BASIC DESIGN AND SIMULATION OF OIL PRODUCTION PIPELINES

A Thesis

by

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The undersigned hereby certify that they have read and recommended to the Faculty of Mineral Resources Engineering for acceptance a thesis entitled "**Basic Design and Simulation of Oil Production Pipelines"** by **Siaw Clement Gyasi** in partial fulfillment of the requirements for the degree of **Master of Science**.

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Abstract

After the successful drilling and completion of wells, the formation fluids are produced to the surface at the wellhead. The produced fluids are then transported at a distance of several miles from the wellhead to the production facility, where they go through separation (into water, oil and gas), conditioning, treatment, processing, measurement, refining or storage. The transportation usually takes place through production pipelines that sit on the land or on the seafloor. The design of the pipelines is crucial both for maximizing the oil/gas throughput, as well as for minimizing the shut in and start up times. The design involves testing and simulation of several possible oil/gas exploitation scenarios, in order to deliver the produced fluids to the separation facility at the recommended pressure and temperature conditions. Several factors should be included in the study, such as fluid pressure, temperature and velocity, phase distribution at each section of the pipeline, heat transfer rate to the environment, erosion-corrosion restrictions and the possible formation of flow restrictions like solid precipitation, deposition, gel effect from the formation of waxes etc.

The thesis is a thorough research of two offshore oil reservoirs with three production wells each that gather into two production manifolds and they are routed through a 12 km long pipeline to the separation facility. Two possible sets of pipeline sizes are tested through simulation, which includes the reservoir performance, the performance of each well and the performance of each set of pipelines. The IPM simulation suite (MBAL, PROSPER and GAP) was used for conducting the necessary simulations. The results include a set of possible solutions for maintaining maximum production for a period of ten years.

Contents

1. PIPING SYSTEMS AND FLUID CHARACTERISTICS

1.1 Introduction

After the successful drilling and completion of wells, the formation fluid is produced through these wells to the surface. At the surface, the produced fluid is transported from the wellhead through surface production pipelines or flowlines to a production facility where separation (into water, oil and gas), conditioning, treatment, processing, measurement, refining or storage takes place. It is therefore important to design and optimize these flowlines to be able to deliver the produced fluid to the facility at the recommended pressure and temperature conditions.

Before the production pipelines or flowlines are set up to transports fluids, it is important to consider the following design parameters:

- What type of fluid is transported at which pressure and temperature conditions and at which rate? Does the fluid contain any solids or impurities?

- What is the distance that the fluid is being transported? Are there any elevation changes on the path?

- What will be the use of the fluid after the transport?

The above questions are used to determine; the type of piping (material, size and thickness), valves and fittings, the type of pressure maintenance and heating device / utilities and the type of support the pipelines would require.

Fig. 1.1 is a simplified block diagram that illustrates the basic "wellhead to sales" concept. The diagram begins with wellhead choke, which is used to control the rate of flow from each well. The fluid from the well travels through a flowline to the production facility where the fluid is separated, conditioned, treated, processed, measured, refined, or stored [1].
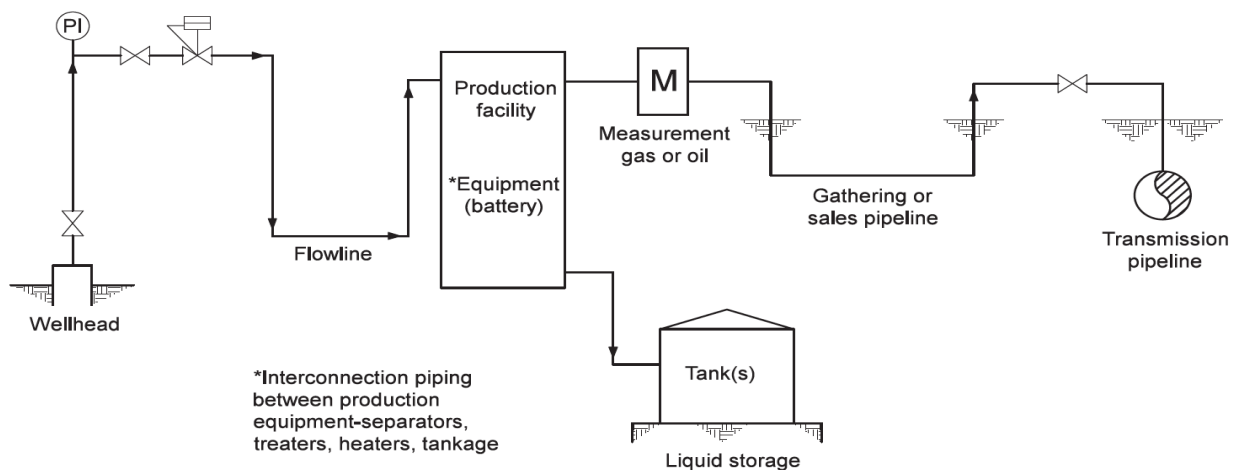
Figure 1.1 Block diagram of "wellhead to sales" concept. [11]

1.2 Fluid types

Fluids, which are transported through pipelines, are of the following types:

Gases

- Unprocessed natural gas or rich gas consisting primarily of methane and some heavier hydrocarbons.
- Processed natural gas or lean gas consisting primarily of methane, although small amounts of heavier fractions may still be present.
- Non-hydrocarbon components such as nitrogen, hydrogen sulfide and carbon.
- Natural gas liquids (NGLs) consisting primarily of the intermediate-molecular-weight hydrocarbon components such as propane, butane, and pentanes plus [1].

Crude oil

Crude oil consists of hydrocarbons that are liquid at atmospheric conditions in storage tanks. Stabilisation of volatile oils is ensured to prevent excessive vapor formation or "weathering" in storage or transport tankers. Pipeline fluid remains liquid during transportation due to adequate operating pressure [1].

Water

Well effluents normally contain dissolved salts and minerals that are usually corrosive; therefore, pipelines must be able to transport water.

Two-phase fluids

Two-phase fluids usually consist of natural gas and condensate or crude oil and associated gas. Flow lines from the well to the production facility are designed for two-phase flow [1].

Combinations

Well effluents contain the following:

- Hydrocarbons
- Water
- Varying amounts of CO_2 and H_2S

- Hydrocarbons and water combine to form emulsions that have higher viscosity than that of oil or water alone.

1.3 Fluid characteristics

1.3.1 Physical properties

The physical properties of the transported fluids play an important role in the determination of the pipe diameter and the selection of the pipe materials and the associated equipment. They are also important in determining the power required to transport the fluid [1]. The following fluid properties mostly affect pipeline design:

- Composition
- Density
- Viscosity
- Vapor pressure
- Water content
- CO_2 and H_2S content

Composition

Well stream compositions are usually stated as mole fractions. Gas streams consist of C1 through C6 and C7+, while oil streams consists of C1 through C10+. Fluid composition is used to predict fluid properties such as density, viscosity and phase behavior. If a compositional analysis is not available, one must rely on the "black oil" characterization in which API gravity, gas gravity, gas-oil ratio, and water-liquid ratio are given. The use of empirical black oil property correlations provides reasonable values for density, viscosity, and phase behavior [1].

Density

Several definitions of fluid density are used in upstream oil and gas operations such as density, specific gravity or relative density, and API gravity. The density of a fluid is defined as the mass per unit volume in lbm/ft^3 or kg/m^3. Density is a thermodynamic property and it is a function of pressure, temperature, and composition [1].

Liquid densities are higher than gas densities and are less affected by pressure and temperature. Gas densities increase with pressure and decrease with temperature.

Liquid density is often specified by giving its specific gravity relative to water at standard conditions of 60 °F and 14.7 psia (15.6 °C and 101.4 kPa) [1].

Field Units

$$\rho = 62.4\,(SG) \qquad \text{(Eqn 1.1a)}$$

SI units

$$\rho = 1000(SG) \qquad \text{(Eqn 1.1b)}$$

Where,

ρ = density of liquid (lb/ft^3 or kg/m^3),

SG = specific gravity of liquid relative to water.

API gravity is a special function of relative density. It is a reverse graduation scale of relative density, where lighter fluids have higher API gravities. For example, a condensate would typically have API gravity between 30 and 40, while water would have an API gravity of 10. API gravity is defined as [1]:

Field Units

$$API = \frac{141.5}{SG\ at\ 60^0F} - 131.5 \qquad \text{(Eqn 1.2a)}$$

SI units

$$API = \frac{141.5}{SG\ at\ 15.6^0F} - 131.5 \qquad \text{(Eqn 1.2b)}$$

The density of a mixture of oil and water is determined by the weighted-volume average of the two densities and is given by:

$$\rho = \frac{\rho_w Q_w + \rho_o Q_o}{Q_T} \qquad \text{(Eqn 1.3)}$$

Where,

ρ = density of liquid (lb/ft^3 or kg/m^3),

ρ_o = density of oil (lb/ft^3 or (kg/m^3),

ρ_w = density of water (lb/ft^3 or (kg/m^3),

Q_w = water flow rate (BPD or m^3/h),

Q_o = oil for rate (BPD or m^3/h),

Q_T = total liquid for rate (BPD or m^3/h).

A liquid's specific gravity is measured relative to water, and that of a gas is measured relative to air. The conditions of pressure and temperature at which specific gravity is measured are as follows:

- For "standard conditions", temperature is taken as 60 °F (15.6 °C) and pressure as 14.7 psi (1.01325 bar).

- For "normal conditions", temperature is taken as 32 °F (0 °C) and pressure as 14.67 psi (1.01325 bar).

The above stated conditions should be specified correctly, as using the wrong conditions by mistake can result in volumetric errors of about 5%, which can be significant. Below is the average specific gravity of some oil field liquids [1]:

Crude oil = 0.825

Condensate = 0.75

LPG = 0.58

The specific gravity of an oil and water mixture can be calculated by:

$$(SG)_m = \frac{(SG)_w Q_w + (SG)_o Q_o}{Q_T} \qquad \text{(Eqn 1.4)}$$

Where,

$(SG)_m$ = specific gravity of liquid,

$(SG)_o$ = specific gravity of oil,

$(SG)_w$ = specific gravity of water,

Q_w = water flow rate (BPD or m^3/h),

Q_o = oil flow rate (BPD or m^3/h),

Q_T = total liquid for rate (BPD or m^3/h).

Gas density depends on pressure, temperature, and molecular weight. The specific gravity or relative density of a natural gas at standard conditions of pressure and temperature is determined by its apparent molecular weight. Since [1] the apparent molecular weight of air is 29 lb/lb-mole (kg/kmole), the specific gravity of a gas is given by:

$$S = \frac{(MW)}{29} \qquad \text{(Eqn 1.5)}$$

where,

S = specific gravity of gas relative to air,

(MW) = apparent molecular weight of gas.

The density of a gas under specific conditions of pressure and temperature is given by:

Field units

$$\rho_g = 2.70 \frac{SP}{TZ} \quad \text{or} \quad \rho_g = 0.093 \frac{(MW)P}{TZ} \qquad \text{(Eqn 1.6a)}$$

SI units

$$\rho_g = 3.48 \frac{SP}{TZ} \quad \text{or} \quad \rho_g = 0.1203 \frac{(MW)P}{TZ} \qquad \text{(Eqn 1.6b)}$$

Where,

ρ_g = density of gas (lb/ft^{3}or kg/m^3),

P = pressure (kPa or psia),

T = temperature (K or °R),

Z = gas compressibility factor,

S = specific gravity of gas relative to air,

(MW) = apparent molecular weight of the gas.

Viscosity

      Viscosity is a measure of a fluid's internal resistance to flow. It is determined either by measuring the shear force required to produce a given shear gradient or by observing the time required for a given volume of liquid to flow through a capillary or restriction. When measured in terms of force, it is called absolute or dynamic viscosity. When measured with respect to time, it is called kinematic viscosity. A fluid's kinematic viscosity is equal to its absolute viscosity divided by its density [1]. The unit of absolute viscosity is poise or centipoise (cP). The unit of kinematic viscosity is stoke or centistokes (cSt). The relationship between absolute and kinematic viscosity is given by:

$$\mu = \gamma\rho \qquad\qquad\qquad\qquad \text{(Eqn 1.7a)}$$

Field units

$$\mu = (SG)\gamma \qquad\qquad\qquad\qquad \text{(Eqn 1.7b)}$$

SI units

$$\mu = 1000(SG)\gamma \qquad\qquad\qquad\qquad \text{(Eqn. 1.7c)}$$

Where,

μ = absolute viscosity (cP or Pa s),

γ = kinematic viscosity (cSt or m^2/s),

ρ = density (lb/ft^3 or kg/m^3),

SG = specific gravity relative to water.

      Fluid viscosity varies with temperature. For liquids, viscosity decreases with increasing temperature. The viscosity of oil is highly dependent on temperature and is best determined by measuring the viscosity at two or more temperatures and interpolating to determine the viscosity at any other temperature [1]. When data are not available, the viscosity of a crude oil can be approximated from figure 1.2, if, the oil

is above its cloud point temperature, that is, the temperature at which wax crystals begin to form when the crude oil is cooled [1].

Figure 1.3 can be used to account for the fact that oil at higher pressures has lighter hydrocarbon components and so has a higher API gravity and lower viscosity than at stock tank conditions [1].
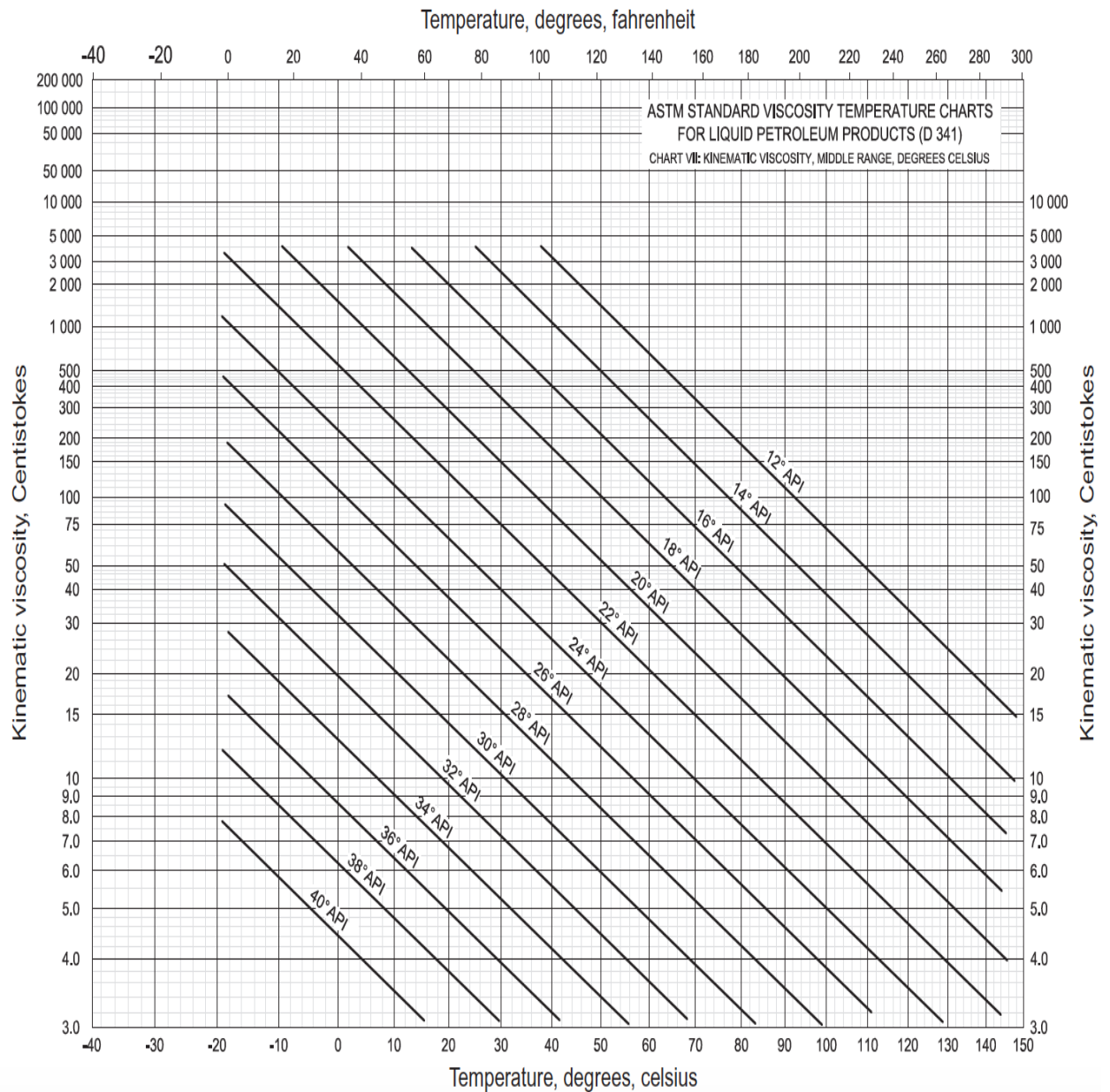
Figure 1.2 Kinematic viscosity versus API for different API gravity oils [1]

The viscosity of an oil-water mixture is not the weighted average of the two viscosities. Depending on the ratio of water and oil and the degree of mixing (shear

rate) in the system, the viscosity of the mixture can be as much as 10-20 times that of oil. The following equation has proved useful in analyzing piping system [1].

