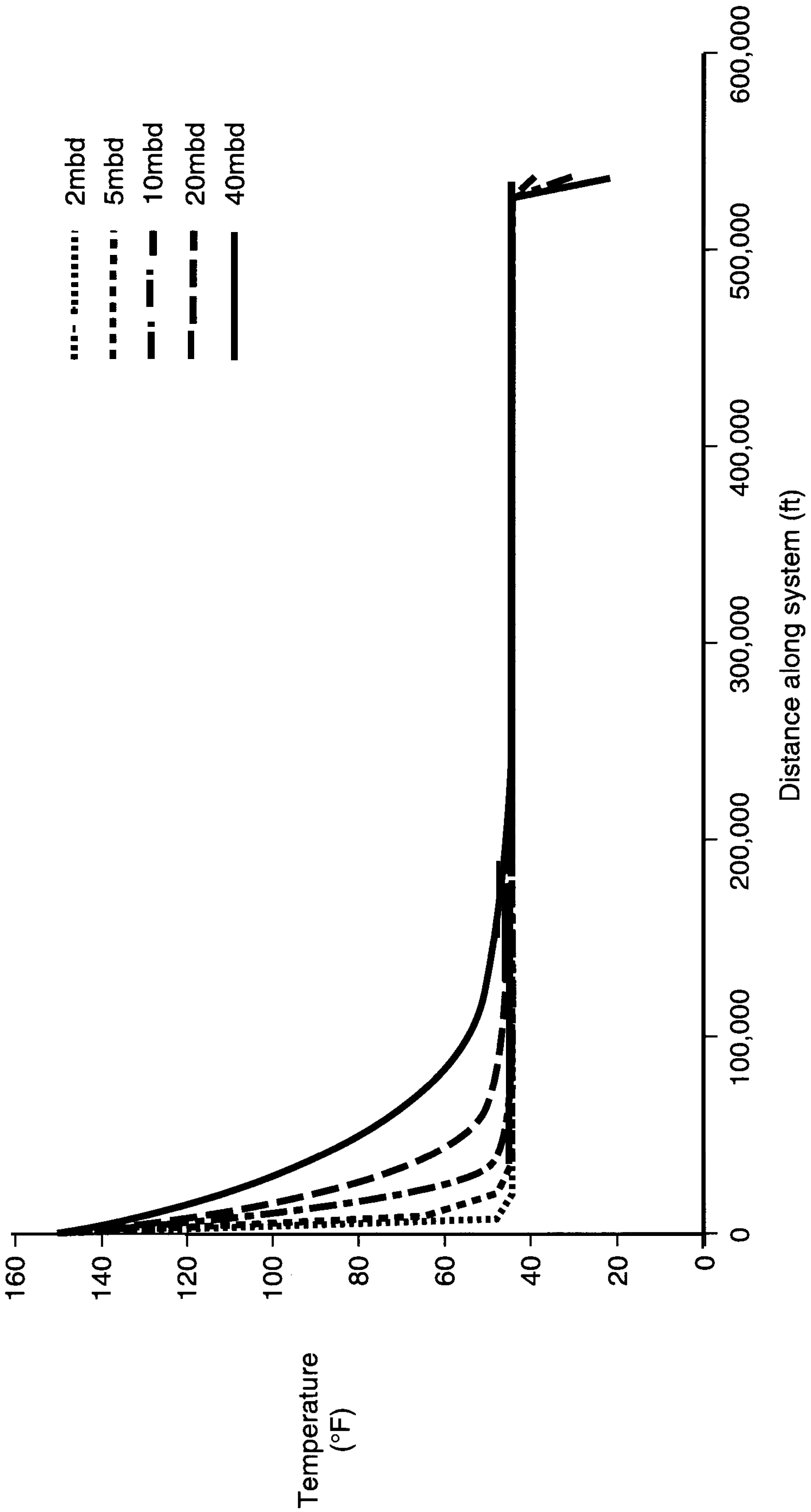


Figure 7. Temperature profiles for pipeline with 2" of concrete coating with burial.

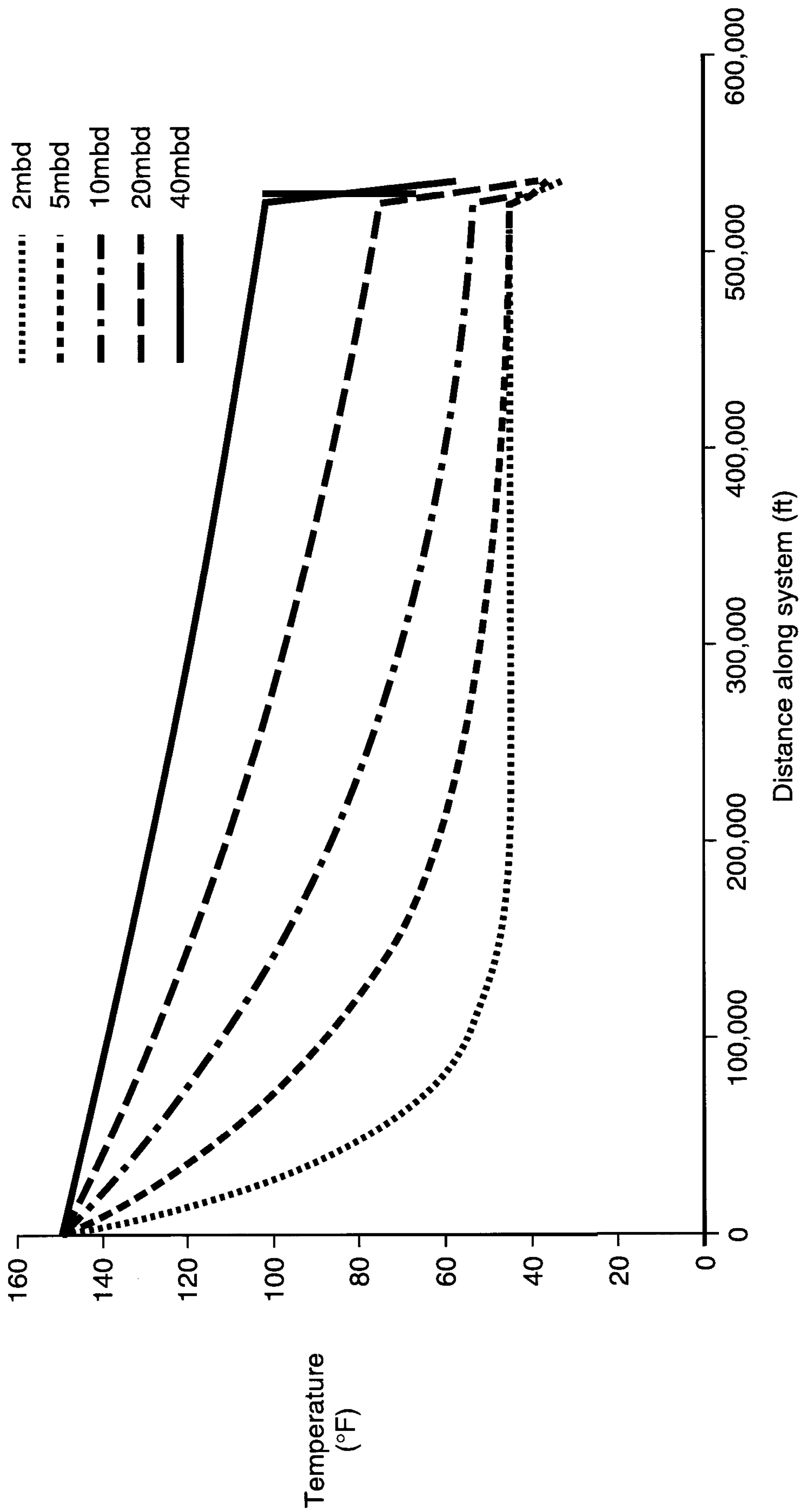
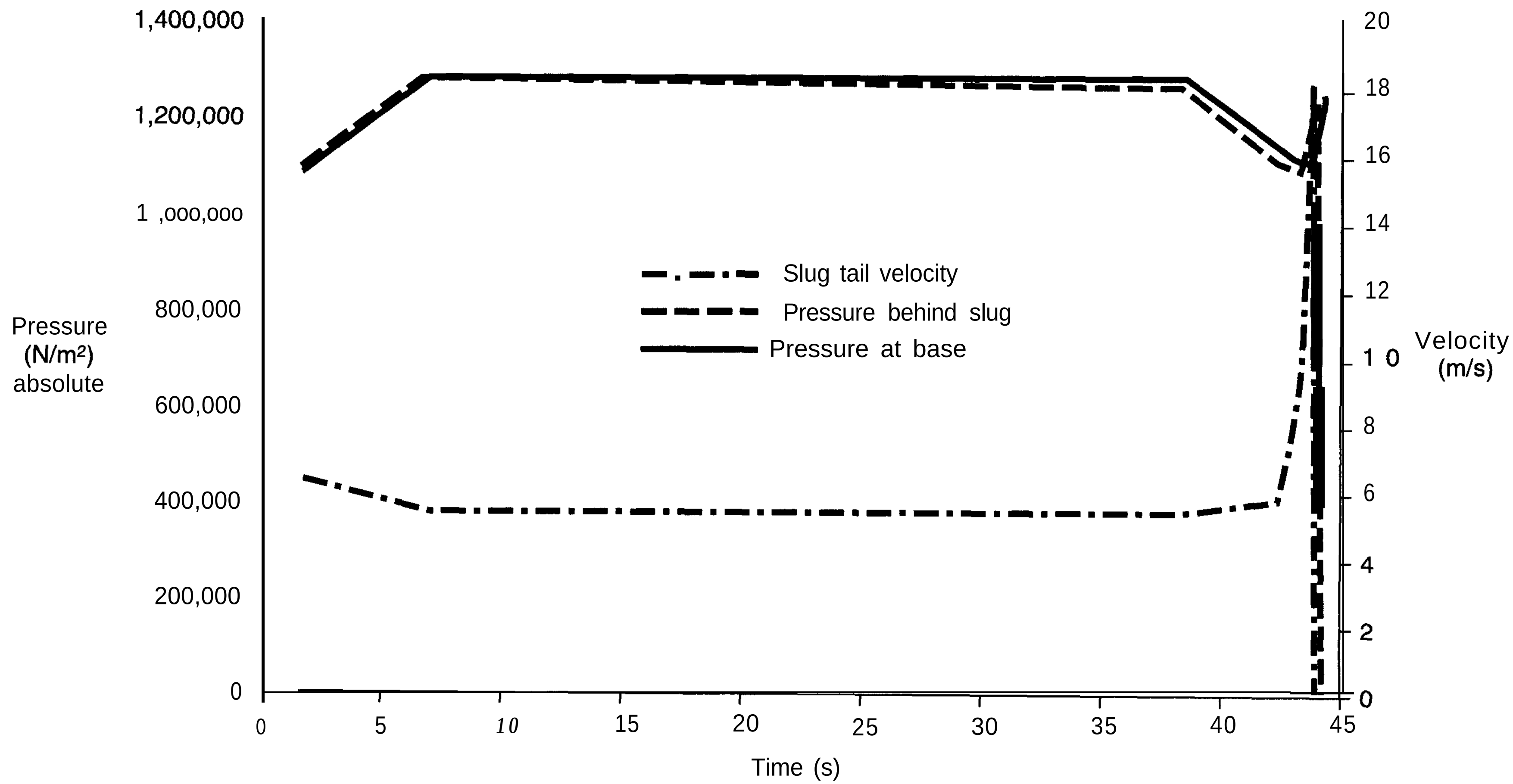


Figure 8. **Temperature profile with low overall heat transfer coefficient of 0.14 btu/hr ft² °F.**



Fi 9. Slug accelerations in a riser

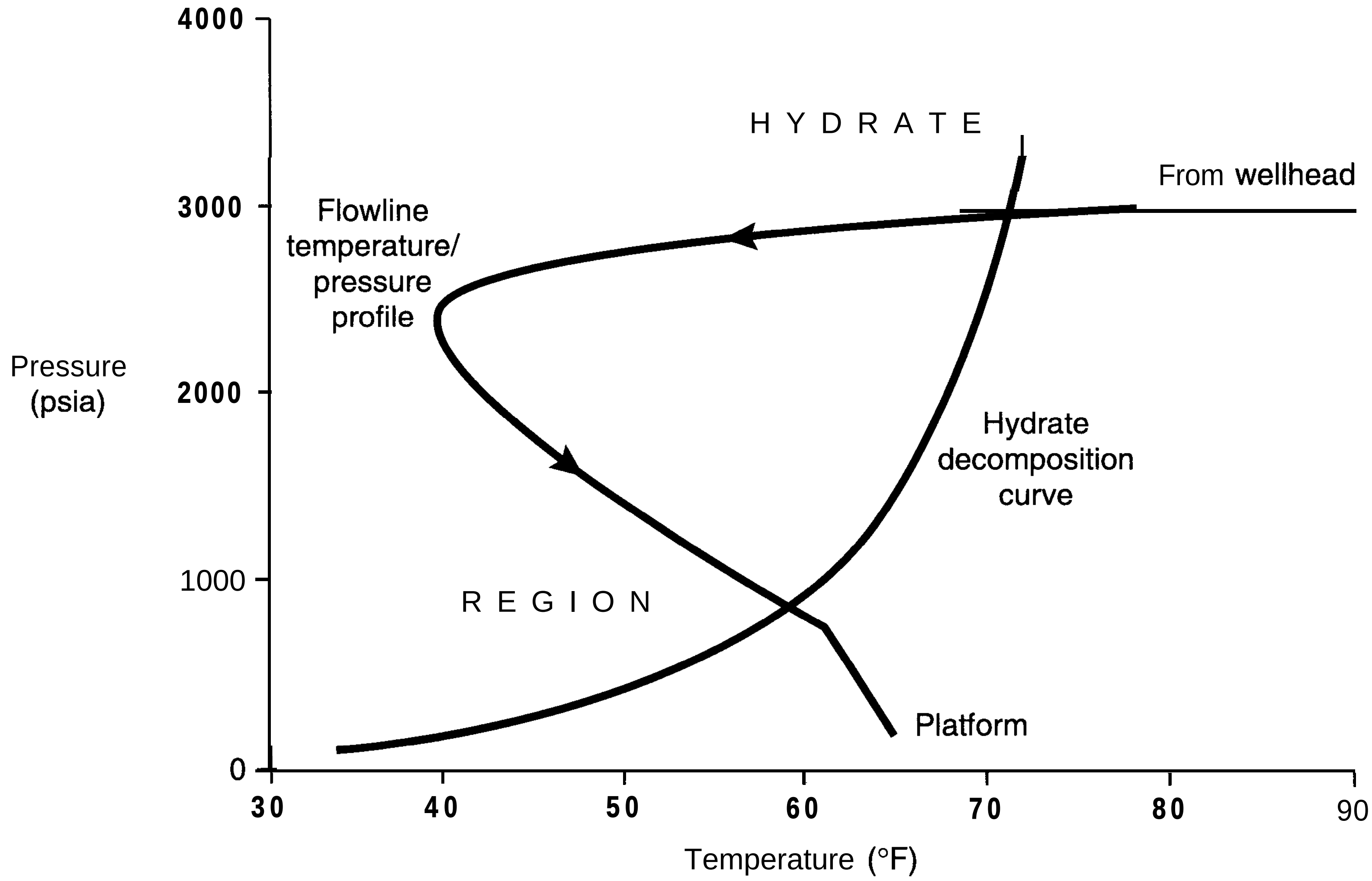
Limitation	Limitation	Solutions
Pipeline blockage due to cold temperatures	• Colder ambient temperatures and increasing heat loss affecting hydrate, wax and emulsion formation. • Colder temperatures due to auto-refrigeration	• Insulate flowlines • Use inhibitors • Add heat – bundle heating – electrical – chemical • Maintain high flowrates • Pig pipelines
Flow impeded by high viscocity	• Colder temperatures give rise to emulsion formation	• Use inhibitor • Keep hot
Flow assurance problems due to low flowrates. Corrosion due to water puddling. Corrosion under solid deposits.	• Flowlines operate at higher pressures and lower velocities allowing water to build up and sand to accumulate. • Higher corrosion rates at higher pressure	• Pig flowlines • Apply minimum flowrates criteria • Inhibit corrosion

Figure 10. Integrity related issues – Avoid impeding flow/contain fluids

Limitation	Limitation	Solutions
Corrosion / Erosion	• Not significantly worse at same steady outlet pressures. However, dynamic effects greater leading to accelerated veolcities could affect inhibitor stripping	• Exotic materials • Lower veocities • Change configuration
High mechanical loads	• Fluids accelerated to greater velocity due to hydrostatic/friction effects	• Limit/control velocites • Support pipework • Change density of slugs
High pipeline pressures	• Deep flowlines operate at high pressure difficult to design for	• Protection systems • Higher design pressure
High temperature	• WHFT higher	• Increase material temperature limits

Figure 11. integrity related issues – Avoid impeding flow/contain fluids

Figure 12. Hydrate decomposition / Flowline temperature-pressure profile



Without inhibition or insulation, production fluids in a 50-mile flow line from deepwa ters are in the hydrate region for most of the length of the flow line

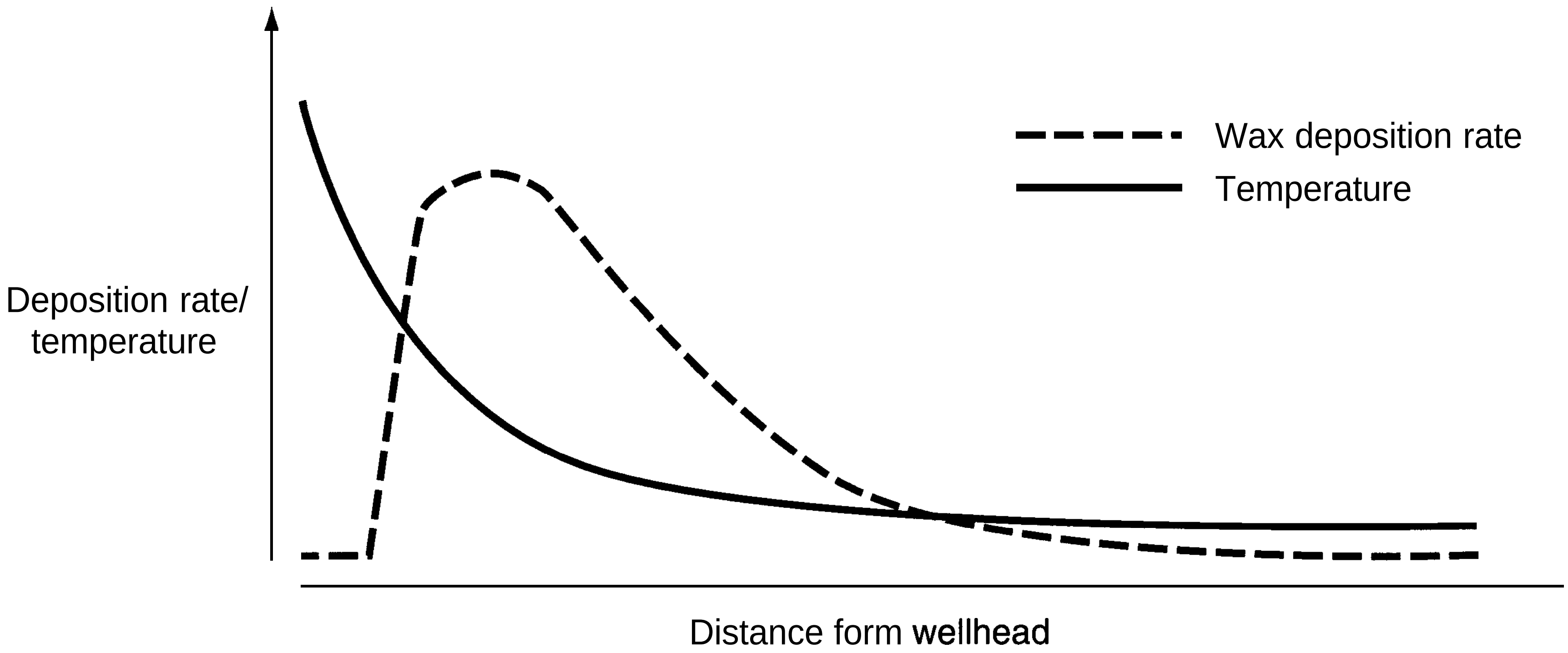


Figure 13. Subsea flowline wax deposition profile

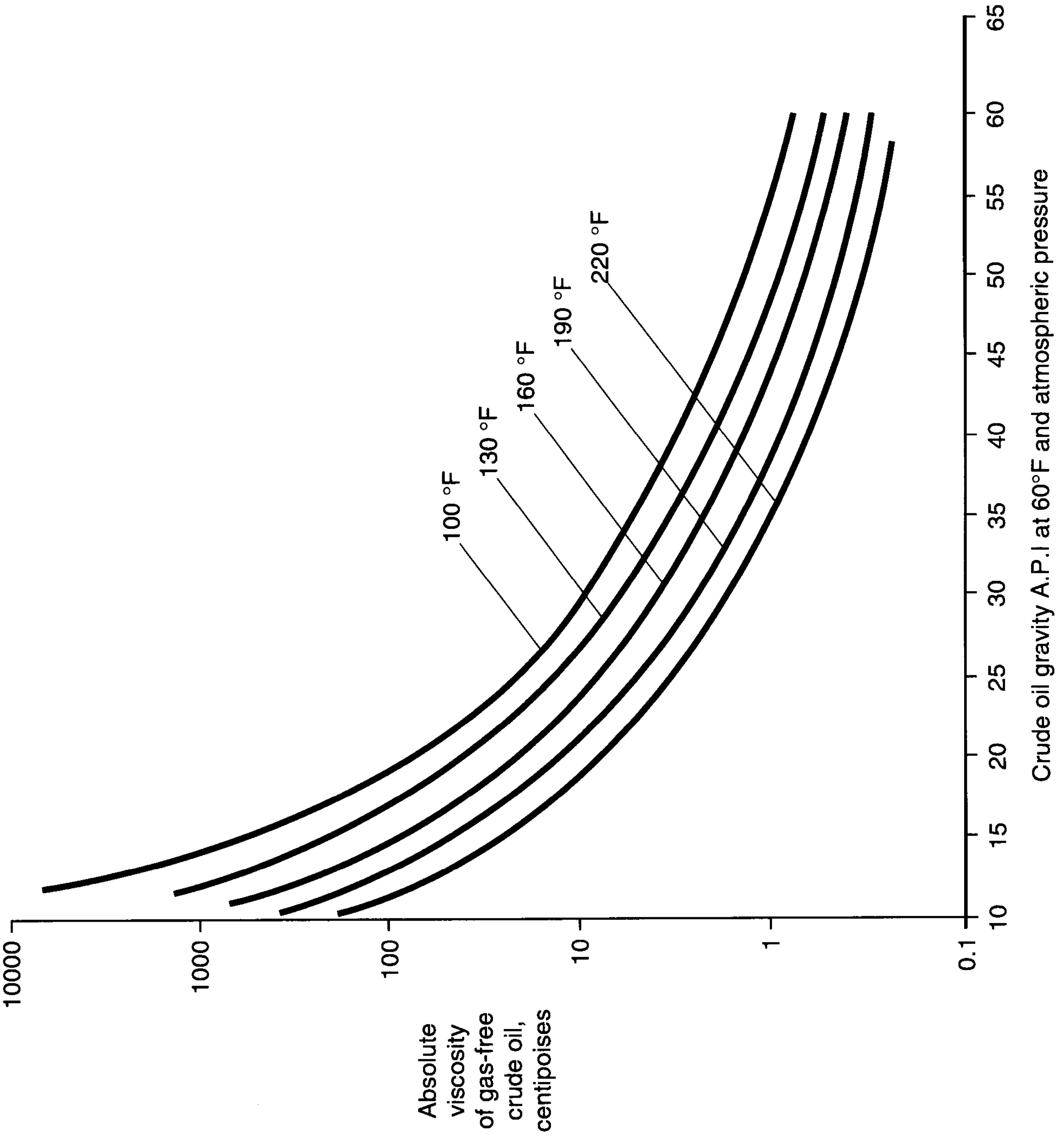


Figure 14. Crude oil viscosity as a function of API gravity and temperature.