$$\mu_{eff} = (1.25\emptyset + 10\emptyset^2)\mu_c \qquad \text{(Eqn 1.8)}$$

$$1cp = 6.72 \times 10^{-4} \text{ lbm/ft.-s}$$

Where,

μ_{eff} = effective viscosity (cP or Pa s),

μ_c = viscosity of the continuous phase (cP or Pa s),

$\emptyset$ = volume fraction of the discontinuous phase.
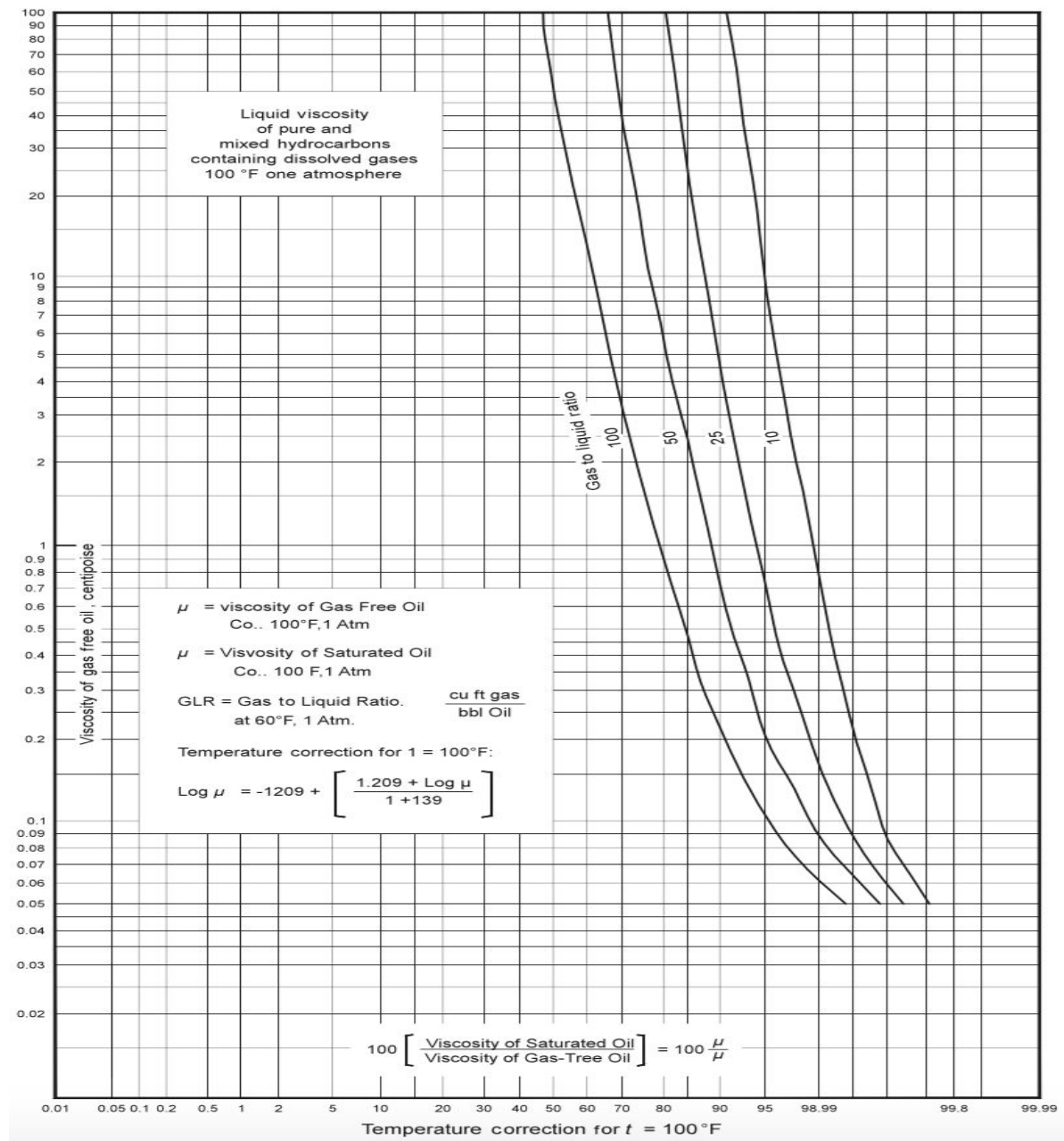
Figure 1.3 Liquid viscosities of pure and mixed hydrocarbons containing dissolved gases.

Gas viscosity depends on temperature, relative density, and pressure. The viscosity of a natural gas can be determined from 1.4. For most production facility gas piping applications, a viscosity of 0.012 cP can be assumed [1].
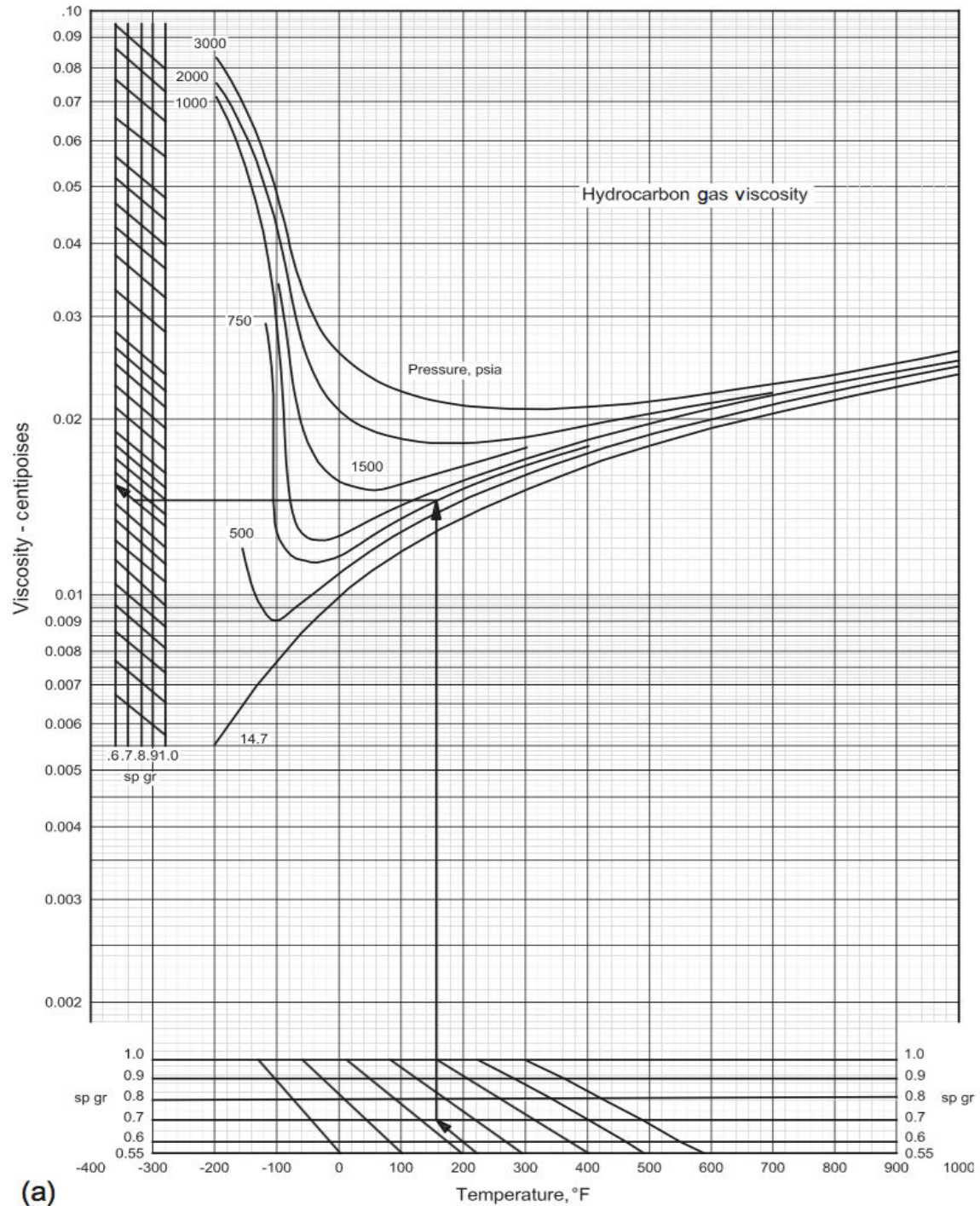
Figure 1.4a Hydrocarbon gas viscosity (field units) (Courtesy GPSA)

(Continued)
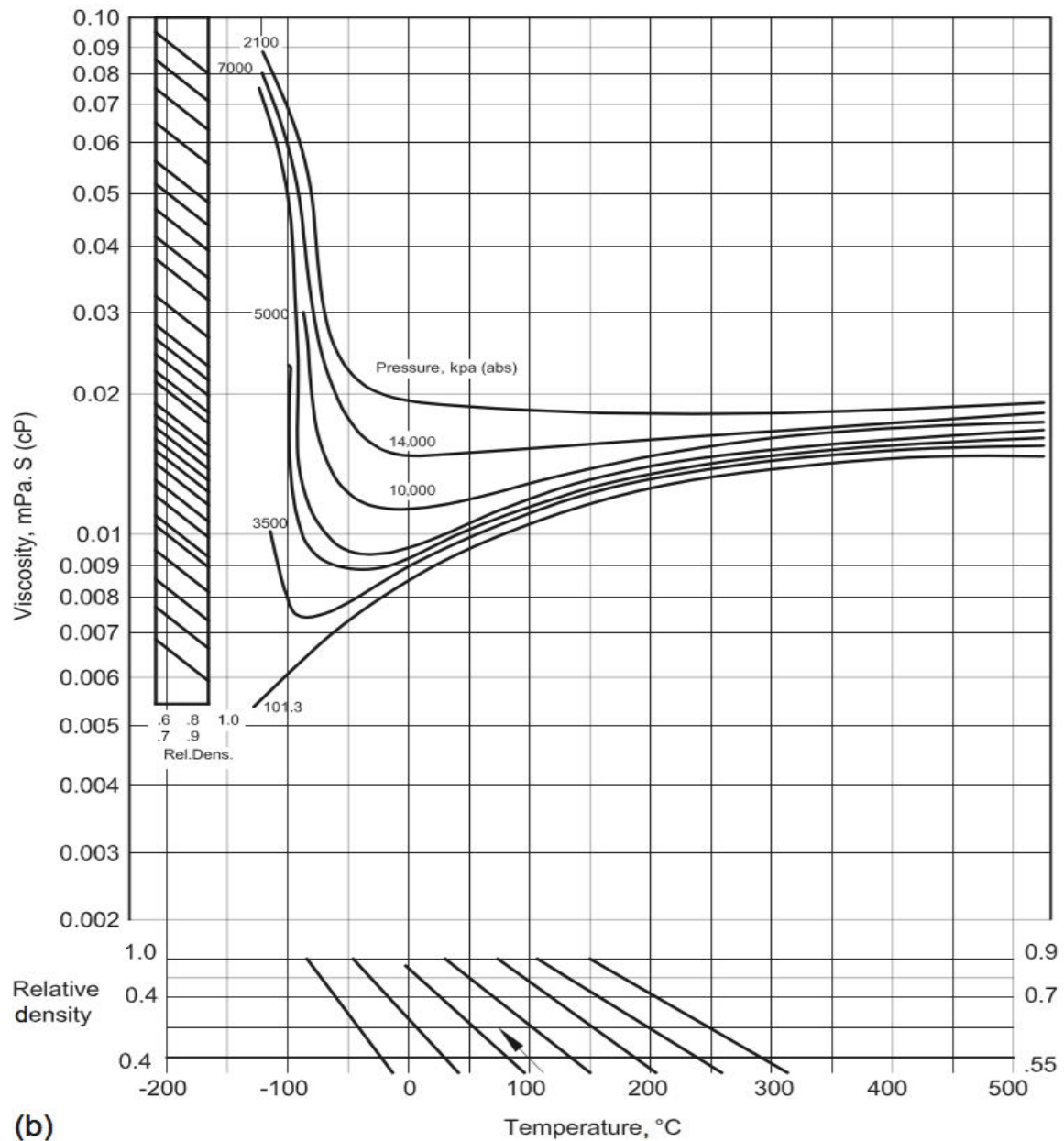
Figure 1.4 Continued. (b) Hydrocarbon gas viscosity

Vapor pressure

Vapor pressure is the pressure exerted by the vapor phase of a fluid in a confined container at a given temperature. Vapor pressure increases with temperature. Figure 1.5 shows the variation of vapor pressure with temperature for certain light petroleum products. Vapor pressure determines the operating conditions at which a fluid moves from liquid phase flow, to two-phase flow [1].
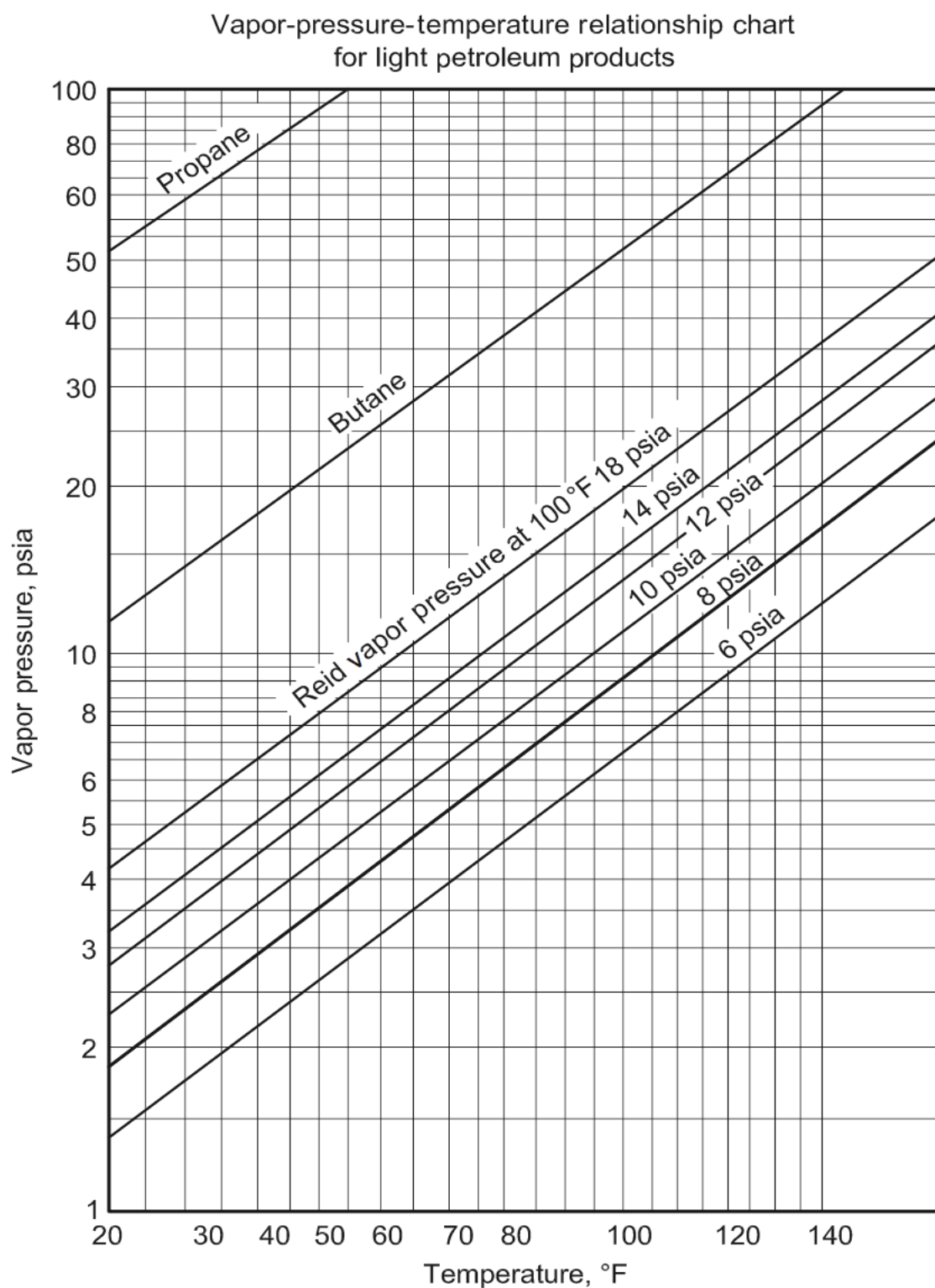
Figure 6.6 Vapour-pressure-temperature relationship chart for light petroleum

products

Water, CO_2, and H_2S content

The presence of water, CO_2 and H_2S in transport fluid increase the risk of internal corrosion in pipes. The knowledge of their amounts present is important to select the right pipe material, or coating, to prevent internal corrosion.

1.3.2 Phase behavior

Phase behavior refers to the mole fraction ratio of vapor to liquid of the fluid as a function of pressure, temperature, and composition. For NGL and volatile oils, calculation of the fluid phase behavior is necessary for the determination of pressure drop-flow rate relations. If the fluid composition is known, flash calculations can be performed to determine the phase envelope and the relative amounts of liquid and vapor in the two-phase region [1].

1.3.3 Deposits forming tendencies

This is thoroughly discussed in chapter 4.