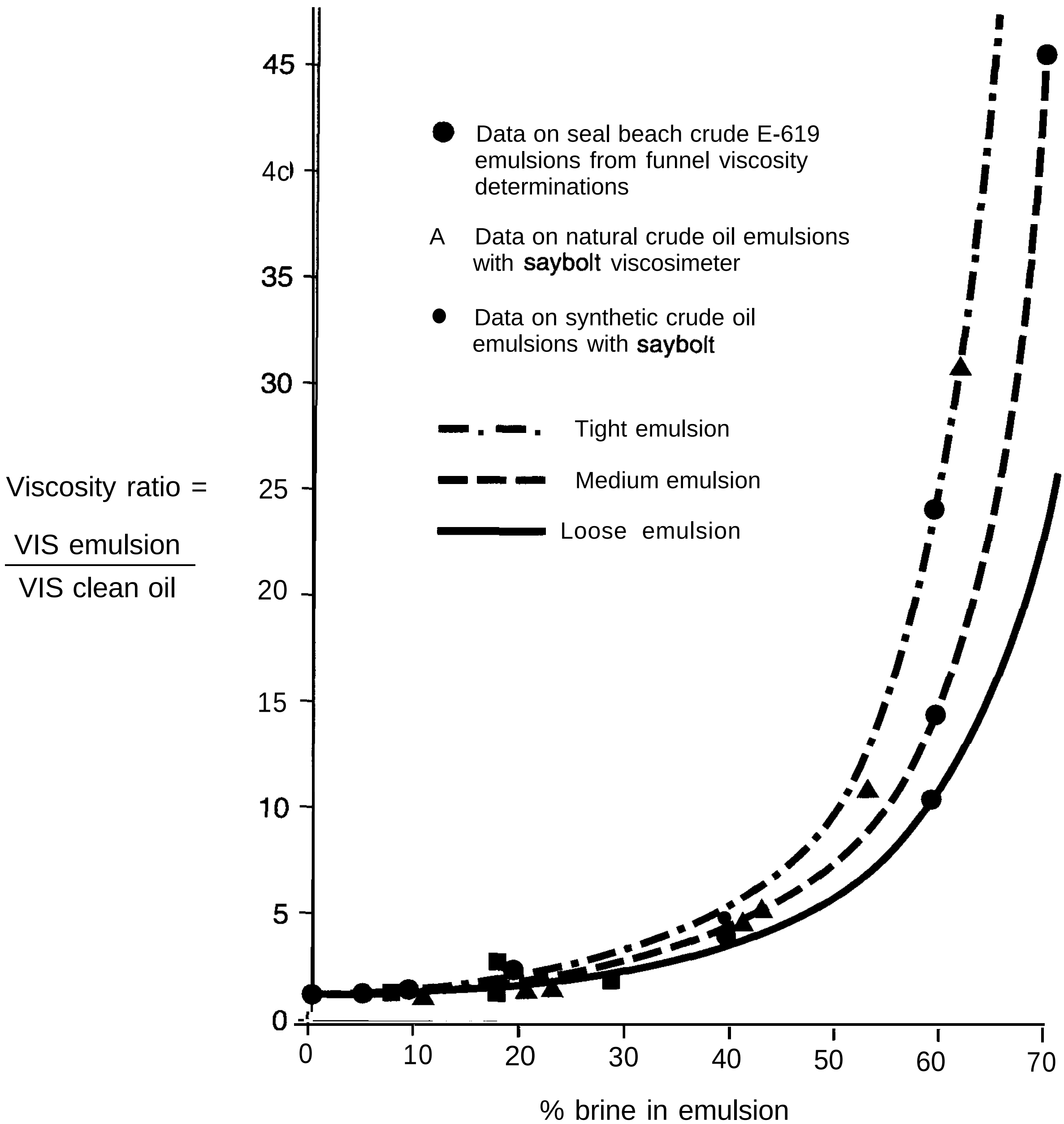


Figure 15. Variation in viscosity of crude oil and brine with brine content

Limitation	Impact of deepwater	Solutions
Flowrate variations	• Slug volumes increased. • Riser geometry can promote severe slugging, greater pressure and velocity swings. • Can lead to plant shutdown	• Change topography of flowline/riser • Use control system • Gas lift • Foam fluids • Operating procedures • Large slug catcher • Compressor control • Process plant dynamic operation
Low temperatures	• Cooler temperatures can give rise to emulsion formation and separation problems. • Problems with handling wax or hydrate increasing cost of treating and causing downtime	• Use inhibition • Insulate • Add heat • Maintain high flowrates

Figure 16. Delivery related issues – Deliver fluids in an acceptable manner

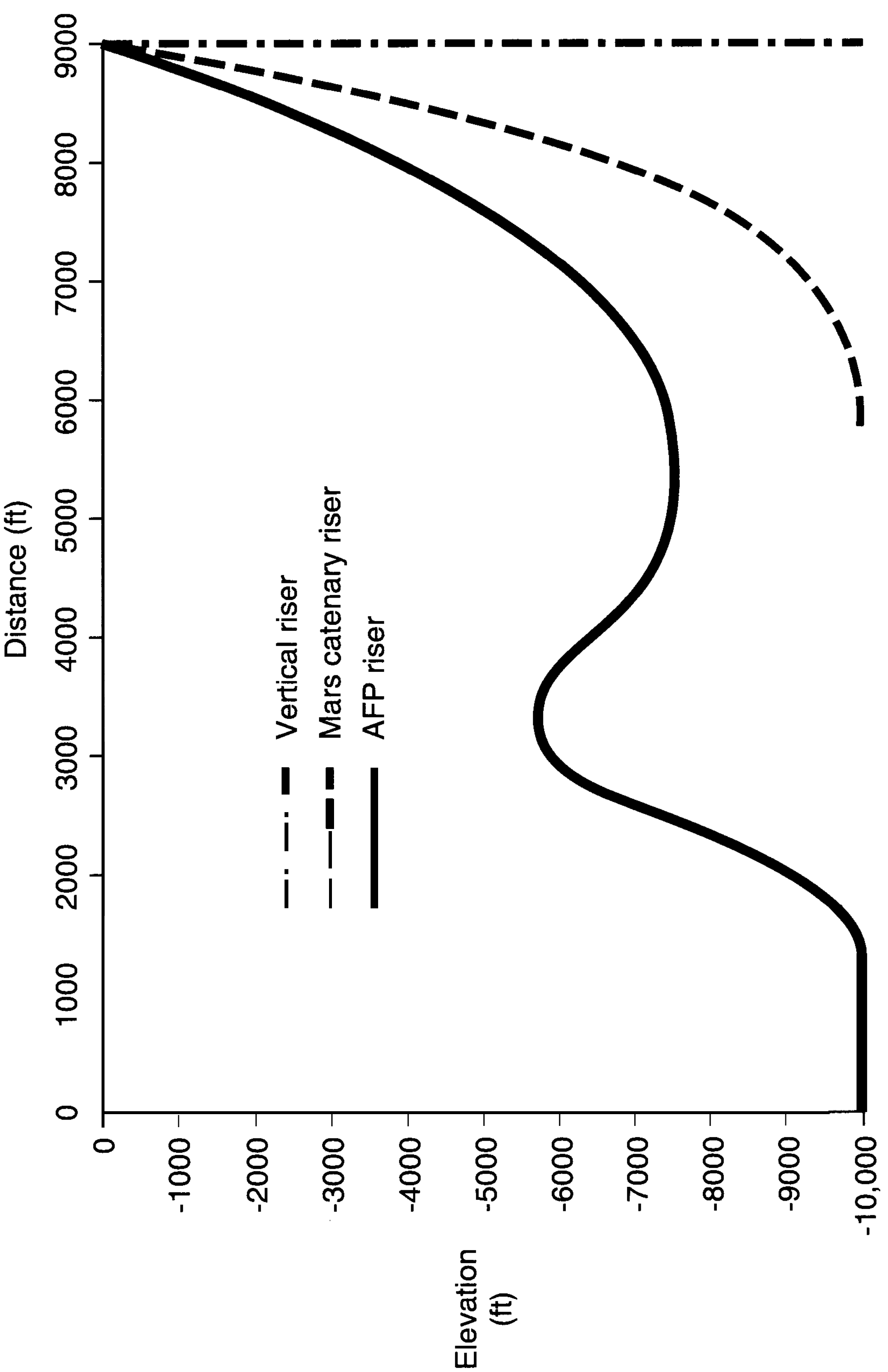


Figure 17. Riser shapes used in analysis

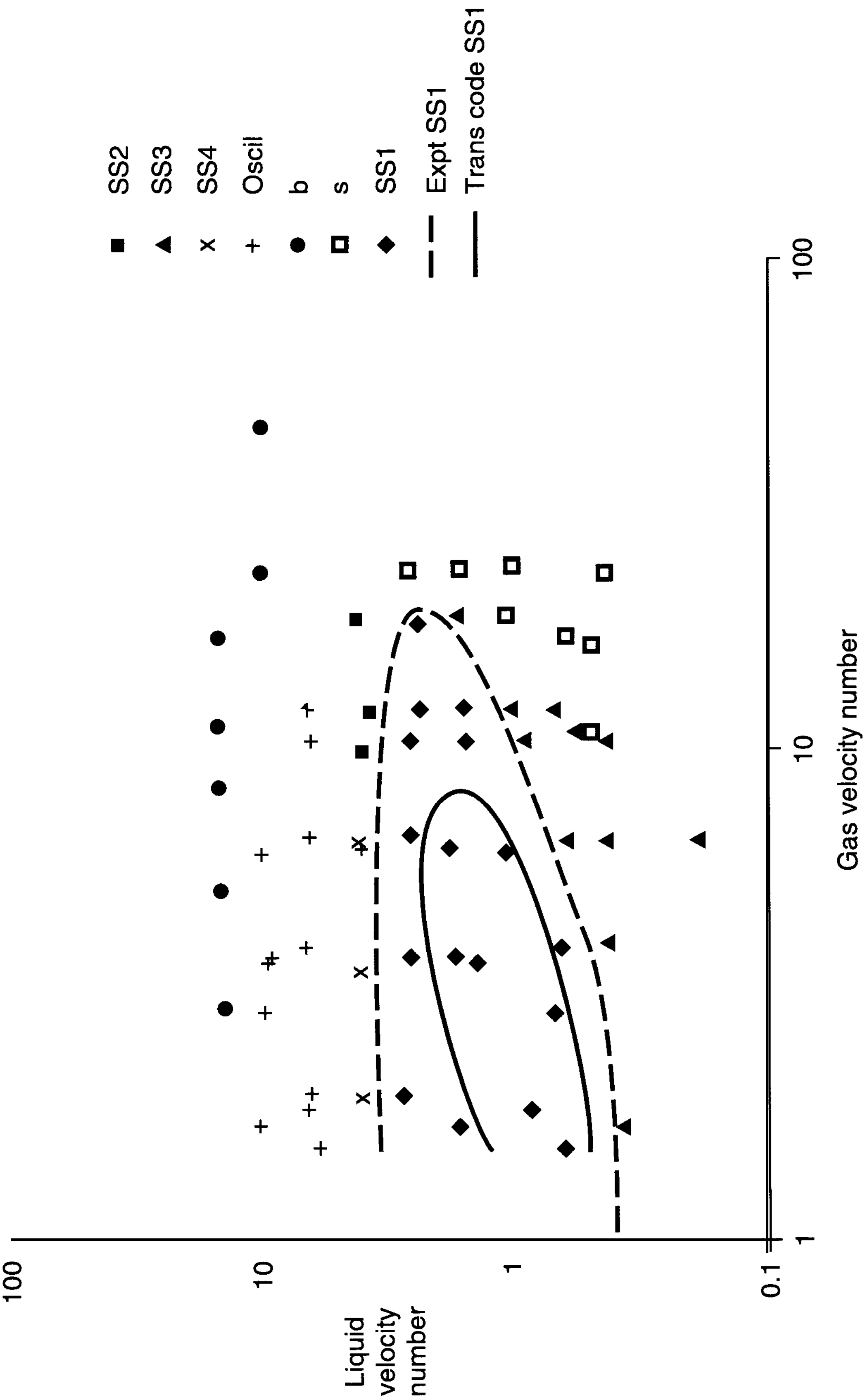


Figure 18. Lazy S riser, flow pattern map

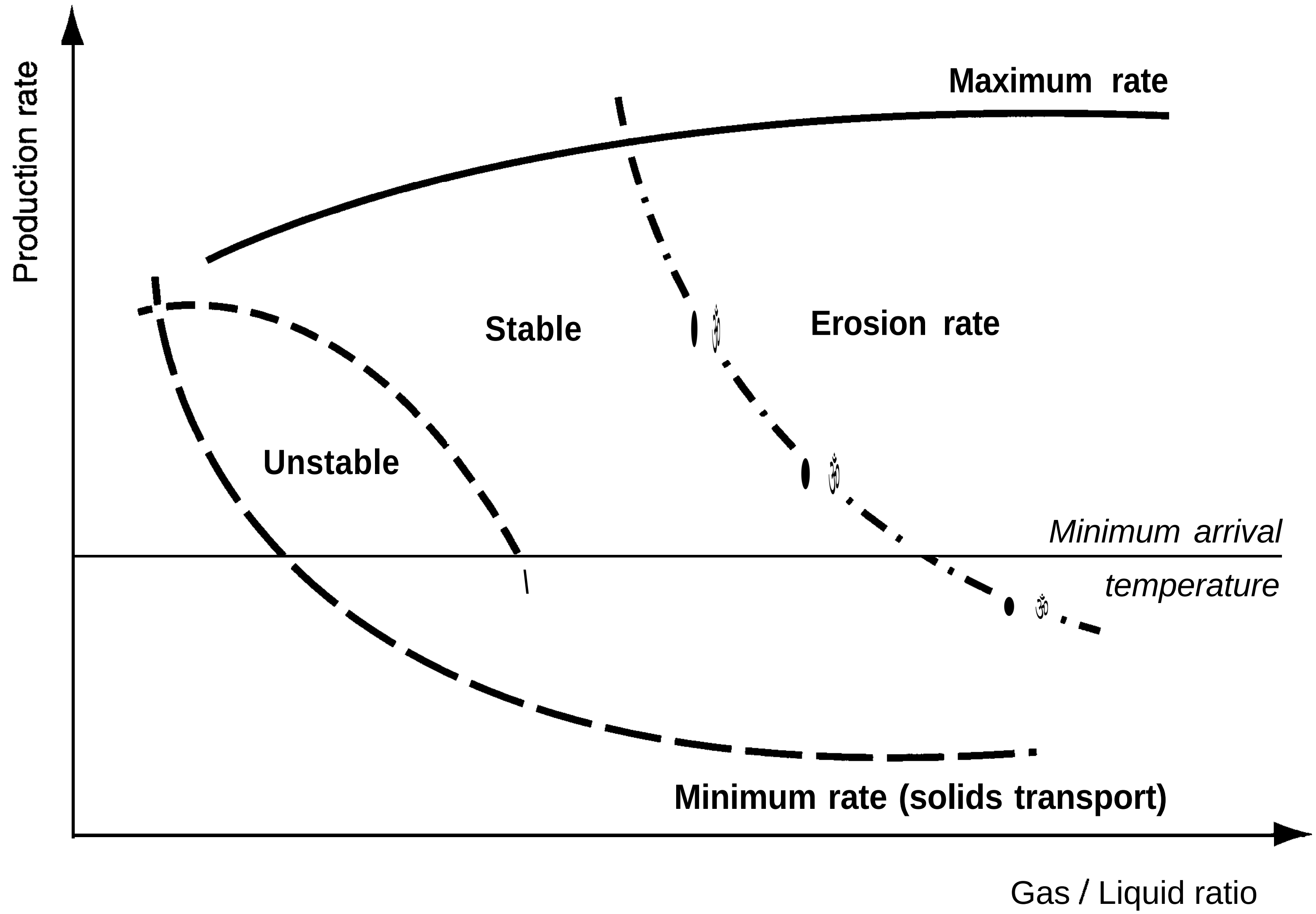
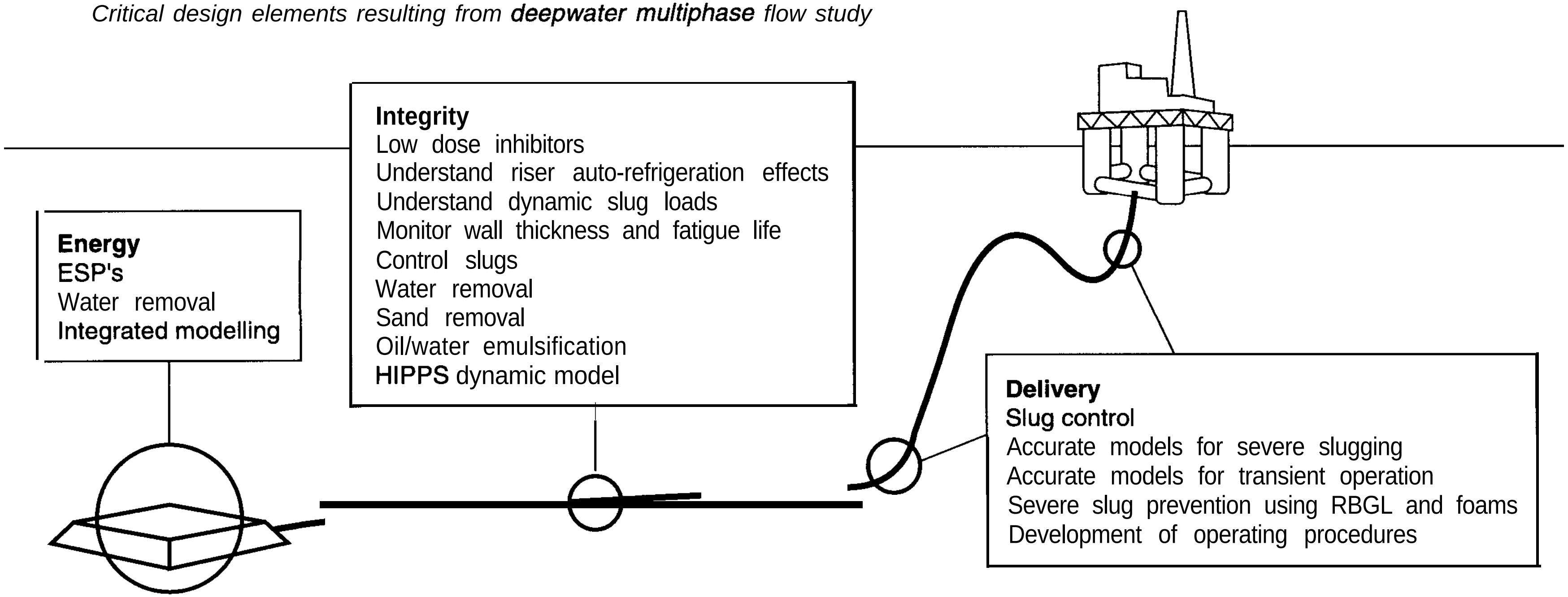


Figure 19. Foinaven operating envelope

Figure 20. Understanding the hydraulic limitations



- **1998 Foinaven Field trials will provide unique data for improved understanding and model calibration**

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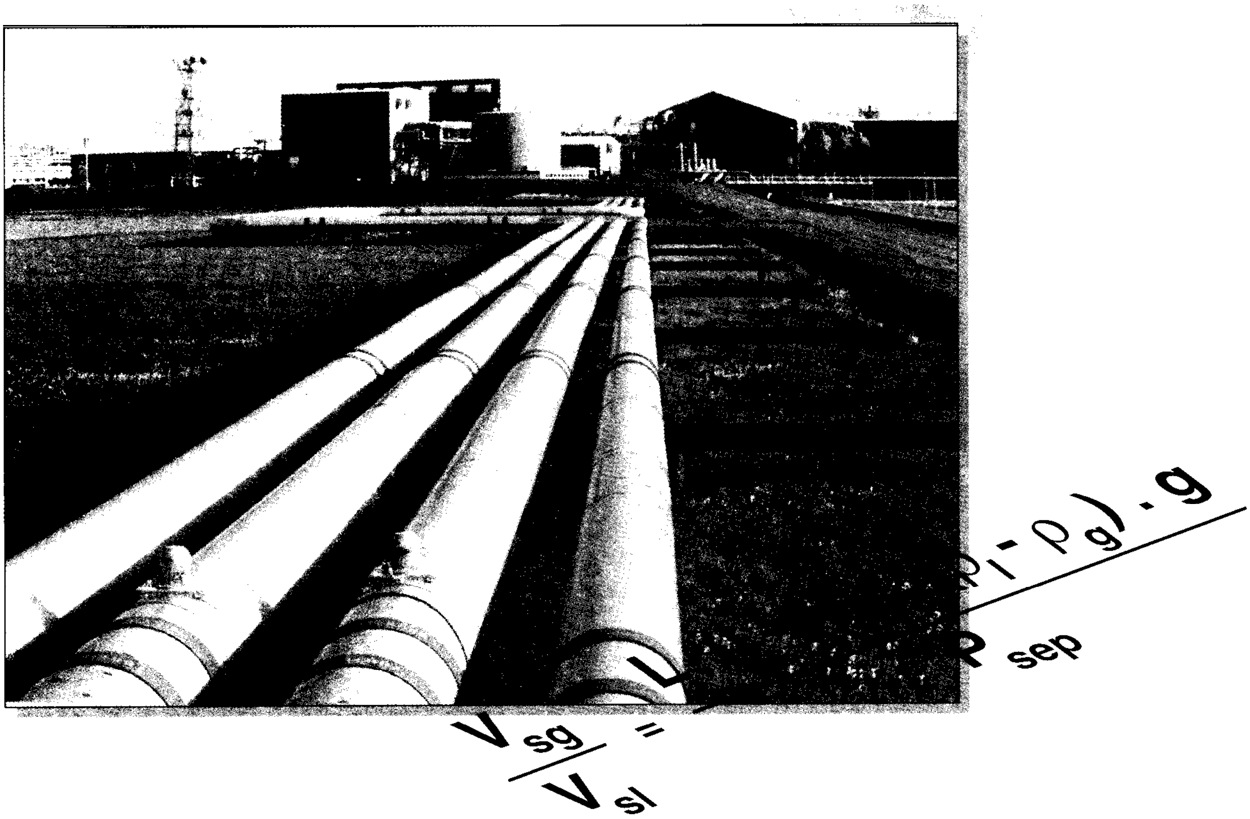
Section 21 Corrosion in Multiphase Flow



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21.1 Introduction

Corrosion in multiphase, carbon steel flowlines is not well understood and the underlying mechanisms are still under research. Fundamental CO, corrosion research is, at present, restricted to simple laboratory flow experiments with idealised fluids. In real field multiphase systems the problem is far greater due to the wide range of uncertainties, i.e. basic flow parameters, gas and crude property variations, effects of flowline geometry and the performance of corrosion inhibitors.