1.4 Types and functions of facility and pipeline systems

Once the type and character of the fluid are known, the designer must determine the suitable type of piping system. Piping and pipeline systems include processing equipment, such as separators, heaters, and tanks, pumps/compressors and the interconnection system of pipes, valves, fittings, instrumentation and their support systems such as racks and/or hangers[1].

The pipeline systems of the producing wells include the well flowline, trunkline, facility (on-plot) interconnecting equipment piping within the production facility, gathering or sales pipelines, and transmission pipelines. A brief description of the aforementioned facility piping and pipeline systems follows [1].

- *Flowline*

A well flowline identifies a two-phase line from a wellhead to a production manifold or separator. The line between a production manifold and the first separator is sometimes referred to as a flowline but is generally referred to as a production header. Flowlines range in size from 2 in. (50.8 mm) to 20 in. (508 mm) [1].

- *Trunkline*

  A trunkline is a larger line that connects two or more well flowlines that carry the combined well streams to the production manifold. Trunklines range from 10 in.(25.4 mm) to 42 in. (106.8 mm) [1].

- *Manifold*

  A manifold is a combination of pipes, fittings, and valves used to combine production from several sources and direct the combined flow into appropriate production equipment. A manifold may also originate from a single inlet stream and divide the stream into multiple outlet streams. Manifolds are generally located where many flowlines come together, such as gathering stations, tank batteries, metering sites, separation stations, and offshore platforms. Manifolds are also used in gas-lift injection systems, gas/water injection systems, pump/compressor stations, and gas plants installations where fluids are distributed to multiple units. A production manifold accepts the flow streams from well flowlines and directs the combined flow to either test or production separators or tanks [1].

- *Production header (and/or test header)*

  In a gathering system, the production header connects flowlines from several wells into a single gathering line that is routed to a test separator. The header has production and testing valves to control the flow of each well, to the production facility or separately to testing vessels. Production headers measure well production of oil, water, and gas rates by directing individual wells through a test header to the test separator with its special metering equipment [1].

- *Gathering system*

  A gathering system consists of a line downstream of field manifolds or wellhead separators conveying fluid from multiple wells and leading to the production facility. The gathering system may handle condensed hydrocarbon liquids, water, and corrosive gas and may require special design considerations [1].

- *Separation station*

  A separation station separates the well stream into gas, oil, and/or water.

Increasing the number of stages of separation increases the liquid recovery [1].

- *Compressor station*

    A compression station consists of a gas compressor with its associated piping, coolers, and scrubbers. Low-pressure gas is normally boosted to sales or injection pressure by the compressor [1].

- *Pumping station*

    A pumping station consists of a pump and associated piping with the purpose of boosting liquid pressure to provide the energy for transmission and distribution of a liquid [1].

- *Production facility*

    A production facility is a facility for the field handling, conditioning, treating, processing, and / or measuring of gas, oil, and / or water [1].

- *Facility (on-plot) interconnecting piping*

    Facility piping consists of piping within a well-defined boundary of processing plants, piping compression stations, or pumping stations. The piping is used for conducting a variety of fluids within those boundaries as required [1].

- *Transmission line*

    A transmission line consists of a cross-country piping system for transporting gas or liquids, usually over long distances. The inlet is normally the custody transfer point or the production facility boundary with the outlet at its final destination, for example, processing plants and refineries. Transmission lines are usually long and have large diameters. Transmission lines also transport hydrocarbon fluids from producing fields to processing plants or refineries and from plants and refineries to marketing centers [1].

- *Injection line*

    An injection system is similar to a gathering line but flows in the opposite direction (toward the wells). Injection lines transport high-pressure gas or water to

wellheads for injection into the producing formation. Fluids are injected for reservoir enhancement, production enhancement, gas conservation, gas lift, produced water disposal, water flood, and steam flood. The injection system pressure is usually greater than the gathering system pressure [1].

1.5 Introduction to fluid flow design

When designing facility piping and/or pipeline systems, it is essential to optimize the line size and determine pump and/or compressor requirements [1]. The following factors should be considered when determining the size of a line to meet the design requirements:

Volume of fluid

The main consideration in line sizing is the volume of fluid that is transported through the piping system. The exact volume is rarely known during the initial design stage. An estimate is normally used for initial design purposes. When the exact throughput is unknown or uncertain, designers must compromise constructing between a large line that is capable of delivering future projected volumes and one satisfying the current peak requirements only. Excess capacity reduces line profitability, while too small a line might need to be expanded in the future [1].

Distance

For pipelines, the distance between the entry point and the delivery point must be known. The designer also needs to know the type of terrain the pipeline must traverse and the elevation profile along the right-of-way as it affects pressure loss and power requirements. The designer must also take into consideration, the environmental conditions, ecological, historical, and archaeological sites as they might influence the pipeline routing, thereby increasing the length of the pipeline [1].

Pressure loss

The pressure loss is a key factor in both facility and pipeline design. An accurate pressure loss projection is critical, as it will directly influence the pipeline's ability to meet design specifications. Available piping inlet pressure must be known, as well as if there is any particular outlet requirement at the delivery point [1].

2. MULTIPHASE FLOW

Multiphase flow in pipelines is whereby liquids (oil and water) and gases flow simultaneously. In some cases, it also includes the flow of solids. Generally, in production operations, two-phase flow exists from the wellbore to the first separator in the production facility. Therefore, in this text, only two-phase flow will be considered.

2.1. Two-phase flow

In two-phase flow, gas and liquid flow simultaneously. The way in which gas and liquid flow simultaneously is characterized by the flow regime. Flow regime indicates the distribution of phases over the cross-sectional area of the pipe. The type of flow pattern depends primarily on the flow rates of the phases. The relative positions of the liquids and gas as they move along the pipeline depend whether the pipeline is vertical or horizontal. The types of two-phase flow regimes, common in pipelines are [1]:

- Dispersed bubble flow
- Elongated bubble or Plug flow
- Annular flow
- Stratified or Laminar flow
- Slug flow

2.1.1 Two-phase flow in horizontal pipes

In horizontal flow, the heavier phase tends to settle at the bottom with the lighter phase on top. A typical two-phase flow in a pipeline is represented on figure 2.1.

- *Dispersed bubble flow:* This type of flow occurs when there is a high liquid-gas ratio at very high flow rates, enough to cause turbulence to disperse the gas within the liquid. The gas bubbles concentrate at the upper part of the pipe.

- *Elongated bubble or Plug flow*: It occurs at moderate gas and liquid rates when smaller bubbles of gas (bubble flow) dispersed in the liquid coalesce to form bigger bubbles called plugs.

- *Stratified smooth flow:* It occurs when both the gas and liquid phases flow at very low velocities. The gas plugs coalesce to form a continuous phase and flows

separately at the upper part whilst the liquid flows at the lower part of the pipe with a distinct interface between them. It has the highest tendency to occur in downhill or horizontal flow lines.

- *Stratified wavy flow*: It occurs when the velocity of the gas increases under stratified conditions. The increased gas velocity drifts away liquid at the gas-liquid interface creating waves at the liquid surface. This transitions the flow from stratified smooth to stratified wavy flow. This type of flow occurs in downhill lines.
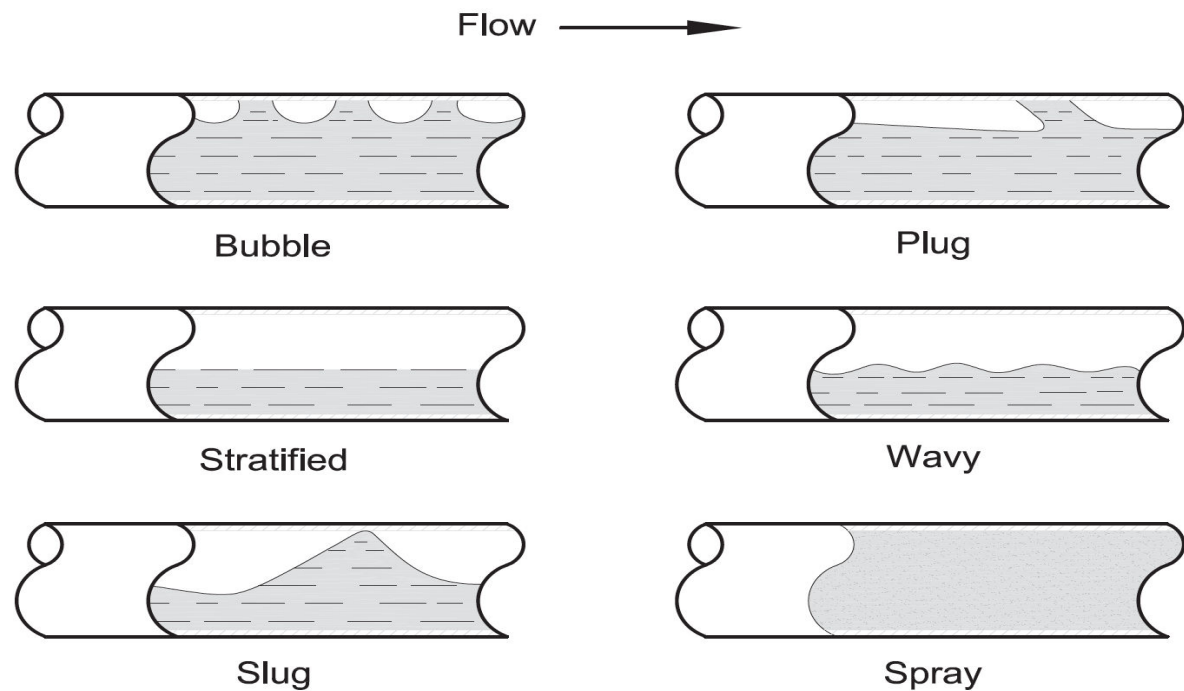
Figure 2.1 Two-phase flow regimes in a horizontal pipe

- *Annular flow:* It occurs at extremely high gas-liquid ratios. Some of the liquid deposits on the surface of the pipe wall and moves as a thin film whilst the rest remains as droplets entrained in the gas stream.
- *Slug flow:* It occurs at moderate liquid and gas flow rates. It is similar to the stratified wavy flow but a much higher gas velocity causes the wave crests to contact the top of the pipe, thereby trapping gas between the wave crests [1]. The gas flow then becomes discontinuous leading to the formation of alternating flow of liquid and gas slugs. The length of these slugs can be several hundred feet long. This type of flow normally occurs in level and uphill lines.

Slugging: Slugs usually lead to significant pressure fluctuations and uneven arrival of gas and liquid at the processing facility causing tanks to flood, therefore are generally unwanted [2]. The different types of slugs are:

- ✓ ***Hydrodynamic slugs***: They are formed in horizontal pipes as already described above and shown in figure 2.1. They are relatively short.

- ✓ ***Terrain generated slugs:*** They are formed when the pipe's elevation profile creates local minima [2]. Liquid flows faster and accumulates at the low point thereby blocking the gas. The gas pressure builds up at the upstream side of the low point until the pressure is enough for the gas to force its way into the liquid-filled uphill section. The liquid section carried by the gas becomes shorter and shorter until the pressure is enough to carry it upwards. As a result, both the liquid and the gas accelerate out of the pipe in the form of a blowout. Unlike hydrodynamic slugs, terrain generated slugs can have several hour time periods [2]. This type of slug is demonstrated in figure 2.2.

Gas pressure similar at both sides of liquid.

Gas pressure builds up at upstream side.

Gas reaches lowest point and blows out the liquid.

Much of the liquid and gas has escaped. The flow is reduced and some liquid falls back.
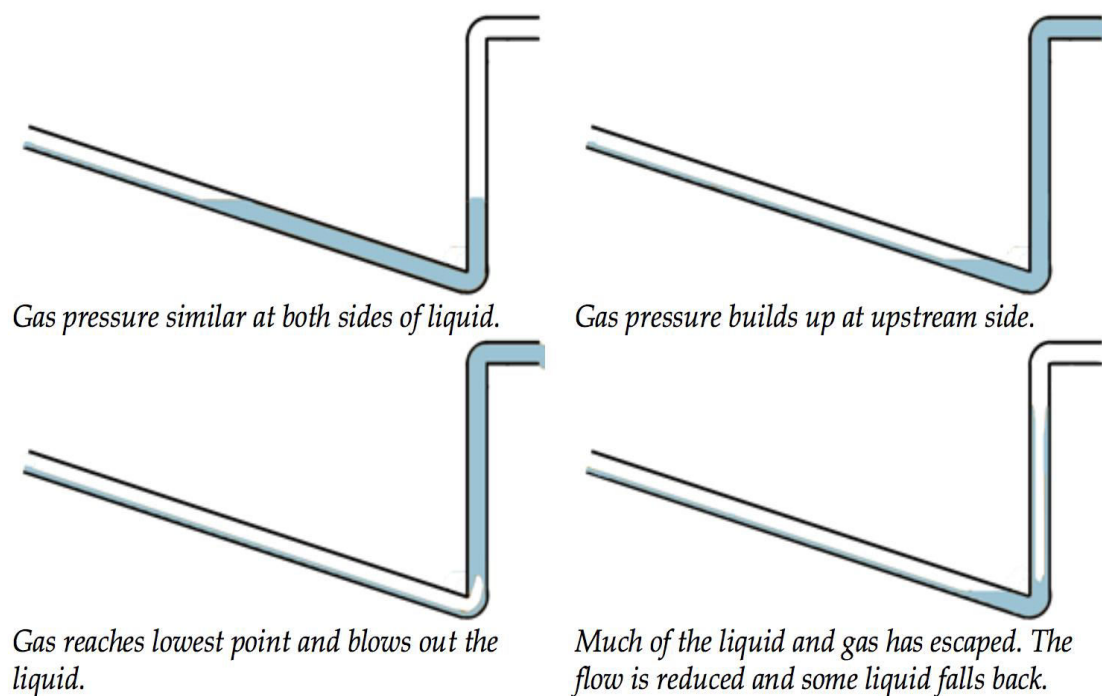
Figure 2.2 Terrain slug formation

- ✓ ***Riser based slugs:*** Some risers have a J or U shape. In addition, if the seabed in front of the riser slopes downwards, slugs can be created at the low points. Controlling the choke at the riser's outlet and/or the choke at the wellhead can greatly reduce the formation of riser-based slugs.

The stratified smooth and stratified wavy flows are grouped together as

"*segregated flow*" because the different phases flow in separate layers. The elongated bubble and slug flows are grouped as "*intermittent flow*" where the flow of each phase is interrupted from the other phases. The annular and dispersed flows are grouped as "*distributive flow*" because the different phases appear in the flow without following any order [3].

Horizontal Flow regime maps

Flow regime maps are used to visually inspect the expected flow regimes for any flow condition in a horizontal pipe. Along the horizontal axis is the superficial gas velocity V_G, and on the vertical axis is the superficial liquid velocity V_L. The superficial velocity of any phase is the volumetric flow rate of that phase divided by the cross-sectional area of the pipe. A typical horizontal flow regime map is shown on figures 2.3.
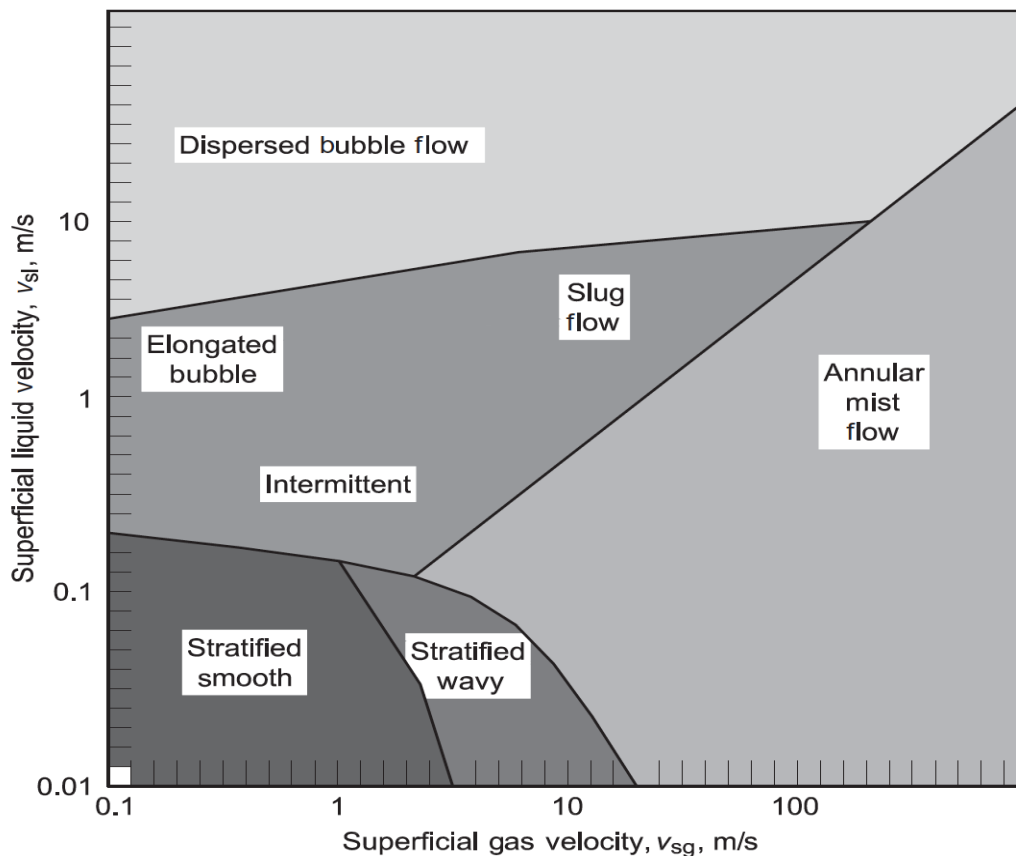
Figure 2.3 Horizontal multiphase flow regime map [4].