Assessment of corrosion rate and potential is an inexact science and its application requires considerable experience. This section of the Design Manual has been written only to highlight the awareness of corrosion to the Facilities Engineer. It must be emphasised that reference to a Corrosion Engineer should be made at the concept stage of any flowline or facilities design study.

Multiphase flow design schemes and the use of carbon steel flowlines present attractive capital cost options for field development. The flowline design team must assess these advantages against high operating costs for inhibition, corrosion monitoring and line repair during field life.

A multiphase, carbon steel flowline is always subject to a corrosion risk. In a design, an assessment of that risk is required to see whether corrosion may be controlled through a corrosion allowance and inhibition. These specialised tasks combine stress analysis and corrosion engineering specialists to examine pipe integrity requirements, corrosion allowance and fluid corrosivity, Refs.1-4 provide more details.

This chapter examines the multiphase flow effects on corrosion and inhibitor performance and where they are important in a flowline design.

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21.2 Corrosivity

CO_2 corrosion is caused by a combination of the CO_2 content in the gas and the presence of produced free water. The corrosivity of the fluid is affected by pH, partial pressure of the gas, temperature, H_2S content, etc. Simple models are available to assess the corrosivity of a fluid under flowing conditions, see Ref 5. This is a specialist task for a corrosion engineer as value judgements need to be made based on experience.

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21.3 Corrosion Inhibitors

Corrosion inhibitors are long-chain polymers that reduce the corrosion rate by adsorbing onto the flowline wall. They are high cost chemicals that fall mainly into either the imidazoline or polyamine chemical groups. Actual oil field inhibitors contain many additional components, such as sutfactants, in a proprietary formulation. Usually an inhibitor is designed to be soluble in either the water or oil liquid phase and dispersible in the other phase.

The physical processes governing the distribution of the inhibitor, and its adsorbtion onto the metal surface, are not well understood, especially in multiphase flow. It is known that the time required for successful inhibitor adsorption may be counted in minutes, whereas, the timescale of turbulence activity near the pipe wall surface, e.g. in slug flow, is measured in milliseconds. This timescale mismatch and the intermittent nature of slug flow are considered to be important factors affecting the corrosion rate.

The design dosing rates are quoted in parts per million, typically in the range 10-50ppm, which requires a known solvent liquid phase flowrate to calculate the inhibitor injection rate. Inevitably the phase flowrates change so that, unless care is taken, widely differing dosages are delivered than those required by the design. In addition, operational problems such as lack of maintenance of the injection equipment can severely affect the delivered dosages; episodes where the inhibitor supply tank has run dry are not uncommon. Some success in reducing the corrosion rates to manageable rates, typically 0.05 mm/yr, in inter-field multiphase lines in Prudhoe Bay, has been attained by increasing the dosing rates to 100 - 200ppm.

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21.4 Database

A database of corrosion rate and flow conditions is available on PC, Ref.6. The system is known as the Corrosion Rate Assessment dataBase System (CRABS) version2 consisting of Excel data worksheets containing empirical and erosion rate data and details of the fluid and chemical conditions under which they were measured.

Information on both field and laboratory measurements is contained in the database and may be used to:

- **Establish representative metal loss rates for use in pipeline design**
- **Investigate for correlation between corrosion and the fluid and chemical condition recorded**
- **Generate data to aid in the validation of computer models**

The database is maintained by the Materials and Inspection Group, Sunbury.

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21.5 General Flow Effects

Multiphase flow regimes in flowline systems have been discussed in Chapters 1&3. If the fluid phases are corrosive, under flowing conditions, the resulting flow regimes can affect the corrosion rate by varying the corrosion inhibitor efficiency, Ref.7. Generally, stratified flows in horizontal flowlines do not present a problem, if adequately inhibited. However, at very low liquid velocities with little or very poor inhibition, considerable damage can occur. Typical damage resembles "tramlines", lines of localised pitting at the liquid gas interface, located at the bottom of the flowline. Sand or solids may remain at the bottom of the flowline and promote under-deposit corrosion. Due to geometrical or terrain changes, the stratified flow may change and alter its characteristics radically, this is discussed in the next section.

The slug flow regime, in a horizontal flowline, may have a very large effect on inhibitor performance. During the passage of the slug, gas is entrained into the highly turbulent slug front and, flow visualisation suggests, is projected downwards to the bottom of the flowline in bursts of small bubbles, Ref.8. High shear stress, the possibility of gas bubble collapse and high turbulence are all possible contributors to reducing the performance of the corrosion inhibitor; current investigations will assist in determining governing mechanisms. Rapid operational transients, e.g. compressor trips or the bringing on of high gas rate wells, may produce very large slugs.

The annular flow regime has a high velocity gas core which entrains and deposits liquid droplets from and to the surrounding liquid annulus. In a horizontal flowline case the liquid annulus is much thicker at bottom of the flowline due to the effects of gravity. There is no evidence that the impact of liquid droplets on flowline walls, alone, causes problems. However, combining sand transport within the high velocity gas core, and a corrosive medium, can produce very rapid rates of erosion/corrosion, Ref.9.

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21.6 Localised Flow Effects

Flow regimes operating in straight, horizontal flowlines may give rise to corrosion problems in the absence of other influences. However, it is much more likely that the first signs of damage will occur at localised flowline features. These would be where details in flow structure and geometry produce potential corrosion "hot spots". Likely areas would be at major geometrical changes, e.g. bends and flowline fittings, tees & wyes. However, minor changes, e.g. weld beads, gaps at flanges and previous mechanical or corrosion/erosion damage, can also produce serious corrosion problems.

Major geometrical changes to the direction of a flowline increase turbulence production, local wall shear stress and re-distribute the gas and liquid (also solids, if present) phases. The degree to which this occurs depends upon the acuteness of the direction change, the change in flow area, especially a constriction, and the proximity of two or more changes. Vertical bends, in nominally stratified flow may create very high local turbulence activity as the geometry change serves to collect the liquid phases, bridging the pipe before being blown through by faster moving gas. Examples of this occurring in a 30" dia. flowline, in nominally stratified flow, and 40% watercut, created wall thickness losses of 80% and corrosion rates in excess of 500 mpy (12.5 mm/yr). Hilly terrain with long upward slopes may produce slugs, which are dissipated in the downhill sections. Here, again, an analysis of the flowrates without taking pipeline angles into account may predict a stratified flow regime, whereas slugs will have formed somewhere in the system.

Minor changes and details to pipewall sections may create local areas of increased inhibitor breakdown due to the abruptness of the change and water retention. Abrupt wall mismatches due to thick walled fittings or field welds of flowline stock manufactured from rolled plate, and the metallurgical changes due to the heat affected zone, are prime targets for water dropout, inhibitor breakdown and accelerated corrosion. Gaps in flange seals for large diameter flowlines may also promote water dropout so that a nominally low watercut of less than 10% could locally be subject to, say, 50% or more. Particular care is required if sections of dissimilar materials are flanged together, e.g. stainless and carbon steels, due to galvanic corrosion, again reference to a corrosion engineering specialist should be sought.

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21.7 Flow Modelling for Corrosion Design

There are no direct methods for predicting enhanced corrosion rates based on the action of multiphase flows. The corrosivity of the fluids can be predicted by a corrosion specialist. Inhibitor integrity under flowing conditions, however, is based on field experience and, mainly, laboratory trials, Models, e.g. PROSPER and HILLYFLO, may be used to determine local flow conditions, e.g. velocities, flow regimes and wall shear stresses but the corrosion potential is a value judgement. For example, in modelling work the shear stresses generated by a slug are estimated by treating it as a piston of liquid travelling at about the gas velocity. This is an over-simplification of the physical processes but allows a degree of quantification and a reference for experienced corrosion specialists to base a judgement.

The steady-state software described in Chapter 3 may be used to determine the local flow conditions and fluid properties, taking into account simple geometrical changes, e.g. slug characteristics in long upward and downward slopes. A quantitative assessment of the damage potential of the slug cannot currently be determined.

Where the geometrical changes are more complicated, such as flow through a tee, or at a greater level of detail, flow in a corrosion pit, the three-dimensional nature of the flow becomes important. In this instance, specialist computational fluid dynamics (CFD) software has to be used to predict velocity, pressure, turbulence and temperature distributions. However, again, the corrosion potential is a value judgement.

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21.8 Research into Corrosion in Multiphase Slug Flow

Research at Ohio University by Dr W.P. Jepson and co-workers has shown that the slug film Froude number may be used to determine both the severity of the slug and its corrosion potential in the presence of CO,. This technique is still under investigation and is yet to be applied fully on an operational problem. Froude number analysis of slugs has been based purely on pipelines, i.e. smooth, constant diameter pipes in the absence of localised areas of turbulence caused by intrusive fittings or junctions.

The most likely effect of an intrusion on the corrosion rate and its possible relationship to the Froude number is shown in the figure below; the shapes of the curves are only approximate and are not based on highly detailed analyses.

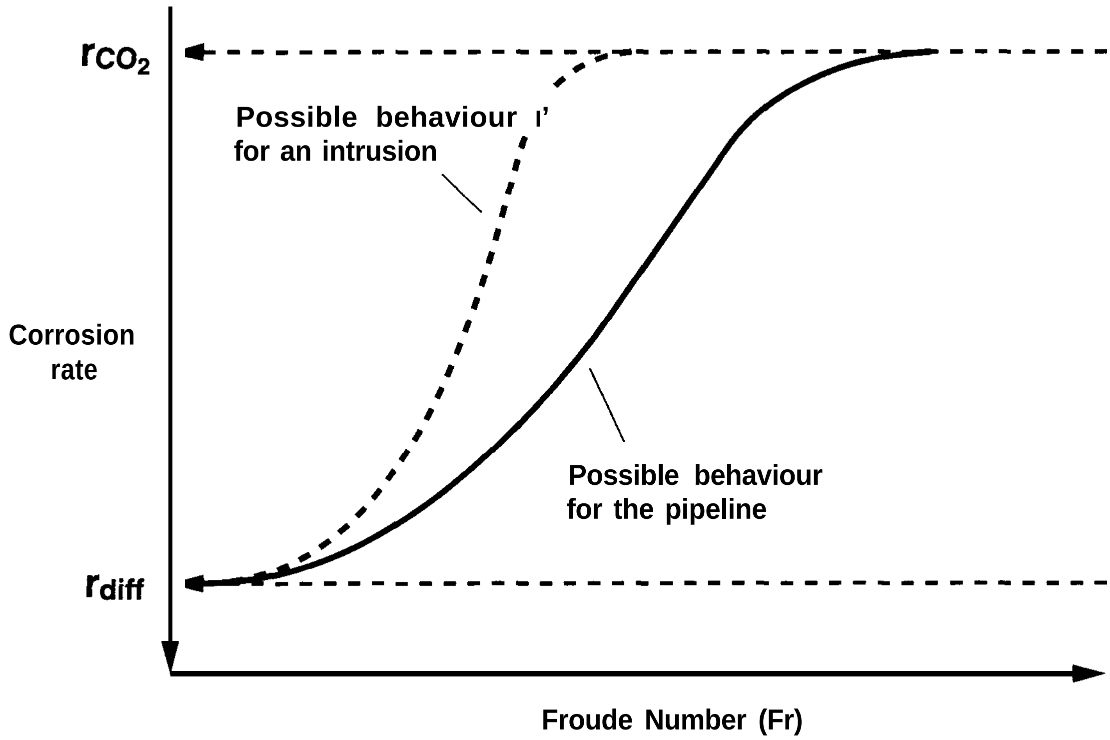

Figure1 . Variation of corrosion rate with Froude Number

The rate of corrosion is shown on the y-axis of the trend chart and demonstrates the two extremes of corrosion. In diffusion controlled corrosion the rate of diffusion controls the rate of reaction, r_{diff} and corrosion occurs as CO, attack on the metal surface. However, the reaction is limited by the mass transfer to and from the surface, i.e. the rate of diffusion through the stagnant and laminar boundary layers is slow, in comparison to the reaction rate.

In reaction controlled corrosion the rate of corrosion is the rate at which the reaction takes place, r_{CO_2}. This corrosion occurs as if the metal were placed in a beaker and stirred vigorously, i.e. the turbulence generated by the flow completely dominates the reaction rate.

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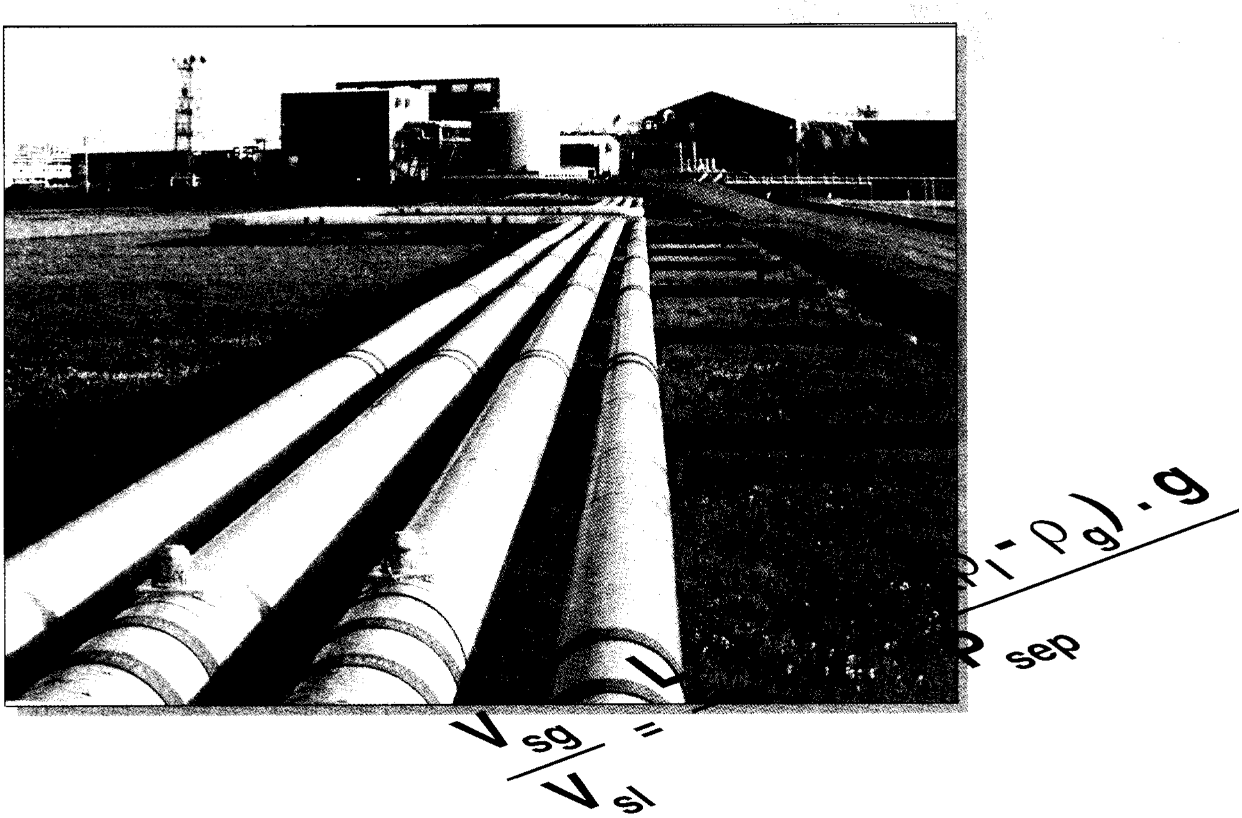
Section 22 Wellhead and Pipeline Closed-h Pressures



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22.1.1 Introduction

Maximum WHCIP's have traditionally been calculated using one of the following two models:

1. Assume wellbore contains dry gas only.

WHCIP is then the bottom hole pressure (BHP), which is assumed to rise to the reservoir pressure, minus the hydrostatic head of gas. This gives rise to a worse case maximum possible WHCIP. For wells that produce large fractions of liquid, this model will lead to unduly conservative predications of maximum WHCIP.

For systems where the tubing may consist predominately of gas when the well is shut-in this model must be used.

2. The Classical Equilibrium Model.

This assumes equilibrium conditions between the oil and gas are immediately reached as soon as the well is shut-in. The oil settles to the bottom of the tubing. The pressure at the bottom of the oil column then rises to the reservoir pressure. The pressure in the oil column reduces with height due to hydrostatic head effects. When the pressure in the oil column drops to the bubble point, a sharp interface between oil and gas is assumed to exist. Above the interface a dry gas column is assumed to be present. The hydrostatic head loss over the **gas** column gives the difference in pressure between the bubble point pressure (at the bottom of the gas column) and the WHCIP.

Applying this equilibrium model to data from Magnus Platform wells gave rise to an under prediction of WHCIP, i.e. the WHCIPs reached during well shut-in exceeded the values calculated using this model. It was believed that the reason why this model under predicts the WHCIP is that on shutting in the well much of the gas present (under relatively low pressure flowing conditions) did not re-dissolve in to the oil. Thus thermodynamic equilibrium did not exist for the oil at the relatively high pressure conditions produced as the bottom hole pressure rises up to the reservoir pressure. Hence, for the shut-in condition, a volume of gas in excess of the equilibrium quantity is left in the tubing. The hydrostatic head of fluid in the tubing is then less than would be determined using equilibrium assumptions and hence the WHCIP is higher than predicted using this model.

A new model has been developed to account for this non-equilibrium affect and is discussed below.

The two models recommended for determining WHCIP are discussed in detail below:

- **The first is for systems where a dry gas column may exist in the tubing on shut-in. (Section 22.1.2)**

- **The second is for systems containing liquid and accounts for non-equilibrium conditions that can exist in the tubing following shut-in. (Section 22.1.3)**

22.1.2 WHCIP Calculations for Wells that may contain Gas only.

For any system where a column of gas may exist at shut-in this approach must be used to predict the worse case maximum WHCIP. This applies to gas, gas-condensate reservoirs, and also oil reservoirs which contain a gas cap. Determination of head loss over a gas column is complicated by the change in gas density, and hence hydrostatic pressure gradient with height. The simple equation traditionally used to compute hydrostatic head loss over a column is:

$$\frac{P_{top}}{P_{bottom}} = -\exp\left(\frac{Mw.g.H}{ZRT}\right)$$

where:

P_{top} = pressure at the top of the gas column (bara)
P_{bottom} = pressure at the bottom of the column (bara)
Mw = molecular weight of the gas
g = 9.81 ms^{-2}
H = height of the gas column (m)
Z = average gas compressibility
R = 8314 J K-l $Kmol^{-1}$
T = average temperature (K)

This equation however does not account for the change in compressibility (Z) with pressure (i.e. it requires input of one fixed value of Z). In order to correctly account for the change in Z, and hence in gas density with height and pressure, the following procedure is required:

- **Set up a model of the tubing.**
- **Ensure that the homogeneous correlation will be used for vertical upflow.**
- **Produce compositional physical properties for the fluids.**

Run the simulation at a very low rate, eg 0.1 MMSCFD, so that the frictional pressure drop is minimal (less then 2 psi). The homogenous correlation is employed so that no slippage occurs between gas and liquid. The simulation should be run with a fixed temperature throughout the tubing. This should be set to the reservoir temperature. The inlet pressure (i.e. the bottomhole pressure, BHP) is set equal to the maximum reservoir pressure. By using a simulation to compute hydrostatic head loss, the change in gas density, and compressibility, with height is correctly accounted for. The calculated outlet pressure (WHP) is a first pass estimate of maximum WHCIP.

If liquid is predicted to be present in the tubing during this calculation there is a danger that the hydrostatic head loss over the tubing will have been over estimated and hence the WHCIP under-predicted. In practice, any liquid present will drop to the bottom and may re-enter the reservoir.

To ensure that the predicted WHCIP is conservative, the initial fluid composition used should be flashed at the WHP conditions to remove all liquid that could exist in the well. The remaining lighter gas from this flash should then be used to produce a second compositional datapack.

Re-running the simulation with this new modified compositional datapack will provide a conservative estimate of maximum WHCIP.

22.1.3 WHCIP Calculations for Multiphase Wells.

This method is applicable for oil reservoirs where there is no danger of the liquid present in the tubing at shut-in flowing back into the formation to leave a gas column only. This method, known as the MPF method, is discussed below:

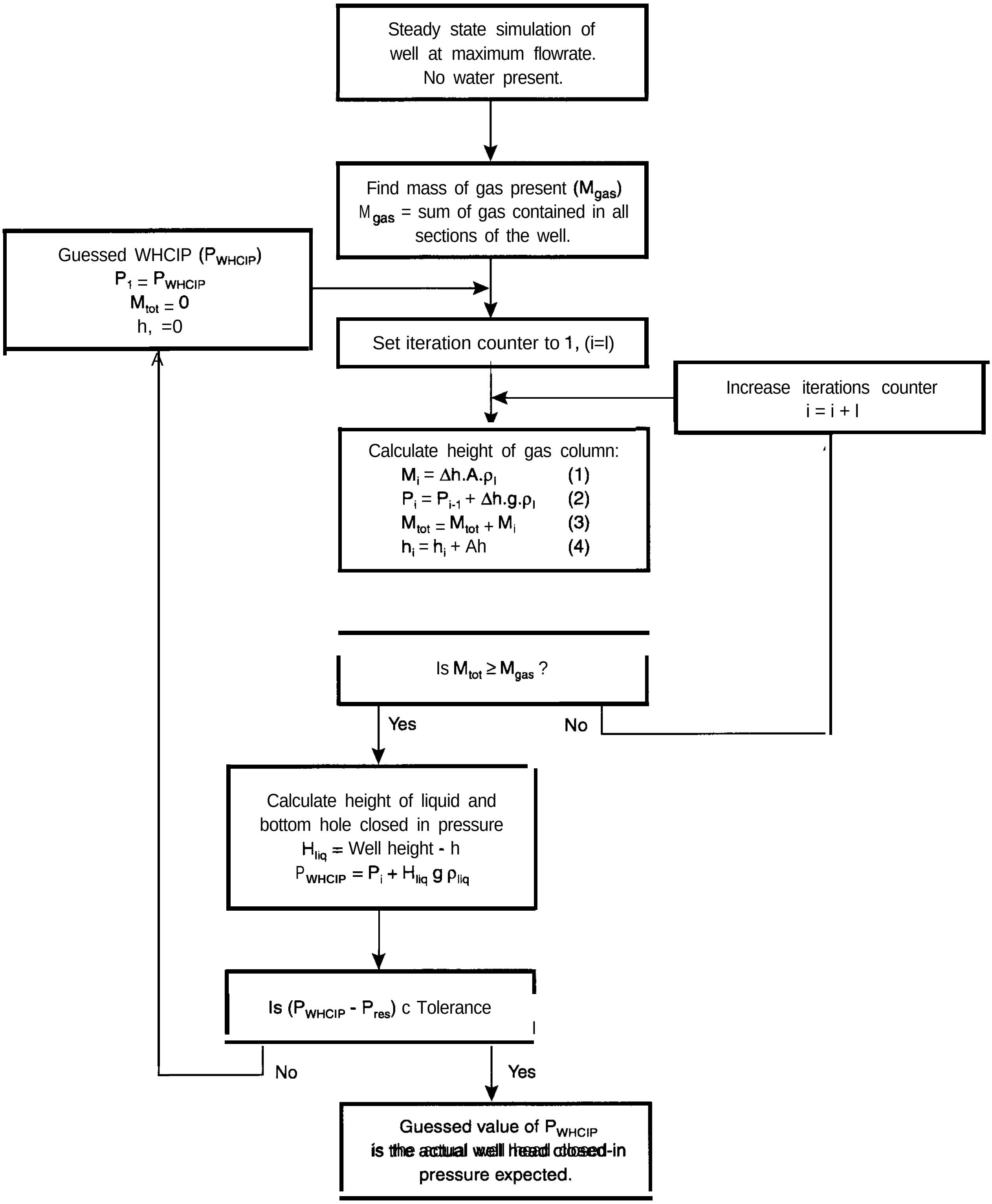
The calculation begins by simulating the well under maximum flow rate conditions and the highest temperature gradient, to determine the maximum gas content of the well. The pressure drop method used should be the one that most closely matches actual data for the field in question. If no data is available, then a pressure drop method likely to overpredict the AP should be used to ensure that a low wellhead flowing pressure is predicted, so that the gas content of the well is overpredicted. The flowing conditions used in this initial simulation should have a zero water cut in order to maximise gas content. The mass of gas present in the well bore under these maximum flowrate conditions is calculated. This data is not directly available in most flow simulation packages. The mass of gas present can be calculated by considering the liquid hold-up in each section and taking an average gas density for that segment. In order to simplify the calculation a spreadsheet model has been developed. This model is available from SPR Transportation. The model is based around a non-equilibrium concept where the gas in contact with liquid at steady state flowing conditions does not have time to equalise with the oil falling back down the well. This leads to the fundamental assumption of the model – that when the well is shut-in the mass of gas present remains the same as that calculated under flowing conditions. This is a conservative assumption as in practice some of the gas will re-dissolve in the oil as the pressure rises. The bottom hole pressure is assumed to rise to the reservoir pressure. The mass of gas settles at the top of the well bore. Below this gas column is the oil column.