It can be seen that, for very low Vs_l and V_{sg}, stratified flow is observed with the gas on top and the liquid at the bottom. As the gas flow increases, waves start forming on the liquid surface due to increased friction at the gas-liquid interface and

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the flow transitions into stratified wavy flow. Further increase in the gas velocity intensifies the turbulence and rips more liquid from the liquid surface, which becomes entrained within the gas stream as tiny droplets. This reduces the liquid level and the previously horizontal surface bends around the inside of the pipe covering the whole circumference of the pipe as a thin film [2]. The droplets carried by the gas occasionally hit the walls of the pipe and fall into the liquid film. This is called annular flow.

In case of high liquid flow, the liquid disperses the gas into tiny bubbles within the liquid phase and carries them along. As the liquid flow decreases, the bubbles move upwards to the upper part of the pipe and coalesce to form larger bubbles called Taylor bubbles. The Taylor bubbles are large gas bubbles separating liquid slugs.

Flow regimes are generated based on pipeline diameter, inclination and fluid properties. Upward inclinations cause slug flow whilst downward inclinations cause stratified flow. Therefore, a flow map generated for a specific input data, is not general enough to be valid for other data sets.

From the flow regime map, it can be seen that multiphase flow attends different flow regimes. These flow regimes are dependent on the difference in rate and velocity between the phases [5]. In figure 2.3, the multiphase flow is simplified to two phase flow, gas and liquid.

2.1.2 Two-phase flow in Vertical pipes

- *Bubble flow*: Small gas bubbles of different diameters become dispersed in the liquid phase. The velocity of the gas bubbles depends on their diameters. The gas phase has little effect on the pressure gradient.

- *Slug flow*: Small gas bubbles coalesce to form bigger ones of approximately the same size and shape which are nearly the diameter of the pipe. Liquid slugs separate these gas bubbles. Both the gas and liquid phases have significant effect on the pressure gradient.

- *Churn or Transition flow*: It is the type of flow whereby there is a transition from a continuous liquid phase to a continuous gas phase. The gas phase is predominant and the liquid becomes entrained in the gas. Gas bubbles combine to form plugs and flow in the center of the pipe. The effect of the liquid is still significant.

- *Annular or Mist flow:* The gas is the continuous phase whilst most of the liquid is entrained in and carried by the gas. The gas flows at a high velocity in the centre of the pipe whilst the rest of the liquid move along the pipe wall as thin liquid film.
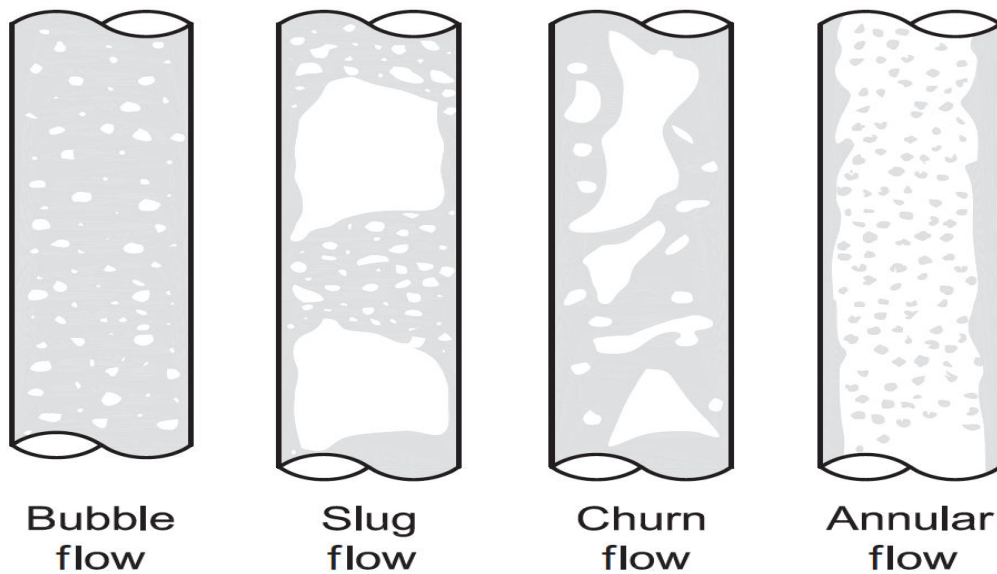
Figure 2.4 Two-phase flow regimes in a vertical pipe [6]

Vertical Flow regime map

Figure 2.5 can be used to determine the type of flow regime to be expected in a vertical pipe.
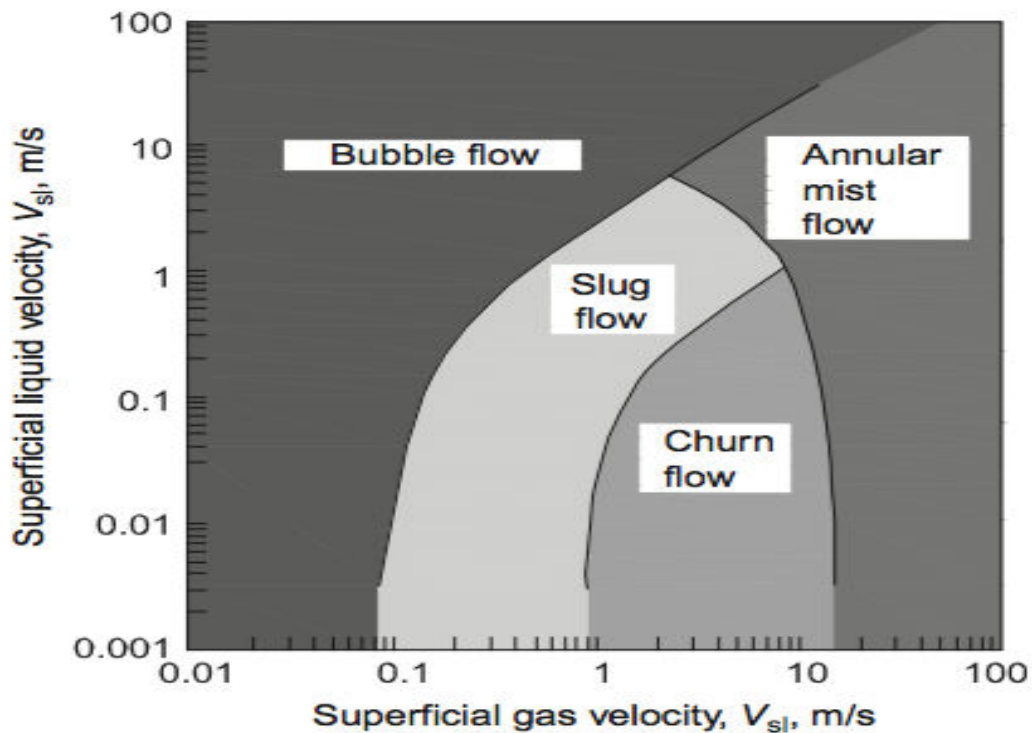
Figure 2.5 Vertical multiphase flow regime map [7]

3. PRESSURE DROP IN FLOWLINES

3.1 Flow Conditions

3.1.1 Flow Potential:

It is the total pressure drop that is required to transport fluid in a pipe section. i.e., the pressure difference between the inlet and the outlet of the pipe section. In an inclined pipe, this pressure difference also includes the pressure loss due to elevation. The flow potential should be equal or greater than that required to flow a given quantity of fluid in a pipeline of a specific diameter. Otherwise, a pump, compressor or a larger diameter pipeline would be needed.

3.1.2 Flow Regimes:

The basic types of flow in a pipe are laminar and turbulent flow. Laminar flow occurs at low velocities. At low velocities, fluid particles flow smoothly in one direction and there is little or no collision between themselves and the pipe walls. As the velocity increases, a critical point is reached when the flow changes from laminar to turbulent. The flow becomes chaotic and fluid particles collide with each other and with the walls of the pipe. This critical point depends on the fluid density and viscosity, pipe diameter and velocity of flow.

In laminar flow, the pipe internal roughness has no effect on the pressure drop, and the pressure drop is inversely proportional to the viscosity of the fluid whilst for turbulent flow, viscosity has little effect on the pressure drop but the pipe roughness plays a very important role in determining the total the total pressure drop.

3.1.3 Reynolds number:

It is the ratio of the inertial forces to the viscous forces for a fluid (density, viscosity and velocity) flowing in a circular pipe (diameter, D). It is dimensionless.

$$Re = \frac{inertial}{viscous} = \frac{\rho v^2}{\mu v / D} = \frac{D v \rho}{\mu} \qquad (3.10)$$

In field units,

$$Re = \frac{1488 D V \rho}{\mu} = \frac{124 d V \rho}{\mu} = \frac{7728(SG)dV}{\mu} \qquad (3.12a)$$

In SI units,

$$Re = \frac{\rho D V}{\mu} \qquad (3.12b)$$

Where,

Re = Reynolds number, dimensionless,

ρ = density (lb/ft^3 or kg/m^3),

D = pipe ID (ft or m),

d = pipe ID (in or mm),

V = average velocity (ft/s or m/s),

μ = absolute viscosity (cP or Pa s),

(SG) = specific gravity of liquid relative to water

For Re < 2000, the flow is considered Laminar and when Re > 4000, the flow is turbulent. For 2000 < Re < 4000, the flow is called transitional flow. It becomes unstable and could be either laminar or turbulent.

For liquids, the Reynolds number can be expressed in terms of liquid flow rate as:

In Field units

$$Re_l = 92.1 \frac{(SG)Q_l}{d\mu} \qquad (3.13a)$$

In SI units

$$Re_l = 353.13 \frac{(SG)Q_l}{d\mu} \qquad (3.13b)$$

Where,

μ = absolute viscosity (cP or Pa s),

d = pipe ID (in or mm),

(SG) = specific gravity of liquid relative to water,

Q_l = liquid flow rate (BPD or m^3/h).

For gases, the Reynolds number can be expressed as

Field units:

$$Re_g = 20000 \frac{Q_g S}{d\mu} \qquad (3.14a)$$

SI units:

$$Re_g = 0.428 \frac{Q_g S}{d\mu} \qquad (3.14b)$$

Where,

Q_g = gas flow rate (MMSCFD or std m^3/h),

S = specific gravity of gas relative to air,

μ = absolute viscosity (cP or Pa s),

d = pipe ID (in or mm).

3.1.4 Pipe Roughness

In laminar flow, the pressure loss is independent of the pipe wall roughness. In turbulent flow, the pressure drop is highly dependent on the pipe wall roughness and therefore affects the pressure gradient. As the velocity increases, the fluid boundary increasingly becomes thinner exposing the irregularities in the pipe wall to the higher velocity outside the boundary layer [1].

3.1.5 Flow Rate

The higher the flowrate, the higher the pressure loss due to friction. The capacity of the pipe also depends on the expected flow rate and velocity at design stage and its allowable pressure drop (flow potential).

3.1.6 Velocity Limitations

Increased fluid flow rate can lead to increased velocity, which can cause pipe damage. The avoidance of pipe damage sets an upper limit on the capacity of the pipe [1]. One criterion used to estimate the critical fluid velocity above which pipe damage may occur is found in API RP 14E, which suggests that a critical erosional velocity is expressed as:

$$Ve = \frac{C}{\sqrt{\rho}} \tag{3.15}$$

Where,

ρ = mixture density (lbm/ft^3),

Ve = erosional velocity threshold (ft/s),

C = 125 for intermittent service,

= 100 for continuous service

= 60 for corrosive service,

3.1.7 Temperature

Fluid temperature affects the pressure drop-flow rate relations and the density of the fluid, which can have an impact on the erosional viscosity limitations.

3.1.8 Deposits and Pipe Damage

This topic is thoroughly discussed in chapter 4.

3.1.8 Phase Flow regimes

Generally, in production operations, two-phase flow exists from the wellbore to the first separator in the production facility. Gas from the separator is considered as single-phase flow even though entrained liquids can be present. Liquid from the separator is also considered as single-phase flow even though gas is present due to pressure drop through the liquid dump valve. Details for the two-phase flow regimes have been discussed in the previous chapter.

1.1.9 Networks

Piping systems are usually constructed of segments with different diameters interconnected with manifolds. Some pipe systems may be so complicated that, empirical correlations are not applicable and computer software simulation should be used.
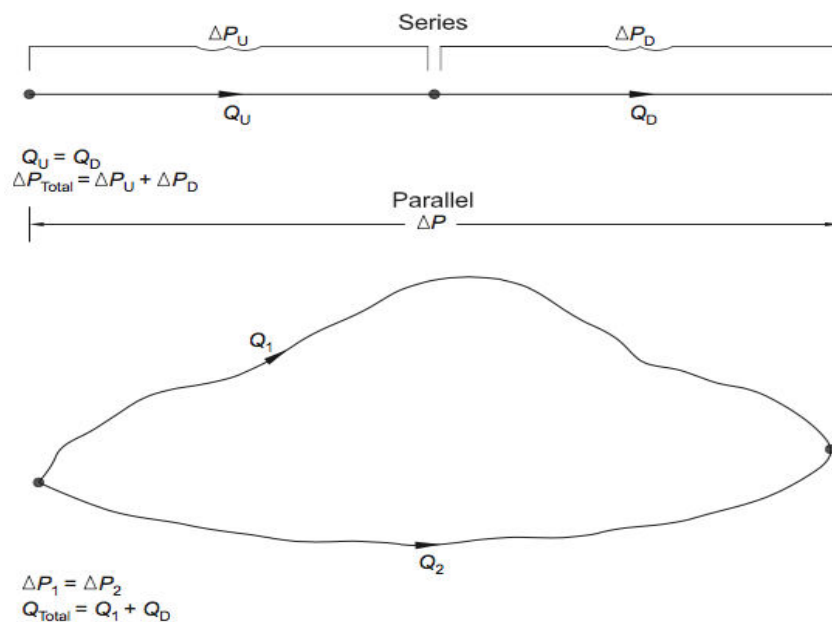
Figure 3.1 Series and parallel piping systems

For pipes of different diameters in series, the flow rate is the same, but the pressure drop is different. Therefore, the total pressure drop is determined by calculating the pressure drops for each individual pipe and then summing them up.

For pipes in parallel, the total flow rate is equal to the sum of the flow rates through each of the individual pipes but the pressure drops are identical for each segment.

3.2 Fluid Head

The term "head" is used to represent the pressure as the vertical height of a static column of liquid. A liquid at any point consists of the following three kinds of head:

Pressure Head: The static pressure head represents the energy contained in the fluid by virtue of its pressure. Its unit is feet or meters.

Velocity Head: It is the kinetic energy possessed by the fluid by virtue of its velocity. It is usually very small, although it is technically incorrect not to include it.

Potential Head: The potential energy contained in a fluid due to its position measured by the vertical elevation above a certain reference level. It can also be defined as the head due to the potential energy and the amount of that required to move a fluid from one elevation to another.

3.3 Friction Losses

Friction losses depend on the following:

- Flow rate
- Diameter
- Type of pipe (relative roughness)
- Length of pipe
- Number and size of fittings, valves and accessories
- Entrance and exit losses

3.4 Bernoulli's Equation

It states that, as a fluid flows from one point to another in a pipe (upstream to downstream), the sum of the static pressure, potential and velocity heads at the upstream point is equal to the sum of the three heads at the downstream point plus the head loss due to friction between the two points.
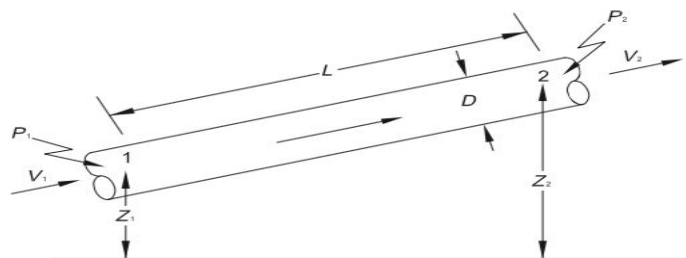
Figure 3.2 Terms used in the momentum equation

$$(H_s)_1 + (H_p)_1 + (H_v)_1 = (H_s)_2 + (H_p)_2 + (H_v)_2 + H_f \tag{3.16}$$

Where,

H_s = static pressure head (ft or m),

H_p = potential head (ft or m),

H_v = velocity head (ft or m),

H_f = frictional head loss (ft or m).

For a pipe section carrying a fluid of constant properties, the equation can be written as:

Field units

$$Z_1 + \frac{144 P_1}{\rho_1} + \frac{V_1^2}{2g} = Z_2 + \frac{144 P_2}{\rho_2} + \frac{V_2^2}{2g} + Hf \tag{3.17a}$$

SI units

$$Z_1 + 1000 \frac{P_1}{\rho_1} + \frac{V_1^2}{2g} = Z_2 + 1000 \frac{P_2}{\rho_2} + \frac{V_2^2}{2g} + Hf \tag{3.17b}$$

Where,

V_1 = upstream fluid velocity (ft/s or m/s),

V_2 = downstream fluid velocity (ft/s or m/s),

P_1 = upstream pressure (psi or kPa),

P_2 = downstream pressure (psi or kPa),

Z_1 = upstream elevation (ft or m),

Z_2 = downstream elevation (ft or m),

H_L = frictional head loss (ft or m),

ρ = fluid density (lb/f^3 or kg/m^3)

g = acceleration due to gravity (32.17 ft/s^2 or 9.81 m/s^2).

The velocity in the above equation is the average fluid velocity and is determined from the steady-state flow equation below:

Field units:

$$V = \frac{Q}{A} = \frac{W_s}{A_P} \tag{3.18a}$$

SI units:

$$V = \frac{Q}{3600 A} = \frac{W_s}{A_P} \tag{3.18b}$$

Where,

V = average velocity (ft/s or m/s),

Q = flow rate (ft^3/s or m^3/s),

A_p = cross-sectional area of the pipe (ft^2 or m^2),

W_s = mass flow rate (lb/s or kg/s).

3.5 Pressure Drop Equations

The Bernoulli equation includes the frictional head loss term H_f, which needs to be calculated. The two main equations used for this calculation are the Darcy-Weisbach and the Fanning equations.

- Darcy-Weisbach's equation

$$H_f = \frac{f_m L V^2}{D \times 2g}$$
(3.19)

- Fanning equation

$$H_f = \frac{4 f_f L V^2}{D \times 2g}$$
(3.20)

Where,

H_f = frictional head loss (ft or m),

f_m = Moody friction factor, dimensionless,

f_f = Fanning friction factor, dimensionless,

L = pipe length (ft or m),

D = pipe ID (ft or m),

V = average velocity (ft/s or m/s),

g = acceleration due to gravity (32.17 ft/s^2 or 9.81 m/s^2).

Neglecting the head difference due to elevation and the velocity changes between the two points, the Bernoulli's equation can be written as:

Field units:

$$P_1 - P_2 = \Delta P = \frac{\rho}{144} H_f$$
(3.21a)

SI units:

$$P_1 - P_2 = \Delta P = 9.81 \times 10^{-3} \rho H_f$$
(3.21b)

Where,

ΔP = pressure drop (psi or kPa),

H_f = pipe friction head loss (ft or m),

ρ = density of liquid (lb/ft^3 or kg/m^3).

Substituting equation 3.19 into 3.21a, the above equation becomes:

Field units:

$$\Delta P = 0.0013 \frac{f \rho L V^2}{d} \tag{3.22a}$$

SI units:

$$\Delta P = 0.5 \frac{f \rho L V^2}{d} \tag{3.22b}$$

where, d = pipe inside diameter (in or mm)

3.5.1 Moody and Fanning Friction factor comparison

Both friction factors are determined experimentally. The moody friction factor is generally accepted and used in pressure drop calculations [1] and will therefore be used in this text. Comparison of the two pressure drop equations reveals that the Darcy-Weisbach and Fanning equations differ only by a factor of 4. This occurs because of the differences in friction factors [1]:

$$f_f = 1/4 f_m \tag{3.23}$$

The friction factor is different for both Laminar and turbulent flows.

3.5.2 Friction factor for laminar flow

For laminar flow (Re < 2000), the friction factor is directly related to the Reynolds number and is expressed as:

Field units:

$$f_m = \frac{64}{Re} = 0.52 \frac{\mu}{dV\rho} \tag{3.24a}$$

SI units

$$f_m = 64 \frac{\mu}{dV\rho} \tag{3.24b}$$

If equation (3.24) is substituted into the equation (3.22), the pressure drop in psia and kPa can be calculated as:

Field units:

$$\Delta P = 0.000676 \frac{\mu L V}{d^2} \tag{3.25a}$$

SI units:

$$\Delta P = 32 \frac{\mu L V}{d^2} \tag{3.25b}$$

Equation (3.25) is used to determine the pressure drop between two points in a piping system for Laminar flow.

3.5.3 Friction factor for turbulent flow

The friction factor for turbulent flows (Re > 4000) is not only related to the Reynolds number, but also to the relative pipe roughness. Pipe relative roughness (ε/d) is equal to the pipe absolute roughness ε, divided by the inside diameter of the pipe d. Roughness is a measure of the smoothness of the pipe's inner surface. Table 3.1 shows the absolute roughness ε, of various new clean pipes.

Table 3.1 Absolute roughness (ε) of various new clean pipes.

Pipe roughness			
Type of pipe	Absolute roughness		
New, clean condition	mm	ft	inches
Unlined concrete	0.30	0.001 – 0.01	0.012 – 0.12
Cast iron – uncoated	0.26	0.00085	0.0102
Galvanized iron	0.15	0.0005	0.006
Carbon steel	0.046	0.00015	0.0018
Fiberglass epoxy	0.0076	0.000025	0.0003
Drawn tubing	0.0015	0.000005	0.00006

For pipes that have been in service for some time, it is often recommended that the absolute roughness to be used for calculations should be up to four times as much as the values in table above.

The pressure drop between two points in a piping system for turbulent flow can be determined from equation (3.22) using the friction factor from the Moody diagram below (figure 3.3).
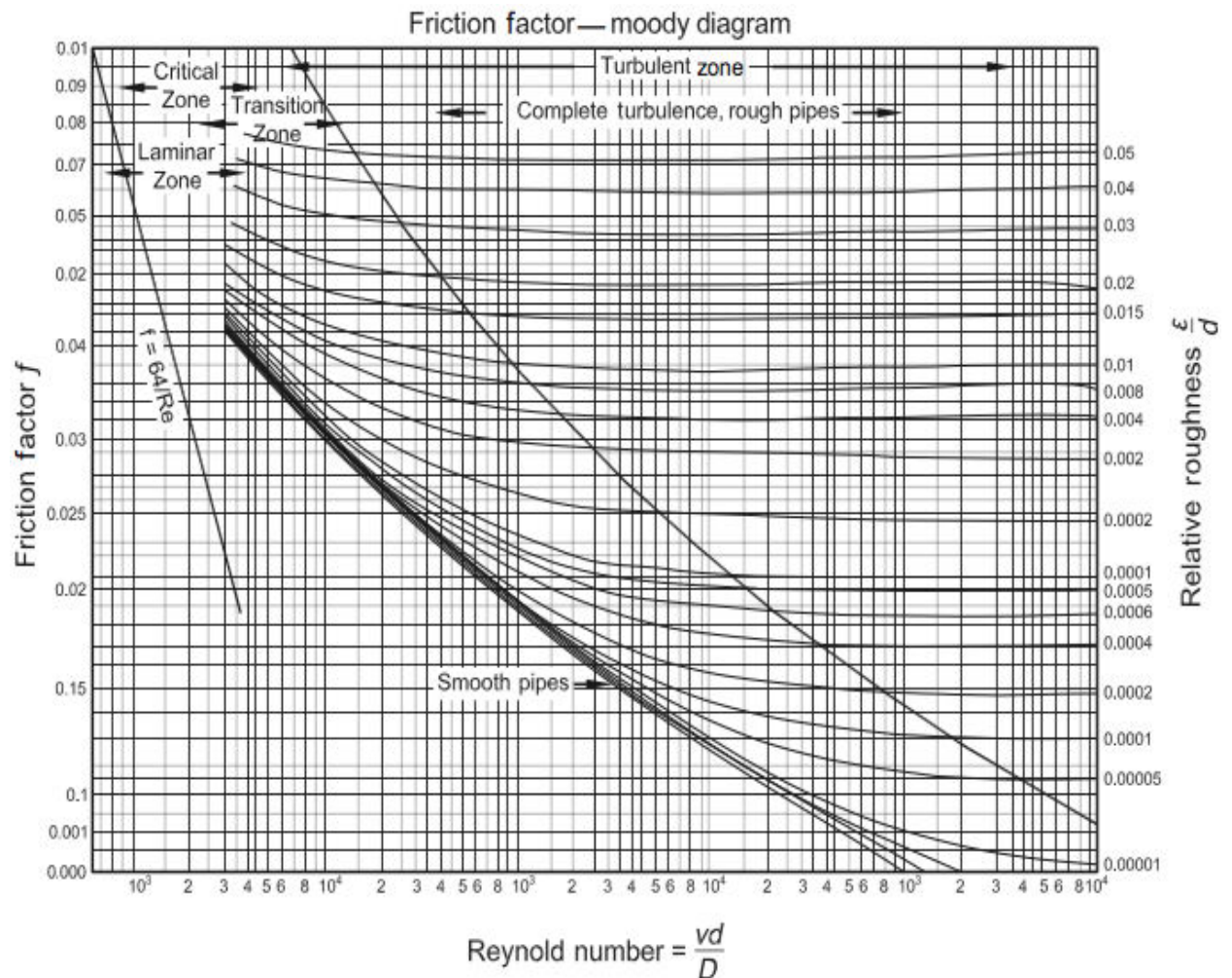
Figure 3.3 Friction factor as a function of pipe roughness and Reynolds number.

(Courtesy API)

NOTE:

- For Laminar flow, the velocity profile is parabolic and all the friction factor curves collapse into a straight line.

- For turbulent flow, the velocity profile is more uniform, the friction factor curves are more spread out and almost horizontal which indicates that, the friction factor is more of a function of the relative roughness.

- The transitional flow (2000 <Re< 4000) represents an unstable region between laminar and turbulent flow and the friction factor is undetermined.

3.6 Two-phase flow

Examples of two-phase flow are found in pipelines transporting natural gas and condensate or crude oil and the associated gas. These conditions occur when

fluids are transported in a pipeline at conditions that are:

- below the bubble point of the liquid,
- above the lower dew point of the gas.

Two-phase flow is further complicated by the presence of a second liquid phase, such as [1]:

- water with oil,
- gas or glycol with gas and condensate stream.

The design of a pipeline for two-phase or multiphase flow is complex because the properties and the flow patterns of two or more fluids must be taken into consideration.

3.6.1 Factors that affect two-phase flow

When designing a two-phase pipeline, the following factors must be considered:

- Liquid volume fraction
- Pipeline profile
- Flow regime
- Liquid holdup
- Two-phase pressure loss

Knowing how much gas and liquid is present in a hydrocarbon mixture is essential in designing a two-phase flow pipeline. The two-phase gas-liquid composition of a hydrocarbon mixture is best illustrated by a phase envelope. The phase envelope is a graphical representation of the relative quantities of gas and liquid present in a mixture at various pressures and temperatures. Figure 3.4 is an example of a typical phase envelope of a gas-condensate reservoir. It illustrates the pressure-temperature operating profiles for a pipeline with varying flow rates. The phase envelope shows the liquid percentage in the fluid at each point. For example, the liquid percentage changes from about 3% at the inlet to about 5% at the outlet [1].
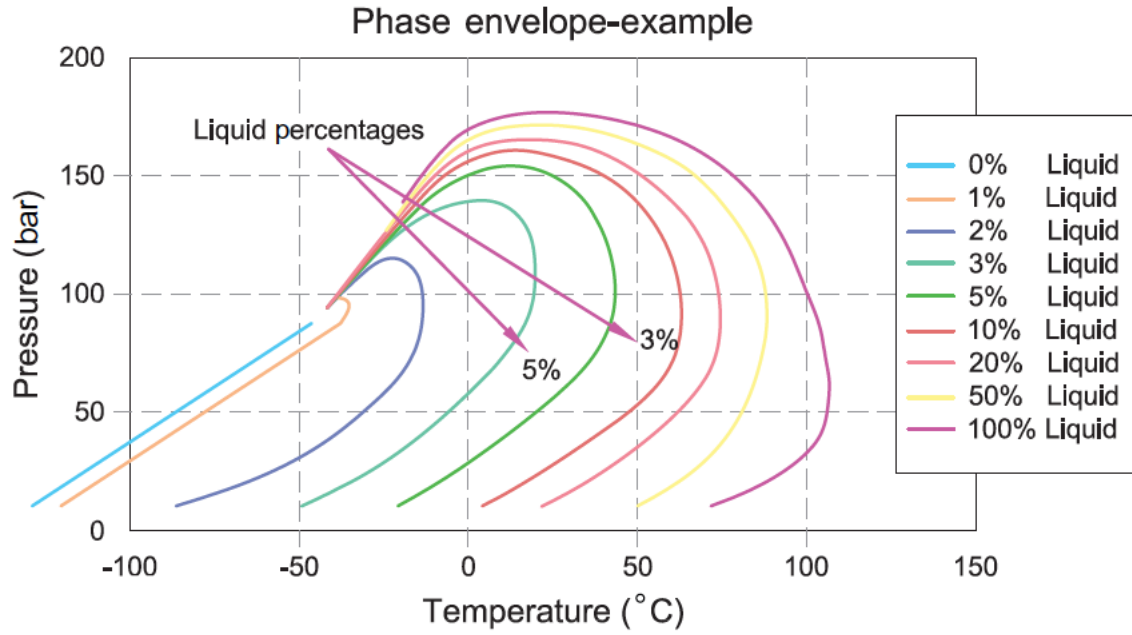
Figure 3.4 Typical phase envelope of a gas-condensate reservoir [8].

3.6.1.1 Liquid volume fraction

The liquid volume fraction is the fraction of fluid flowing as a liquid present at a local point in the pipeline. It is determined by assuming that, thermodynamic equilibrium exists between the two phases at the local pressure and temperature [1].

$$V_{SG} = v_G \cdot H_G \qquad V_{SL} = v_L \cdot H_L \qquad H_L + H_G = 100\%$$

(3.26)

Where,

V_{SG} = superficial gas velocity,

V_{SL} = superficial liquid velocity.

v_L: the actual velocity of the gas phase [ft/s]

v_G: the actual velocity of the gas phase [ft/s]

H_L: the hold-up fraction of the liquid phase [%]

H_G: the hold-up fraction of the gas phase [%]

The superficial liquid velocity is the velocity the liquid would have in the pipeline, if only the liquid phase was present. Likewise, the superficial gas velocity is the velocity the gas would have, if only the gas phase was present [1].

3.6.1.2 Liquid holdup

The liquid holdup is the accumulation of liquid in the pipeline due to the

difference in velocity, "slip," between the phases and the influence of gravity. In a two-phase pipeline, the designer must predict the amount of liquid expected to be present. An understanding of how the liquid holdup in a pipeline will vary with flow rates will greatly assist in establishing a suitable liquid management system. This could include a slug catcher, pigging facilities, or operational limits on ramp-up times. When designing a two-phase pipeline, a graph, similar to that shown in Figure 3.6, is usually generated showing how liquid holdup varies with throughput and with varying pipeline diameters [1].

Hold-up H_L
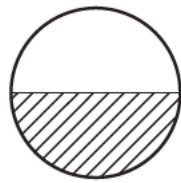
$\frac{A_L}{A}$ = H_L = liquid hold-up

where
A   = total cross-sectional area
A_L = liquid cross-sectional area

Figure 3.5 Liquid hold-up in a pipeline. [1]

Figure 3.6 Liquid hold up vs. gas throughput [1]

3.6.1.3 Pipeline profile

Changes in elevation of the pipeline route play an important role in two-phase flow pipeline design and operation. The elevation profile influences the liquid holdup and determines the flow regime. This in turn influences the pressure loss incurred during the two-phase flow [1].
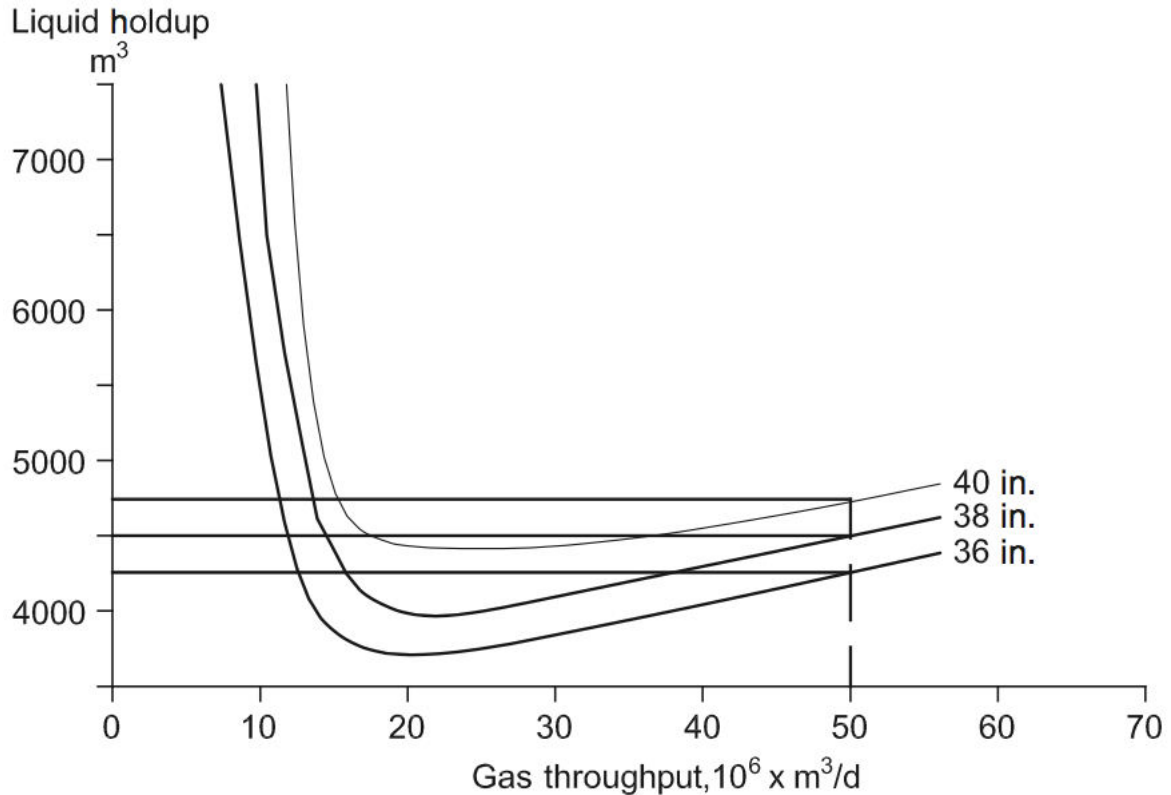
3.6.1.4 Two-phase Flow regime

It is discussed in detail in chapter 2.