A WHCIP must be guessed and then the pressure due to the hydrostatic head of the gas can be calculated. (Integration is used to eliminate undue errors) This is the pressure at the interface between the oil and gas.

This pressure now becomes the starting point to calculate the bottom hole pressure at the bottom of the oil column. This is done by assuming a constant oil density, determined by the reservoir conditions. A check must then be made between the calculated bottom hole pressure and the reservoir pressure. If they don't match then the chosen WHCIP is wrong and must be recalculated.

A simplified flow diagram is shown overleaf.

22.1.4 The Algorithm to Calculate WHCIP for Multiphase Wells



22.2 Pipe Shut-In Pressure with Upstream Inflow

22.2.1 Introduction

When flow lines are shut in at the downstream end it is not always possible or desirable to stop production into the upstream end of the line. For example line packing in gas export systems, and lines with downstream pressure relief systems.

It is therefore often necessary to simply shut off the exit to the pipe for some operational reason, such as a plant trip. The pipeline pressure is therefore allowed to rise as more fluid enters the pipe.

This can be simply represented by treating the pipeline as a tank in which liquid accumulates and reduces the volume. This results in an increase of the pressure in the pipeline, as well as an increase in liquid volume.

Below is outlined a method in which this localised increase in pressure can be calculated.

Consideration must be given to the resulting change back to flowing equilibrium conditions when the pipe is restarted as transient events will occur, often resulting in larger than normal liquid slugs.

It is important to note here that to fully understand the phase distribution that can exist in a pipe when it is shut-in, a full transient code such as PLAC or OLGA should be used. These codes also have the advantage of fully modelling thermal and hydrodynamic effects as well as of being able to simulate the restart of a pipeline. However, these codes require a large quantity of input data and take a considerable time to run.

22.2.2 Pressure Increase due to Outlet Closure

A spreadsheet program based in Excel has been developed to help in the above calculation as an alternative to full transient modelling.

This spreadsheet is called Pipeshut and is available through SPR Transportation.

Pipeshut requires the following information:-

1. Pipeline dimensions (ID and length).

2. Densities of the gas and liquid phases under flowing conditions.

3. The flowrates of the gas, oil and water.

4. The initial temperature, pressure and holdup of the pipeline under normal flowing conditions.

5. Correlations for calculation of Rs, Bo and Ov.

The following assumptions are made in the calculation:

1. **Pipeshut does not model any heat transfer and assumes that the pipeline remains at a constant temperature.**

2. **The flowrates of oil, water and gas into the pipe are assumed constant and therefore unaffected by the increase in pressure. The amount of gas dissolved in oil is however considered.**

3. **Ideal gas is assumed throughout.**

The calculation procedure is as follows:

1. **Calculate the mass of gas and liquid in the pipeline from the flowing conditions and holdup, at the time of initial shut-in.**

2. **Calculate the velocities and densities of the flowing gas and liquid phases, using the Rs, Bo and Ov correlations.**

3. **Calculate new mass of gas and liquid in the pipeline after a time step, and the new holdup.**

4. **Calculate the pressure in the gas phase after a time step using the ideal gas equation, with the gas volume, molecular weight, temperature and mass of gas.**

5. **Repeat from step 2, for as many time steps as required.**

22.2.3 Areas for Further Development

Areas where the code could be made more sophisticated and accurate are:

- **Calculating gas compressibility at each calculation step.**

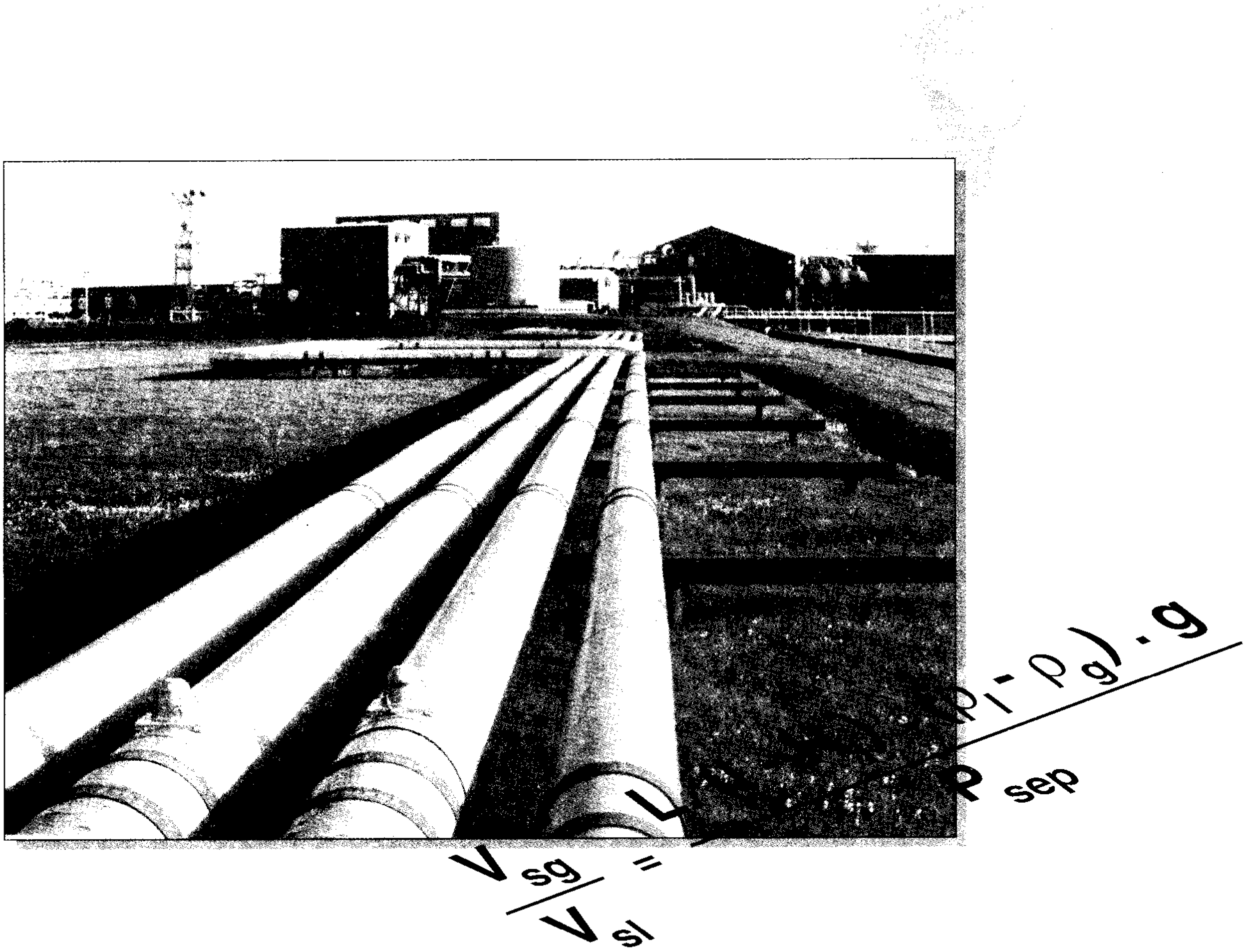
- **Allowing a mass flowrate which varies with pressure (inlet pressure). This would follow a well performance curve.**

- **Include thermal effects.**

- **Adding the ability to model fluids compositionally.**

- **Providing a starting point for the depressuring of a pipeline. (Same assumptions are made for this program as those made in the depressuring program.)**

Section 23 Pressure relief systems involving multiphase flow



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23.1 Introduction

The intent of this document is to concentrate on the specific issues arising from needing to relieve pressure from multiphase lines of significant length, and also to cover issues arising from having multiphase flow in any relief/disposal system, from flowlines or from process plant.

For a full treatment of the design and operation of relief systems please refer to the Relief Systems Team, Engineering and Operations Support, Oil Technology Centre, BP Oil, Sunbury. Everything contained within this document should be applied only in the context of, and as a supplement to, a comprehensive relief system analysis performed by suitably qualified individuals or organisations.

This section should also be read in conjunction with Section 8 De-pressuring and Blowdown, in which particular aspects relating to the generation of cold temperatures due to gas expansion are given a fuller treatment.

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23.2 Defining the reasons for requiring a relief system

The principal reason for requiring a relief system is to prevent overpressure of equipment and to keep the equipment within its design envelope. Additionally, blowdown may be a requirement to protect against external fires. Any release of inventory from the system that results from overpressure relief or from blowdown must be discharged in a safe manner.

Preventing over-pressuring

In oil and gas production over-pressuring could arise when the outlet from a system or line is shut in (e.g. by valve closure due to process upset), yet the supply continues to flow into the system due to failure, or absence, of the supply shut-in control system. Integrity could be compromised, with possible loss to the environment, if the maximum pressure seen by the system at any point exceeds the maximum allowable operating pressure (MAOP) of the system or any component thereof.

The time taken for the system pressure to build up to the MAOP is an important variable. The relief rate to prevent further pressure build up must be specified for the different possible failure scenarios. The relief rate is likely to be steady (on average over several minutes) until the source of supply can be shut-off. A relief control system is required to activate at a suitable pressure, and pass a flowrate sufficient to protect the overall system, such that MAOP is not exceeded.