3.6.1.5 Two-phase pressure loss

The pipeline pressure loss in a two-phase flow is higher than that of a single-phase. The increased pressure loss is caused by the additional friction loss due to the formation of waves and slugs [1].

The calculation of pressure drop in two-phase flow is very complex and is mostly based on empirical relationships that take into account the phase due to pressure and temperature changes along the flow, the relative velocities of the phases, and the complex effect of elevation [7]. Correlations and simulation softwares are generally employed to predict pressure loss in a pipeline for two-phase flow at varying levels of uncertainty.

The formula below, presented in the American Petroleum Institute's Recommended Practice API RP 14E, provides an approximate solution for friction pressure drop in two-phase flow problems that meet the assumptions stated below.

Field Units

$$\Delta P = \frac{3.36 \times 10^{-6} fL W_h^2}{\rho_m d^5} \qquad \text{(Eqn 3.27a)}$$

SI units

$$\Delta P = 62.561 \frac{fL(W_h^2)}{\rho_m d^5} \qquad \text{(Eqn 3.27b)}$$

Where,

ΔP = pressure drop (psi or kPa),

L = length of pipe (ft or m),

W_h= flow rate of liquid and vapor (lb/h or kg/h),

ρ_m= mixture density (lb/ft^3 or kg/m^3),

d = pipe inside diameter (in or mm),

f = Moody friction factor.

The flow rate of the mixture to be used in this equation is calculated as:

Field Units

$$W_h = 3180Q_gS + 14.6Q_1(SG)$$ (Eqn 3.28a)

SI units

$$W_h = 1.21Q_gS + 999.7Q_1(SG)$$ (Eqn 3.28b)

Where,

W_h = flow rate of liquid and vapor (lb/h or kg/h,

Q_g = gas flow rate (MMSCFD (std or m³/h),

Q_1 = liquid flow rate (BPD or (m³/h),

S = specific gravity of gas relative to air,

SG = specific gravity of liquid relative to water.

The density of the mixture to use in equation (3.27) is given by:

Field Units

$$\rho_m = \frac{12,409(SG)P + 2.7RSP}{198.7P + RTZ}$$ (Eqn 3.29a)

SI Units

$$\rho_m = \frac{28,814(SG)P + 34.81RSP}{28.82P + 10.0RTZ}$$ (Eqn 3.29b)

Where,

ρ_m = mixture density (lb/ft³ or kg/m³),

P = pressure (psia or kPa),

SG = specific gravity of the liquid relative to water (the average gravity for the hydrocarbon and water mixture),

S = specific gravity of gas relative to air,

R = gas-liquid ratio (std ft³/bbl or std m³/m³),

T = temperature (°R or °K),

Z = gas compressibility factor, dimensionless.

The formula is applicable if the following conditions are met:

- ΔP is less than 10% of the inlet pressure.
- Bubble or mist exists.
- There are no elevation changes.
- There is no irreversible energy transfer between phases

3.7 Head loss in valves and pipe fittings

The pressure drop equations described above are useful for calculating the pressure drop-flow rate relations in straight pipes. Additional pressure drop occurs through valves, pipefittings, and enlargements and contractions. For piping systems within production facilities, the pressure drop through fittings and valves can be much greater than that of the straight pipe itself. In long pipeline systems, the pressure drop through fittings and valves can be often ignored [1].

A pipe restriction that changes velocity or direction of the flow stream causes pressure drops greater than that, which would normally occur in a straight piece of pipe of the same length. The additional frictional effects can be evaluated using an extension of the Darcy-Weisbach equation. The extension involves the determination of either of the following [1]:

- Resistance coefficients for fittings
- Flow coefficients for valves
- OR equivalent lengths for both valves and fittings

3.7.1 Resistance coefficients

The head loss in valves and fittings can be calculated with resistance coefficient as:

$$H_f = K_r \frac{V^2}{2g} \qquad \text{(Eqn 3.30)}$$

Where,

K_r = resistance coefficient, dimensionless,

H_f = head losing fitting (ft or m),

V = average velocity (ft/s or m/s),

g = acceleration of gravity (32.2 ft/s^2 or 9.81 m/s^2).

Comparing equation (3.30) with the Darcy-Weisbach equation shows that, for a straight pipe,

$$K_r = \frac{fL}{D} \qquad \text{(Eqn 3.31)}$$

Where,

K_r = fitting resistance coefficient, dimensionless,

f_m = Moody friction factor, dimensionless,

L = length (ft.),

D = pipe inside diameter (ft.).

Total system frictional losses in valves and fittings can be determined by summing up the head loss in all fittings, as defined by the following equation [1]:

$$\sum H_{Lf} = \sum K \frac{V^2}{2g}$$ (Eqn 3.32)

Where,

$\sum H_{Lf}$ = total system frictional losses in valves and fittings measured in feet (ft).

Once the resistance coefficients for the fittings and valves have been determined, the frictional head losses due to the fittings and valves can be calculated and then added to that of the straight runs of pipe.

Approximate values of K_r are given in Table 3.2 for various pipefittings. Figure 3.7 (APPENDIX) shows the resistance coefficients for sudden contractions and enlargements and for pipe entrances and exits. Figure 3.8 (APPENDIX) is a listing of representative resistance coefficients from Crane Technical Paper No. 410 [1].

Table 3.2 Resistance coefficients for pipefittings

Fitting	Kr
Globe valve, wide open	10.0
Angle valve, wide open	5.0
Gate valve, wide open	0.2
Gate valve, half open	5.6
Return bend	2.2
Tee	1.8
90°	0.9
45°	0.4

3.7.2 Flow coefficients

The flow coefficient of a valve or fitting is a relative measure of its efficiency at allowing fluid flow. It describes the relationship between the pressure drop across the valve and the corresponding flowrate [8].

The flow coefficient for liquids, Cv, is determined experimentally for each valve or fitting as the flow of water in gpm at 60 °F for a pressure drop of 1 psi through the fitting. The relationship between flow and resistance coefficients can be expressed as follows:

Field Units

$$C_V = \frac{29.9d^2}{(\frac{fL}{D})^{\frac{1}{2}}} = \frac{29.9d^2}{(K_r)^{\frac{1}{2}}}$$ (Eqn 3.33a)

SI units

$$C_V = \frac{0.0105d^2}{(\frac{fL}{D})^{\frac{1}{2}}} = \frac{0.0105d^2}{(K_r)^{\frac{1}{2}}}$$ (Eqn 3.33b)

Where,

Cv = flow coefficient (gpm or m³/h),

D = fitting equivalent ID (ft or m),

d = fitting equivalent ID (in or mm),

L = fitting equivalent length (ft or m),

K = resistance coefficient, dimensionless.

For any valve with a known Cv, the pressure drop can be calculated for different conditions of flow and liquid properties with the equation below:

Field Units

$$\Delta P = 8.5 \times 10^{-4}(\frac{Q_l}{C_V})^2(SG)$$ (Eqn 3.34a)

SI Units

$$\Delta P = 6.89(\frac{Q_l}{C_V})^2(SG)$$ (Eqn 3.34b)

Where,

Q_l = liquid flow rate (BPD or m³/h),

ΔP = pressure drop (psi or kPa),

SG = specific gravity of liquid relative to water,

Cv = flow coefficient (gpm or m³/h).

3.7.3 Equivalent lengths

The head loss associated with valves and fittings is calculated by considering equivalent "lengths" of pipe segments for each valve and fitting. It is often simpler to treat valves and fittings in terms of their equivalent length of pipe. The equivalent length of a valve or fitting is the length of an equivalent section of pipe that has the same frictional pressure drop as that of the fitting. Total pressure drop can then be determined by adding all equivalent lengths to the pipe length [1]. The equivalent length, L_{eq}, can be determined from K_r and C_V as follows:

Field Units

$$L_e = \frac{K_r D}{12 f_m} \quad \text{and} \quad L_e = \frac{74.5 d^5}{f_m C_V^5} \qquad \text{(Eqn 3.34a)}$$

SI Units

$$L_e = \frac{K_r D}{1000 f_m} \quad \text{and} \quad L_e = \frac{1.108 \times 10^{-7} d^5}{f_m C_V^2} \qquad \text{(Eqn 3.34b)}$$

Where,

Kr = resistance coefficient, dimensionless,

D = pipe diameter (ft or m),

d = pipe diameter (in or mm),

Cv = flow coefficient (gpm or m^3/h),

f_m = Moody friction factor, dimensionless,

L_e = equivalent length of valve or fitting under turbulent flow (ft or m).

For laminar flow, equivalent length can be calculated from the following [1]:

$$L_{e(LAMINAR)} = \frac{Re}{1000} L_e \qquad \text{(Eqn 3.35)}$$

$L_{e(LAMINAR)}$ = equivalent length to be used in pressure drop calculations (never less than actual fitting length),

L_e = equivalent length of the valve or fitting if the flow were turbulent.

Table 3.3 (APPENDIX) summarizes the equivalent lengths of various commonly used valves and fittings. Figure 3.9 (APPENDIX) shows equivalent lengths of 90° bends, and Figure 3.10 (APPENDIX) summarizes equivalent lengths of fabricated bends of different radii [1].

4. DEPOSITS AND PIPE DAMAGE

4.1 Hydrates

Hydrates are ice-like solids, which are formed when gas molecules of lower molecular weight form hydrogen bonds with water molecules at certain range of temperatures and pressures. For hydrates to form, light gas molecules like methane, ethane, propane, CO_2 and H_2S should be present with water at high pressures and low temperatures. The range of temperatures and pressures at which hydrates can form is determined from testing fluid samples in the lab, or from correlations and simulation tools. Hydrate formation poses the largest challenge in multiphase flowlines and can therefore result in:

- complete blockage of pipeline and flow line.
- production downtime.
- time and cost for remediation

Hydrate formation is most likely to occur under the following operations:

- Shutdown: Shutdown operations cause pipeline temperature to drop, as heat is lost to the environment. Temperature drop well within the hydrate region can cause hydrates to form.
- Subsea operations: At the seabed, temperatures are so low that, unrealistically low pressures are required to prevent the formation of hydrates.

Hydrate curve

Figure 4.10 below shows the "hydrate curve" for a natural gas. The curve depends on the gas composition. It can be seen that, outside the hydrate region, hydrates do not form and the opposite occurs for the non-hydrate region.
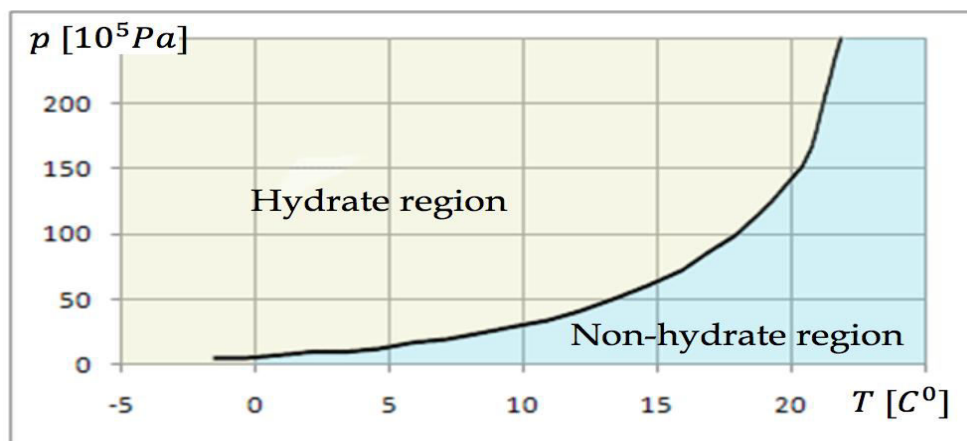
Figure 4.10 Typical mixture hydrate curve for a natural gas (mixture) [2].

Figure 4.12 also shows a typical hydrate curves for pure gases. The hydrate curve is affected by both the hydrocarbon and water composition.
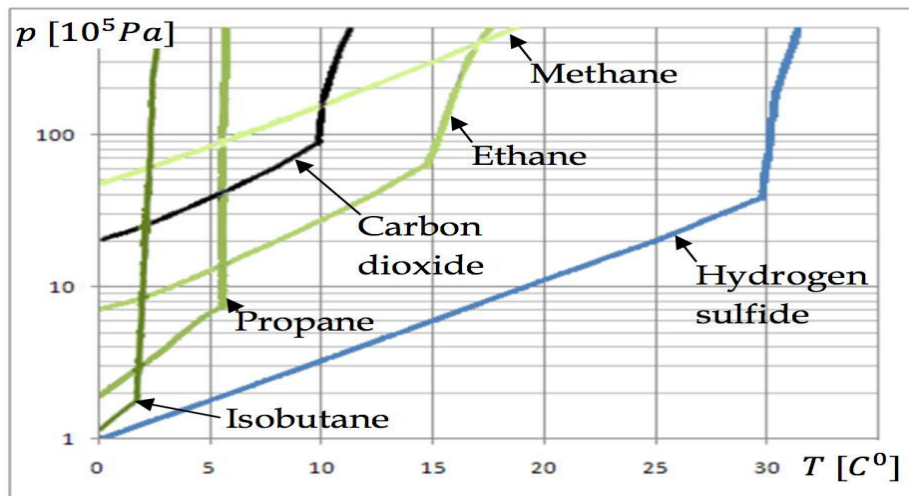
Figure 4.11 Hydrate curves for various pure components [2].

Prevention of hydrate blockage

- *Injection of hydrate inhibitors:* Such as methanol and mono-ethylene glycol (MEG) enlarges the non-hydrate region or shifts the hydrate curve to the left. Figure 4.12 shows the results of adding methanol to water. The methanol percentage is calculated as the ratio of the mass flow rate of methanol to the combined mass flow rate of both methanol and water. Gas flow is not included in the calculation. It should be noted that, any methanol which is not mixed with the water does not take part in the prevention of hydrate formation and therefore poor mixing reduces the effect of methanol.
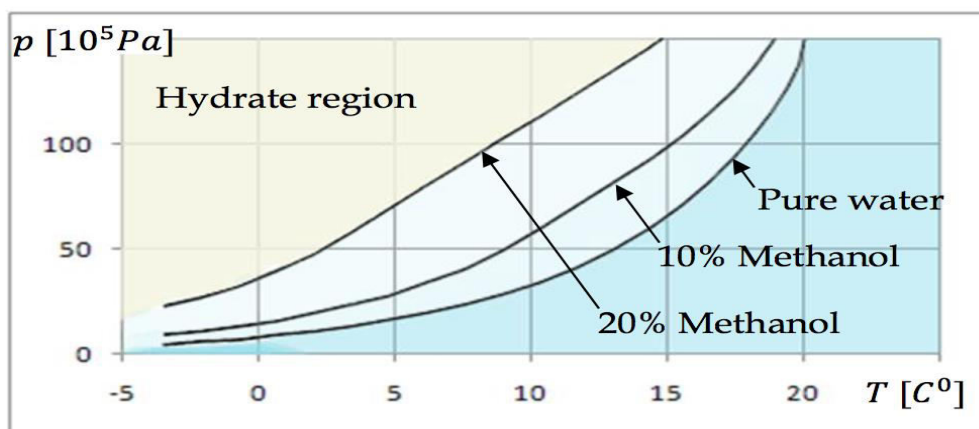
Figure 4.12 Hydrate curves for various amounts of methanol inhibition for pure methane [2]

Addition of salts will have the same effect as that of methanol and MEG but there is a high risk of pipe corrosion and therefore, it is not recommended.

- *Temperature control*: The main purpose is to prevent the fluid from cooling down. This is done by thermal insulation of the pipe to prevent loss of heat to the surroundings and to keep the interior temperature of the pipe above the hydrate formation limit. Heat can also be added from an external source. Example, hot water pipes embedded in the flowline bundle, electrical heating etc.

- *Injection of "anti*-agglomerant" and "kinetic inhibitors": They do not prevent the formation of hydrates. The kinetic inhibitors delay the formation of hydrates for a period of at least 24 to 48 hours. They do not usually work when the temperature falls more than 10 ^{0}C below the one for hydrate formation, and they have little effect after the maximum delay time. Anti-agglomerate inhibitors prevent the formulation of large hydrate crystals. This prevents hydrates from sticking to the wall and makes it easier for the crystals to be carried away if the flow velocity is high enough.