In some cases it may be possible to protect pipes or vessels from the possibility of over-pressure by using instrumentation of acceptable integrity. This is designed to reduce to acceptably low levels the risk (multiple of frequency of occurrence and consequence) of a high pressure source being opened to a closed system which cannot contain the full shut-in pressure. In such a case a relief system may not be required.

Note: HIPS, HIPPS, Cat 1, 2a . . . are now old terminology. BPX follows the risk based method annotated in the UKOOA guidelines referring to SILs (safety integrating levels)

The judgment of whether to invest in full pressure containment, lower pressure rating with instrumentation of acceptable integrity, or lower pressure rating with relief, must be made with full knowledge of the risks and costs associated with each option.

Inventory removal

In case of fire or other emergency that may lead to over-pressure a relief valve may be desirable, but only if a non-reclosing device is used. This is rare. We would normally use a blowdown valve separate to the PSV or process equipment through a controlled disposal mechanism, rather than run the risk of large hydrocarbon inventories in the affected vessel/area being released if there was a leak or rupture due to the over-pressure.

In such cases the blowdown rate is likely to have an initial peak with subsequent decrease to zero as the system, which has no supply after shutdown, de-pressurises to a lower pressure setting or to atmospheric pressure.

Establishing possible relief scenarios for the complete system

It is necessary to determine the cases under which the relief system could be called to operate. As stated in the introduction, this is specialist work requiring expert help (quantified risk assessment). Contact the BP Oil experts directly, or via the Multiphase Group on request.

However, by way of introduction, these are the basic questions that need to be addressed.

What is the range of flowrates from the source?

The range needs to be specified as a function of time from start of relief, and, in the case of oil and gas wells, also as a function of time from commissioning (to take into account the likely changes in reservoir performance through the years of production). The flowrate range will affect the number and rating of relief devices in the system.

Is the disposal system connected to more than one source?

If so then both the frequency of each source being connected to the relief system, and the likelihood of multiple sources being connected to the disposal system at the same time, needs to be determined. This is vital in the determination of the maximum required capacity of the disposal system.

The answers to these questions will begin to form the design case for the relief and disposal systems. Detailed design will, however, need substantially more detailed investigation.

23.3 Describing the elements of the complete relief system

In order to design a relief system, it is necessary to predict correctly all the interdependent effects in order to ensure safe operation, and to have sufficient detail of all the elements comprising the system.

For any system it is necessary to specify the fluids properties over the full range of pressures and temperatures that could be seen during a relief event. It is possible to make simplifying assumptions, but the latest models for relief valve performance are dependent on the fluid properties so the more accurate the fluid compositional representation, the more reliable will be the prediction of system performance.

Any one part of the system will be affected by the way the fluids behave in another part. Slugging flow in a multiphase line being relieved will arrive at the relief valve, which then will see widely varying liquid and gas flow rates. The relief valve imposes a back pressure on the system being relieved, thereby affecting the flowrates delivered. The route from relief valve to disposal may also impose a significant back pressure on the relief valve which affects the fluids flow rate it can pass. The generation of slugs in the downstream pipework will require design of adequate pipe supports, appropriate design of the liquid knock-out system to achieve the required performance and careful design of the disposal system (e.g. to prevent liquid carryover into the flare).

It is therefore very important to carry at out an integrated assessment of the system over the whole range of relief scenarios to ensure that operation of this ‘safety’ system does not itself result in damage or loss.

Interaction with any relief valve vendor is important, to ensure they have full understanding of the conditions under which their product may be called for relief device selection to operate.

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23.4 Design for the 'worst case' relief scenario

What is the worst case?

In many systems the 'worst case', in terms of behaviour of the relief system, and its ability to protect the upstream elements without damage to itself or overloading of the disposal mechanism, is the case of the highest flowrates requiring relief.

There are also instances (such as the upstream pipework being long multiphase lines over hilly terrain) when low flowrate cases can be more severe in terms of ensuring protection against over pressurisation of the upstream system. Lower flowrates may also be more problematic in terms of behaviour of any multiphase flow occurring in the pipework between the relief devices and the disposal system. This can be true even when the flow arriving at the relief device is single phase.

So it is necessary to determine the worst case scenario for the system. As previously stated, from a process plant upset and shutdown, requiring inventory reduction or removal, then the highest flowrates experienced by the disposal system are likely to be just after the shutdown. Rates then decrease with time as the inventory reduces. Depending on the shutdown sequence, high rates may persist for some time, as a rate decrease from one part of the process is balanced by new flow from another part.

The natures of flow arriving at the relief and disposal systems need to be determined for a range of relief rates. If slug flow arrives at a relief valve, for example, then the pressure drop across the valve, and the flowrates the valve will pass, will vary with time. The length and liquid content of the slug may then determine the 'worst' case, which might not be at the highest flowrate.

A further extension to this occurs in the specific example of relief from hilly terrain flow. In long hilly terrain lines a plot of pressure loss across the line vs. flowrate through the line has a minimum. At flowrates less than the minimum pressure drop, the pressure loss is higher due to the dominance of gravitational over frictional pressure loss. So in such an instance when the relief system is required to protect a line, the highest pressures seen at the start of the line may occur at low flowrates. The relief system must lift at a condition such that the start of the line is protected from exceeding MAOP over all flowrates.

A logical step is to protect the system by installing relief at the most likely point of maximum pressure, or by installing a suitably reliable instrumented system to 'eliminate' the over-pressure risk. However, where reliability of an instrumented system and the access, logistics, and practicality of installing a disposal system are in question, then the compromise of having downstream relief may be the most satisfactory option.

Addressing the critical aspects of design

The issues which must be addressed having a requirement for detailed multiphase understanding and analysis are listed below.

- **The flow rate vs. time that must be relieved to protect each element of the system.**

- **The condition at which the relief system must come into operation.**

- **The back pressure exerted on the relief control systems by the downstream pipework and disposal mechanism for the various relief scenarios, and as a function of time.**

- **The nature of the flow arriving at the relief control system (steady, transient, cyclic . ...).**

- **The type, size and control method for the relief control system to allow the required flowrates to pass so the system is protected, taking into account the pressure upstream of the relief control system, the fluid properties (which will be time varying), the nature of flow arriving from the upstream pipework, and the back pressure exerted on the relief control system by the downstream pipework and disposal mechanism.**

- **The potential for erosion damage.**

- **The characterisation of any slugging in upstream and particularly downstream pipework to ensure adequate design of pipe supports.**

- **The changes in fluid temperature particularly across relief device, to ensure appropriate metallurgy is used.**

- **The maximum predicted gas and liquid flowrates, to ensure that these are accommodated by the disposal mechanism.**

To address the critical aspects of the system design listed above, it is currently necessary to use a variety of methods and software. For the relief rates from process plant SimSci's VISUAL FLARE or Hypertech's FLARENET could be used. For pressure drop along lines to the relief device, and then on to the disposal system, INPLANT, PIPESIM, PIPEPHASE or any other steady state multiphase simulator could be used. One issue here though is the prediction of the onset of critical flow in the low pressure line to the disposal system. Some programs may not use appropriate, or even any, checks on critical flow, and it should be noted that critical flow for two phase mixtures can occur at a much lower flowing velocity than for either of the single phase components.

The nature of flow in the lines upstream and downstream of the relief device may need to be simulated using both the steady state multiphase programs previously mentioned, and the transient simulators PLAC and OLGA. Slug sizes due to normal, slug flow terrain effects, or transients, will need to be ascertained.

Across a relief valve, the fluid behaviour in terms of pressure loss can be modelled using one of a number of methods (homogeneous, DIERS, . ..). For steady multiphase flow arriving at a valve the currently preferred method is DIERS for which a sizing Excel spreadsheet is available from the BP Oil Relief Team. However, the behaviour of a relief valve with slug flow arriving at its inlet has not undergone any experimental observation, and only some very preliminary modelling has been undertaken using PLAC.

Incorrect sizing of the valve in either possible direction can cause severe difficulties. If undersized then the pressure may not be sufficiently relieved, and over-pressure may not be prevented. If oversized then slugging issues may arise as the valve opens and closes, and the flow starts and stops. Such opening and closing is also not good for long term valve reliability.

Downstream of the relief device the nature of the flow is generally assumed to be given by steady state multiphase programs. However, the behaviour of a slug arriving at a relief valve, then passing through and experiencing a very large pressure drop, is not well defined. It is likely that large amounts of dissolved gas would flash off in, or just downstream of, the valve. Hopefully this will cause break up of the slug. The question of whether a slug can re-form in the downstream pipework is a very important one. If slugs do form, or re-form they could be travelling at very high velocity. If the design case is a high flowrate relief then it is likely that the flowing velocities could be too high to sustain slug flow.

The performance of any downstream knock-out pot can be assessed (assuming no slugging) using PROSEPARATOR software. Flare backpressures will need to be obtained as a function of flowrate from the manufacturer (assuming no slugging).

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23.5 Ensure integration of all design aspects

Assessment of the different elements of the design will be undertaken using a variety of methods and software, as described above. These various computer programs are not yet well linked from a software point of view. There is no direct feedback/feedforward between the programs, so a once off calculation is not possible. Several iterations through the different programs may be required to ensure the overall ability of the system to relieve the required flowrates and give adequate protection.

As yet there is no software package which can incorporate all the required models for different parts of the system under different relief scenarios. It is therefore important to integrate all information obtained and generated to ensure the relevant interactions (especially due to backpressures) are thoroughly understood and taken into account.

If a source flowrate is identified, then the different software packages will initially be used to determine pressure loss in upstream pipework, pressure loss across a given type and size of relief valve, and pressure loss in downstream pipework and disposal system. If the system does not generate a consistent overall pressure loss from supply to discharge then iterations on relief valve size or type, and upstream/downstream pipe diameters, CAPEX of pipe and vessel sizes must be undertaken until a satisfactory solution is obtained.

The extent of effort required here will depend on whether the work is retrofitting into an existing system, or a completely new design with fewer existing constraints.

It is very important that an individual is nominated to carry the principal coordinating role. All interactions between sections of the system need to be covered, to ensure that the action or response of one part of the system is not detrimental to the action or response of another part. This integrating role should also be the one to work as much as possible towards designing out the requirement for multiphase relief.

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23.6 Assess sensitivity to uncertainty

All of the preceding argument assumes that input data, especially in terms of flowrates and fluid properties, is fixed and accurate. In reality these variables may have a degree of uncertainty for which the significance on the design of the over-pressure protection system needs to be assessed. We have to ask by how much does the predicted behaviour of the relief system, and its ability to protect, change with variations in input data which have ranges of uncertainty. What is the worst realistic combination of the uncertainties?

Variations in possible flowrates of gas, oil and water from the source as a function of months and years should be checked for the influence on total mass flowrate and slugging tendency.

Water and gas fraction variations will also affect the performance of any relief valve (because of changes in fluid properties), which should be checked across the expected range of likely values.

Best judgment should be made of what is the worst realistic combination of the uncertainties, in order to develop the worst case.

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23.7 Check critical aspects of design

If the 'worst case' can still be accommodated in the design basis generated then the design is sufficient. If the 'worst case' is not compatible with the design then initial work should focus on a better definition of the input data. If this is not possible then the design should be modified to suit, with note taken to ensure any CAPEX increase is not disproportionate to the cost of getting better input data.

In terms of improvement of design methods, the behaviour of relief valves under multiphase flow, and slug flow in particular, requires investigation, whether through DIERS or a separate joint industry research project. The integrity of the relief system will often depend on slugs not being present in the downstream pipework to the disposal process, especially if slugs are arriving at the inlet of the relief device, so tracking of slug behaviour through the system is also important.

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23.8 Find specification of relief control mechanism, downstream pipework and disposal mechanism.

Once all issues listed above have been addressed, and this information linked into the specialist studies of relief system experts, then final specification of the relief system can be made.

Much of this process should be repeated when additional flow/pressure sources are to be connected into the system.

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