- **Short shutdown periods** or injection of extra inhibitors in case of expected long shutdown periods.

- **Cold Flow Technology (CFT)**: The main concept is the controlled formation of hydrates and the prevention of destructive blockages. The Cold Flow Technology is based on the conclusion of scientific studies and observations [9] that the hydrates do not accumulate when the temperature is kept constant. When the flow is entering the pipeline, hydrates start to form, because the ambient temperature is low. Before the hydrates start to create agglomerates and depositions, the flow is separated in gas and liquid phase. The liquid phase is cooled down in a heat exchanger reaching a final temperature similar to the ambient one and then it is boosted to the reactor, where it is mixed again with the gas phase. By this way, the hydrates are formed but they do not accumulate to create blockages in the flow as they remain in suspension since there is no heat convection between the ambient environment and the flow in the pipeline.

Reduction techniques of hydrates already formed in the pipelines

- *Heating up the flow*: If heat is provided to the flow, the temperature will increase, which will cause the hydrogen bonds that are responsible for the formation of the hydrates to loosen up and eventually break.

50

- *Depressurization***:** Dissociation of the hydrate plug by reducing the system pressure causes the gas and the solid water molecules to liquefy. The pressure reduction must be done simultaneously and uniformly from both sides of the hydrate plug otherwise the pressure difference between the sides of the hydrate plug might cause violent acceleration and cannon-ball like behavior when the plug comes loose, leading to the destruction of the pipe.

4.2 Waxes

Wax is class of hydrocarbons with high carbon numbers (C18 to C60) which are soluble in oil. They are heavier components of crude oil and occur as natural constituents. The formation of waxes in pipelines creates many problems in production. The following factors affect the formation of waxes in pipelines:

- *Low Temperature***:** Wax precipitation occurs on the walls of pipelines as the fluid cools due to lower temperature at the wall. The temperature below which wax starts to form at a specific pressure is called "*wax cloud point*" or "*wax appearance temperature*" (WAT). WAT is different for each hydrocarbon mixture. It is one of the important parameters when characterizing wax deposition. Shutdown may also cause loss of temperature as the oil cools down.

- *Fluid compositio***n**: Waxy crudes (crude oil with high wax content) have high tendency of forming waxes in pipelines. The higher the heavier compounds, the higher the wax content.

- **Flow rate:** High flow rates prevent the deposition and accumulation of wax particles on the wall of the pipeline as they are swept away by the turbulence.

- **Pipeline internal roughness:** High pipeline internal roughness creates a good platform for wax deposition, while smooth surfaces or surfaces wetted by oil hinder wax deposition.

- Pressure drop: When oil flows out of the well, the lighter components evaporate as the pressure reduces. Thus, the heavier components in the remaining oil become less soluble and increase the tendency of wax formation.

Effects of Waxes Formation

- *Increase in viscosity*: Reduction of temperature leads to increased viscosity of the oil. Increase in viscosity leads to high friction between the oil and the pipe wall, which causes huge pressure losses.

- *Blockages***:** When wax crystals form on the wall, they tend to trap oil and form a wax-oil gel. Further decrease in temperature causes the wax layer to grow and trap more oil, which reduces the effective diameter of the pipe. If no appropriate action is taken, this may lead to blockage.

Prevention / Reduction of wax formation:

- *Injection of wax inhibitors***:** They are chemical surfactants that inhibit the formation of waxes by reducing the WAT by as much as $10^{o}C$.

- *Injection of chemical solvents***:** They increase the solubility of any formed wax crystals and re-dilute them.

- *Temperature control:* By Pipeline insulation, heating or steam tracing the pipeline.

- *Regular pipeline pigging:* If regular pigging is not done, wax accumulation can become large so that the pig might get stuck, especially in relatively small diameter pipes. Wax also hardens with time and therefore it is easier to take out after a short period from when it is formed. During pigging, the oil trapped in the wax in front of the pig is squeezed out which also hardens the wax and increases the possibility of the pig getting stuck.

    If a pipe of diameter d and length l, has a layer of wax d_{wax}, then the wax volume would be approximately $\pi dd_{wax}l$ and the length of the wax plug $l_{wax\ plug}= \frac{4dwaxl}{d}$. This means that, pigging a pipeline of 200 mm diameter with a 1.2 mm wax layer will lead to the wax plug length of 24m for each kilometer pigged. The equation above also explains why pigs easily get stuck for relatively smaller diameter pipes compared to the larger ones. Larger diameter leads to shorter wax plug length per kilometer of pigging for the same wax layer thickness [2].

4.3 Asphaltenes

Asphaltenes are defined as the heaviest oil compounds that are insoluble in n-pentane and n-hexane but soluble in benzene and toluene. They are solids of dark brown or black colour and they are one of the four main compound groups of oil

(saturates, aromatics, resins and asphaltenes). Asphaltene precipitation depends on the oil's composition, pressure and temperature regime.

Asphaltene precipitation increases as the pressures changes towards the bubble point and reaches its maximum at the bubble point. Above the bubble point, its precipitation decreases because, asphaltene molecules become soluble in the oil and therefore remain as part of the oil as a single phase.

Problems arise when the asphaltenes deposition reduces the effective cross sectional area of the pipelines and the oil production is either reduced or completely stopped.

Asphaltene precipitation prevention is similar to that for waxes: By keeping the pressures and temperatures at values where no precipitation is possible, by injecting additives, or by cleaning the pipes with pigging. Asphaltenes formulate harder deposits than waxes, so the pigs must be specifically designed to cope with the challenge. The determination of the suitable additives to inject is more difficult for asphaltenes than for hydrates or waxes so it is recommended to test them on the specific oil in question (rather than trying to predict it from the oil's composition) before using them [2].

4.4 Scales

Scales form from the inorganic chemicals present in produced water. The main and most common scales are inorganic salts like barium sulphate ($BaSO_4$), strontium sulphate ($SrSO_4$), calcium carbonate ($CaCO_3$), though some partly organic scales may also be present (naftenates, MEG-based etc [2].

Scales can be grouped into two main categories namely: sulphate scales and carbonate scales.

Sulphate scales: They are formed when incompatible waters mix. (Seawater and formation water). This normally occurs at gathering networks when production from different wells mix. Production brines from these wells may contain different minerals, which can result in a chemical reaction and the deposition of scales. It can also occur during water injection when minerals of the seawater react with minerals of the formation water to produce sulphate scales.

Carbonate scales: Carbonate scales are formed because of pressure decrease and increase in the pH of the water. When pressure decreases, dissolved CO_2 in the

formation water evaporates causing an increase in pH (decrease in acidity). This makes some minerals less soluble leading to the formation of insoluble salts.

Scale formation can increase the surface roughness, reduce the cross-sectional area of the pipe or even lead to complete blockage. It can also cause problems to valves, pumps, and other components in the flow-path [2], as well as corrosion beneath the deposits.

Scale formation can be prevented by continuous injection of chemical inhibitors. Sample analysis of produced water is required to know its scale forming tendencies and to select the proper type of inhibitors. Scales already formed in the pipelines can be removed by regular pigging.

4.5 Corrosion

Corrosion is the abruption of metal particles from the surfaces that are in contact with the flow, due to chemical or electrochemical reactions. The main causes of corrosion are the reaction of the metal surfaces of the equipment with carbon dioxide (CO_2), hydrogen sulfide (H_2S) and water (H_2O).

The corrosion that may appear on the oil and gas facilities can be categorized to external and internal corrosion, according to the parts of the equipment where it may be developed.

External corrosion is caused by the oxidation of the steel when it is exposed to air. As a result, the metal surface loses its strength and becomes more brittle. If the phenomenon continues for a long time, the corrosion can extend to the interior of the pipeline and may cause the collapse of the pipeline.

Internal corrosion is caused by the chemical reaction of some of the transported fluid ingredients and the steel material of the pipelines' walls or equipment.

According to the nature of the corrosive agents, the following corrosion types may appear:

CO_2 in the flow (sweet corrosion): CO_2 may exist in the flow as one of the non-hydrocarbon compounds of oil or due to direct CO_2 injection during enhanced oil recovery purposes. When it comes in contact with water, it reacts, causing the formation of carbonic acid (H_2CO_3). The corrosion due to CO_2 is caused by the electrochemical reaction between the carbonic acid and the iron of the metallic

surface of the pipeline [10].

H₂S in the flow (sour corrosion): H_2S may appear in the flow as one of the non-hydrocarbon compounds of oil. As with the CO_2corrosion, the prerequisite for sour corrosion is the presence of water in the flow. Corrosion due to H_2S is caused by the chemical reaction between the hydrogen sulphide, water and the iron of the metallic surface of the pipeline [10]. The result of this reaction is the creation of iron sulphide (FeS), which is observed as a film on the surface of the pipe.

O₂ in the flow (oxygen corrosion): O_2 may appear in the flow through leakages in pumps' seals, open hatches and process vents [11]. It reacts with the metallic surfaces causing its oxidization.

Galvanic corrosion: galvanization is referred to the phenomenon of the electrochemical reactions between the molecules of two metallic surfaces with different electrochemical potentials, when they are in contact. The result of those reactions is the charging of the surfaces with opposite electric loads thus altering their chemical structure. The material properties change due to the change of the intergrannular chemical bonds, resulting into corrosion damage.

Crevice corrosion: is a type of corrosion that appears on the surface of the metallic equipment due to the concentration of different corrosive agents in these spots. It appears in crevices and gaps of the equipment where fluid may get stagnant [11].

Microbiological corrosion: Microorganisms such as bacteria are accumulating and create colonies in reservoirs and in spots of equipment where stagnant water is concentrated. Common products of the digestion procedures are CO_2, H_2S and organic acids that cause corrosion on the metallic surfaces [11].

Stress Corrosion Cracking (SCC): is a combination of corrosion and tensile stress failure. It occurs when the material is reaching its fatigue thresholds and cracks start to propagate to the rest of the material meanwhile the corrosive environment contributes to the expansion of the cracking [11].

Factors that affect (internal) corrosion

- ***Flow temperature:*** high pH, high temperatures enhance the formation of iron carbide on the surface of steel equipment, which creates a corrosion layer on the equipment [11].

- ***High pressure***: under high pressures the solubility of $CO_2$and H_2S in water increases,

so the environment becomes more corrosive.

- *pH of the environment*: The pH of a flow is indicative of the number of cations (positively electrically charged molecules) of hydrogen (H^+) in the flow, which contribute to the neutralization of the electrochemical potential. High pH is connected to the low solubility of iron carbonate ($FeCO_3$). Hence, iron carbonate does not dissolve fast and can accumulate on the metallic surfaces, as a protective layer, lowering by this way the corrosion rate [10].

Effects of corrosion

Corrosion has no direct effect on the flow but affects the properties of the pipe's material. As the corrosion keeps spreading in the surface of the metallic equipment, the properties of the material change. More specifically the steel loses its elasticity and becomes more brittle. The impact of corrosion accumulates with time and may even lead to the complete failure of the material and its breakage.

Prevention

- *Use of corrosion inhibitors:* They are chemicals that reduce the rate of corrosion by forming a protective layer on the pipe wall to prevent direct contact with the corrosion agents or neutralize the effect of the corrosion agents.
- *Using corrosion resistant material*s for making of the pipes. Eg. steel alloys.
- *Galvanization:* i.e. coating the internal and external surfaces of the pipe with protective coatings especially zinc, (which is highly resistant to corrosion) to form a protective layer to prevent corrosion.

4.6 Erosion

Erosion is a common problem that is met when fluids, that contain solid particles, flow over metallic surfaces. Erosion damage is caused by the solid particles that are transported with the effluents flow or from the liquid droplets that are detected in the flow. Solids in the flow may come from sand, proppants and deposits of hydrates, asphaltenes, waxes and scales. Erosion is referred to the abruption of particles from the surfaces that are in contact with fluid flow that contains solids. The erosive damage affects the protective oxide film of the metallic surfaces and the result is observed as the creation of grooves, pits and other types of deformation of the

metallic surfaces [12]. The main erosion mechanisms are the following [13]:

- *Liquid droplets impingement*: In case of multiphase flow or wet gas flow, liquid droplets may exist in the flow at high gas velocities and volumetric flow rates. The extension of erosion in this case depends on the droplet size, impact velocity, impact frequency, type of liquid, gas density and viscosity [12]. The constant collision of the liquid droplets on the walls of the pipeline may cause material fatigue and abrasion [13].

- **Solid particles impingement**: in case of multiphase flows that contain solids, the solid particles may be deposited on the bottom of the pipe or they may be transported as dispersed flow in other fluids. When they collide with the pipeline walls, they gradually cause material fatigue and abrasion [13].

Factors that affect erosion

- *The velocity of the solid particles*: the higher the velocity of the solid particles, the rougher the surface will become and thus the more severe the erosion.

- *The shape and size of the solid particles*: the size of the solids particles affects the erosion rate up to a certain size (50 to 1000μm) while the particles with spherical shape contribute less to the erosion damage than the ones with angular shapes [10].

- *Angle of attack*: it has been observed that solid particles that collide on the metallic surface with non-orthogonal angle of incidence have greater contribution to the deformation of the surface [10].

- *Flow rate:* when the volumetric flow rate increases, more liquid and solid particles will collide on the walls of the pipe, which results in increased erosion damage [13].

- *Restrictions in the flow*: the existence of restrictions along the pipeline indicates that downstream of them the flow will exhibit high velocities. In case of sand or liquid droplets in the flow, they will collide on the walls of the pipeline with greater velocity, thus increasing the erosion damage.

- *Sudden changes of flow direction*: when the pipeline network has sharp elevation changes, the sand particles and/or liquid droplets in the flow may collide at the points of elevation change on pipeline's walls, resulting in their abrasion.

Effects of erosion

Related studies and experiments have concluded that the eroded metals

become more brittle [13]. When it is combined with external corrosion, the impacts are more severe and may become destructive for the pipeline. The corroded particles that have created a layer on the metallic surface are removed by the transported solid particles thus destroying the corrosion layer, which acts as a surface protection, preventing the corrosion from intruding further into the surface [11].

The components of the pipeline network that are more vulnerable to erosion are the following [12]; chokes, sudden restrictions, partially closed valves, check valves and valves that are not full bore, standard radius elbows, weld intrusions and pipe bore mismatches at flanges, reducers, long radius elbows, miter elbows, blind tees and straight pipes etc.

Prevention techniques

- *Modeling the solids flow*: There are several correlations about the dependence of the erosion rate and the solids flow that can be used in combination with flow modeling [15]. This measure implies the control over the production flow rate.
- *Use of improved pipe materials:* The pipelines are made of alloys with improved resistance on erosion. The alloys UNS N06625 and UNS S32750 are some of the materials used for subsea applications [14].
- *Sand removal*: Downhole sand screens and gravel packs or sand separation devices are used to remove the sand from the flow [12].

Erosion can intensify the corrosion effect. This is because the iron carbonate scale and/or inhibitor layer that is normally slowing down the corrosion process is less wear-resistant than the steel itself and tends to get more easily removed by erosion, thus allowing corrosion to accelerate the damage to the pipeline. The same problem goes for internal coatings: Erosion can remove the coating and accelerate corrosion. In pipelines carrying refined gas to customers, the coating should be able to resist erosion for a long time. In flowlines, however, it comes off more rapidly. It is difficult to select the appropriate internal coating and to predict how long it lasts in flowline [2].

4.7 Cavitation

Cavitation is caused by gas bubbles collapsing nearly instantly when the pressure suddenly increases. When the velocity increases, for instance due to reduced cross-section, the pressure is reduced. Local high velocities inside valves and pumps

can cause the pressure to fall down to boiling pressure. Once the velocity is reduced again some distance downstream, the bubbles collapse. This collapse can occur at the low pressure side of valves and pumps, in the pipe immediately downstream of a restriction [2].

Cavitation can cause small pieces of the surface to be knocked off and in time, weaken the surface where the bubbles collapse. When inspected, the cavitated surface may look as if somebody had been hacking on it with a needle. Pressure surges due to valve opening or closure, or pump startup or shutdown can also cause cavitation [2].

4.8 Water hammer/Hydraulic Shock

Hydraulic shock (or water hammer) is a common problem that occurs in hydraulic installations of single or multiple phase flow pipelines. Usually flow under these conditions is called "surge flow". It is referred to a sudden pressure drop or increment in the flow. Usually sudden pressure fluctuations are met when sudden closing of non-return valves (valves that allow the flow through them only by one direction) and chokes occur upstream of the flow or when a pump starts or stops suddenly. Under these violent pressure fluctuations, the fluid flow stops abruptly. The change of its kinetic energy creates a great pressure difference that is induced to the fluid as wave that propagates in the inverse direction of the flow. This phenomenon is met in flows that have at least one liquid phase, because the liquids are highly incompressible, so the surge pressure deviations are more intense [15].

Effects of hydraulic shock
Pipeline rupture: The pressure waves produce vibrations, which are transformed to sound waves. The vibrations affect the strength of the pipeline. The forces applied on the walls of the pipelines are big enough to lead to the pipeline's burst [15].
Well completion damage, unset packers due to the great pressure surge.
Leakage at pipeline's joints due to the high forces that are imposed on the joints.
Damage to pumps downstream of the flow.

Operations and facilities during which hydraulic shock may occur
Downward inclination of the pipeline: The damage and burst of the pipeline may be greater if it is inclined downwards. In this case, the pressure difference that will be

created will be higher due to the additional hydraulic pressure [15].

Well Shut in and Start up: during these operations, both bottomhole and wellhead pressure increase and reduce respectively. The pressure fluctuations may lead to hydraulic shock occurrence if they are not controlled [15].

Cavitation: when a valve close or a pump stops, the pressure difference downstream creates an inverse flow backwards, towards the pump or the valve. In case that the pressure downstream the valve or pump decreases more than the bubble point of the mixture, gas bubbles will appear in the flow. When the gas bubbles meet the surface of the closed valve/pump, they will break resulting to an extended and violent pressure surge [16].

Prevention/Reduction techniques

The noise that is created as an effect of the sharp pressure fluctuations in the flow, by unexpected opening of pressure relief valves and when the flow shows pulse periodicity, are indications of the hydraulic shock occurrence [16]. When one of these indications in the flow occur the following reduction techniques may be applied:

Reduction of flow velocity: by using pipelines of larger diameter or by lowering the volumetric flow rate, the sudden pressure drop that may occur will not be harmful for the pipeline.

Surge tanks: they are used to absorb liquid in order to reduce the pressure surge. Examples are surge alleviators, pressure relief valves, Air inlet valves, Injection of nitrogen or air into the fluid.

5. SIMULATION AND RESULTS

5.1 Case Description

In this case, the effect of flowline diameter on oil production and pressure drop in pipelines is evaluated. The data used for the simulation is based on Example 2a of the GAP software of IPM SUITE, which is taken from an offshore oilfield at the Loggie Mill block.

The field has two reservoirs A and B. Reservoir A has been in production since 01/01/2003 and reservoir B was brought online five years later on 01/01/2008. Each reservoir is drained by three wells, which are then connected to two manifolds installed at seabed, approximately 600ft below sea level. Wells drained by reservoir A are connected to manifold A whilst that of reservoir B are connected to manifold B. A flowline of 7 km transports effluents from manifold A to manifold B. The commingled fluids from all 6 wells are then transported from manifold B through a larger 12 km pipeline to a separation facility. The separator is maintained at a pressure of 200 psig and its liquid capacity is set to 20,000 stb/day.

Figure 5.10 is a computer sketch of the offshore field at the Loggie Mill Block.
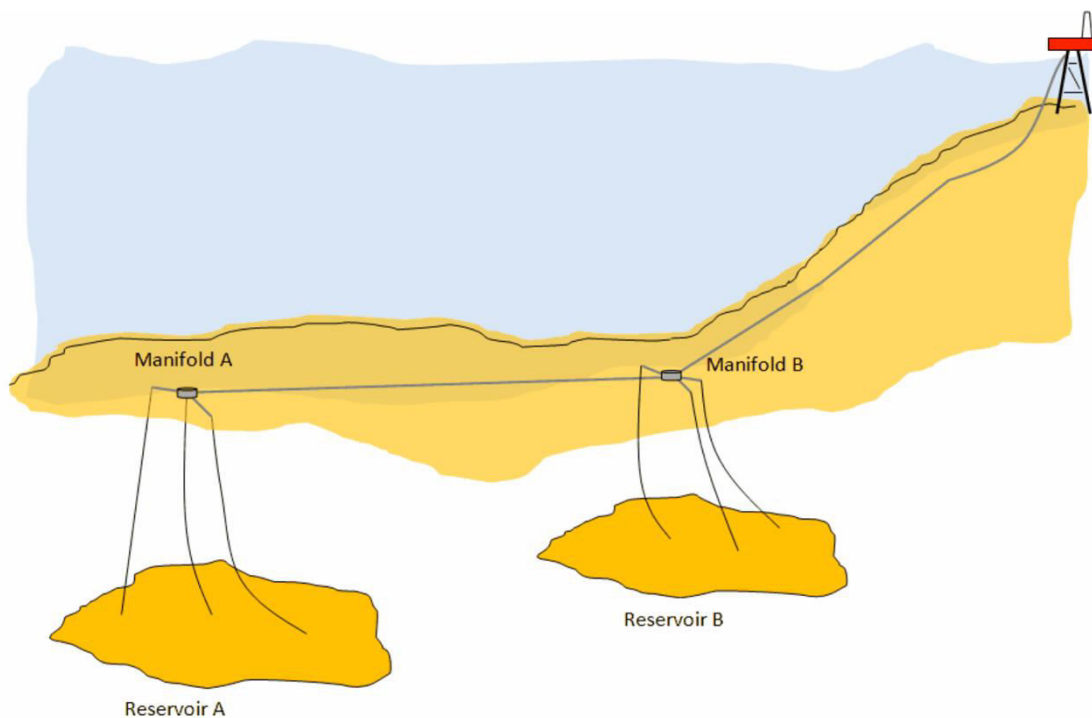

Figure 5.10 Sketch of the Loggie Mill Block's offshore field

5.1.1 Pipeline Simulation Softwares

Table 5.10 lists the current commercially available state-of-the-art pipeline design and operations software for hydraulic modelling and flow assurance for single-phase, two-phase and multiphase flow systems [1].

Table 5.10 Hydraulic modelling and flow assurance softwares

Multiphase Transient Simulation	Multiphase Steady-State Simulation	Single phase gas/liquid Transient Simulation
OLGA ProFES	Pipe flow, well flow	Pipeline Simulator Winflo
	PIPEPHASE	NATASHA PLUS
	PIPESIM-SUITE	TLNET
	GENNET-M	TGNET
	Flowsystem	
	Prosper, Gap	
	Perform	

The **GAP** software in combination with **PROSPER** and **MBAL**, all from the Petroleum Experts IMP Suite were used for this project.

GAP is a multiphase oil and gas optimiser tool that models the surface gathering network of field production systems. When linked with the well models of **PROSPER** and reservoir models of **MBAL** a full field production optimisation and forecast can be achieved. GAP can model production systems containing oil, gas and condensate, in addition to gas or water injection systems [17].

GAP's powerful non-linear optimisation engine allocates gas for gas lifted wells or sets wellhead chokes for naturally flowing wells to maximise revenue or oil/gas production while honouring constraints at any level in the system. GAP is used to perform the following operations [17]:

- Pipeline Flow Assurance Studies
- Multi-well optimisation for artificially lifted and naturally flowing production networks
- Optimisation of gathering system against constraints
- Reservoir Management
- Fully Compositional from the Reservoir to the Process side

The **MBAL** software was used to build the reservoir models of reservoir A and B described above using material balance after which the developed models were imported as input files into the **GAP** software to represent both reservoirs. The **PROSPER** software was also used to build individual well models for all six wells producing from both reservoirs after which the output files were also imported into the GAP software to represent the wells.

With both softwares introduced into GAP, GAP was used to perform a full pipeline simulation and flow assurance study with respect to all reservoirs and wells to observe their response when certain pipeline parameters like diameter is altered. More information about the well and reservoir models has been given below.

The surface network layout of the offshore field at the Loggie Mill Block described above was designed with the GAP software as illustrated in Figure 5.11.
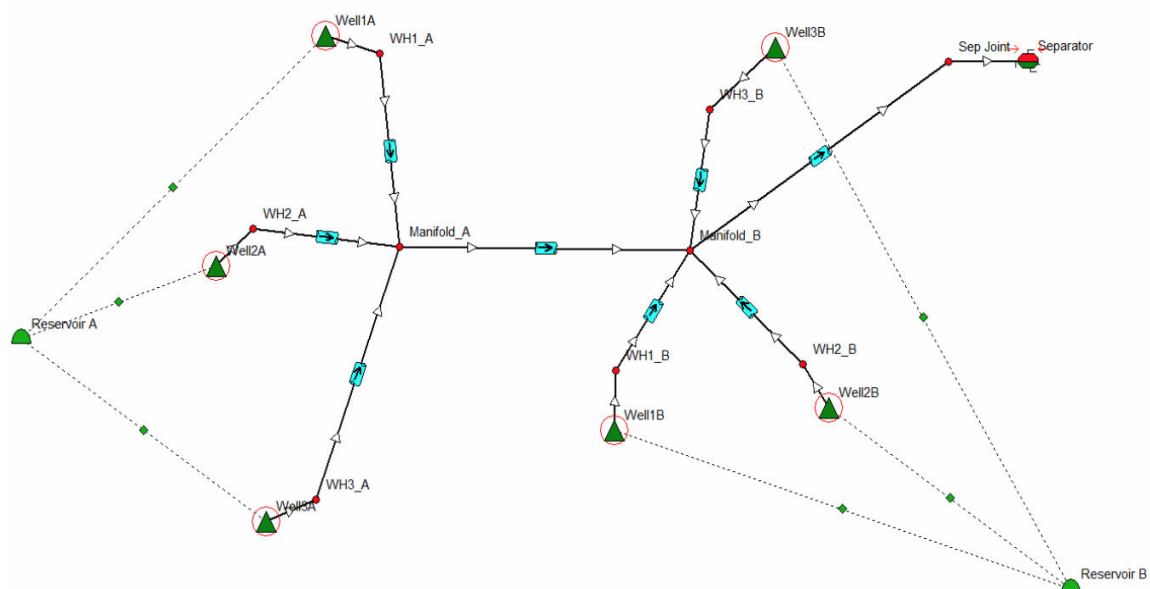

Figure 5.11 Surface network layout design of Loggie Mill Block's offshore field with GAP

The types of data that were used for this project are:

- Reservoir data
- Well Data
- PVT data
- Pipeline data
- Pipeline Environment Data

5.1.2 Reservoir Description

The offshore field at the Loggie Mill block has two reservoirs A and B. Both reservoirs are drained by 3 wells each. Production data was recorded for reservoir A from 01/01/2003 at the start of production to 22/12/2010 and that of reservoir B was recorded from 01/01/2008 to 01/12/2010. The Production histories for the reservoirs are provided in Tables 5.11a and 5.11b. The initial reservoir pressure was 4000 psia for both reservoirs. At the end of the production history, the average reservoir pressure was 3752 psig for reservoir A and 3542.19 psig for reservoir B.

Table 5.11a Production History of Reservoir B

Time	Reservoir Pressure	Cum. Oil produced	Cum. Gas Produced	Cum. Water Produced	Cum GOR
Date d/m/y	psig	(MMSTB)	MMscf	(MMSTB)	scf/stb
01/01/2008	4000	0.000	0	0	500
01/03/2008	3927.5	0.279	139.5	0	500
01/05/2008	3879.14	0.551	275.5	0.001	500
01/07/2008	3842.41	0.814	407	0.003	500
01/09/2008	3811.96	1.075	537.5	0.007	500
01/11/2008	3785.98	1.325	662.5	0.012	500
01/01/2009	3762.36	1.570	785	0.018	500
01/03/2009	3741.1	1.802	901	0.024	500
01/05/2009	3720.07	2.036	1018	0.032	500
01/07/2009	3699.8	2.266	1133	0.041	500
01/09/2009	3679.77	2.495	1247.5	0.052	500
01/11/2009	3660.2	2.717	1358.5	0.063	500
01/01/2010	3641.09	2.934	1467	0.075	500
01/03/2010	3623.04	3.141	1570.5	0.087	500
01/05/2010	3604.68	3.350	1675	0.1	500
01/07/2010	3586.62	3.555	1777.5	0.115	500
01/09/2010	3568.44	3.761	1880.5	0.13	500
01/11/2010	3550.71	3.960	1980	0.146	500
01/12/2010	3542.19	4.056	2028	0.154	500

Table 5.11b Production History of Reservoir A

Time	Reservoir Pressure	Cum. Oil produced	Cum. Gas Produced	Cum. Water Produced	Cum GOR
Date d/m/y	psig	(MMSTB)	MMscf	(MMSTB)	scf/stb
01/01/2003	4000	0	0	0	800
02/04/2003	3913	0.455	364	9.53E-05	800
02/07/2003	3925.39	0.91	728	0.000312	800
01/10/2003	3903.37	1.365	1096.1	0.000747	803.004
31/12/2003	3876	1.82	1456	0.000925	800
31/03/2004	3871.14	2	1820	1820	800
30/06/2004	3858.95	2.73	2181.27	0.00174	799
29/09/2004	3848.35	3.185	2548	0.00225	800
29/12/2004	3866	3.64	2912	0.0028	800
30/03/2005	3830.73	4.095	3276	0.0033	800
29/06/2005	3822.86	4.55	3635.45	0.0038	799
28/09/2005	3750	5.005	4004	0.0043	800
28/12/2005	3809.11	5.46	4373.46	0.0059	801
29/03/2006	3802.94	5.915	4743.83	0.0077	802
28/06/2006	3797.1	6.37	5096	0.0095	800
27/09/2006	3791.52	6.825	5460	0.0105	800
27/012/2006	3786.14	7.28	5824	0.014	800
28/03/2007	3780.92	7.611	6111.63	0.0161	803
27/06/2007	3775.82	7.821	6256.8	0.0187	800
26/09/2007	3770.82	8.231	6584.8	0.021	800
26/12/2007	3765.89	8.501	6800.8	0.023	800
26/03/2008	3761.01	8.9	7120	0.028	800
25/06/2008	3766	9.23	7384	0.035	800
24/09/2008	3765	9.654	7710.65	0.0467	798.7
24/12/2008	3763	10.354	8283.2	0.0646	800
25/03/2009	3762.5	10.521	8427.32	0.0789	801
24/06/2009	3760	10.7456	8596.48	0.0925	800
23/09/2009	3732.27	10.987	8800.59	0.111	801
23/12/2009	3759	11.213	8970.4	0.128	800
24/03/2010	3758	11.345	9076	0.1543	800
23/06/2010	3757	11.564	9285.89	0.178	803
22/09/2010	3787	11.79	9432	0.201	800
22/12/2010	3752	12.01	9632.02	0.22	802

MBAL was used to build separate models for each reservoir. In order to estimate the stock tank oil initially in place (STOIIP), history matching of production data was conducted with "material balance" method for both reservoir models. The

history matching was conducted against reservoir pressure and oil production. The results are shown on Figures 5.12a and 5.12b.

The initial assumption during the history matching was that, both reservoirs have no aquifer support. The MBAL results of this matching deviated substantially from the measured values as indicated by the red line on Figure 5.12. Therefore, an aquifer model (Hurst-van Everdingen-Modified) was introduced into the history matching and an acceptable match was obtained as indicated by the blue line on the same figure plots.
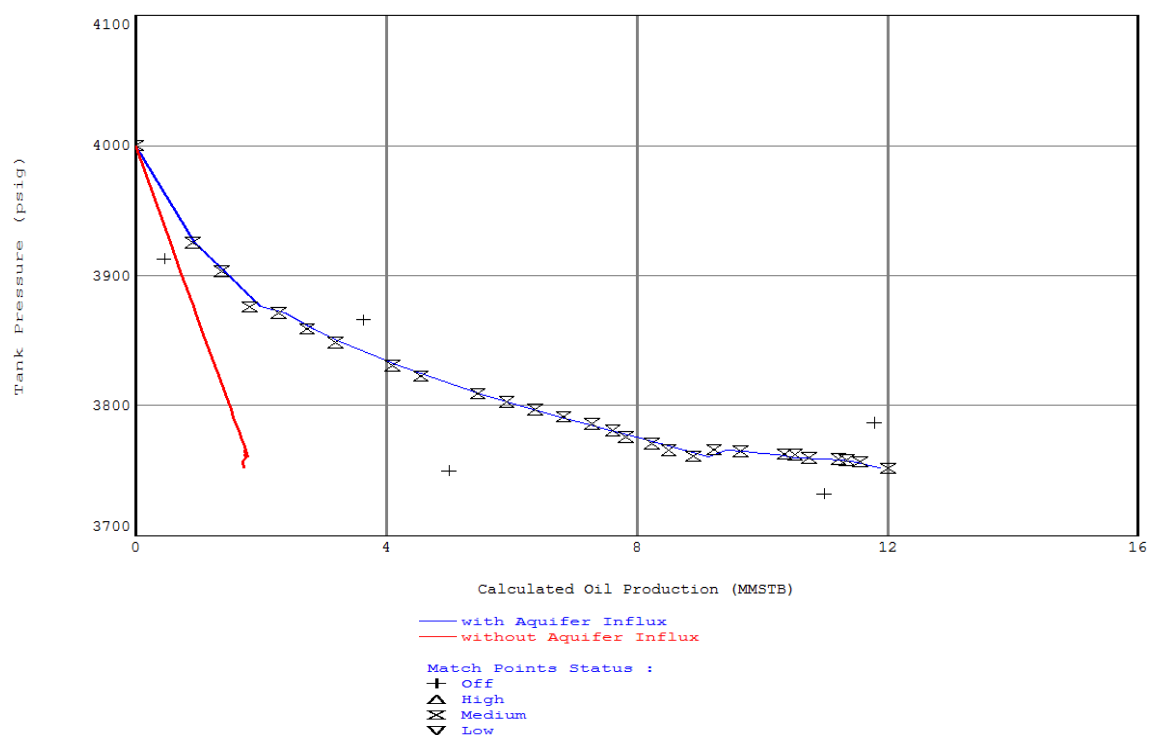

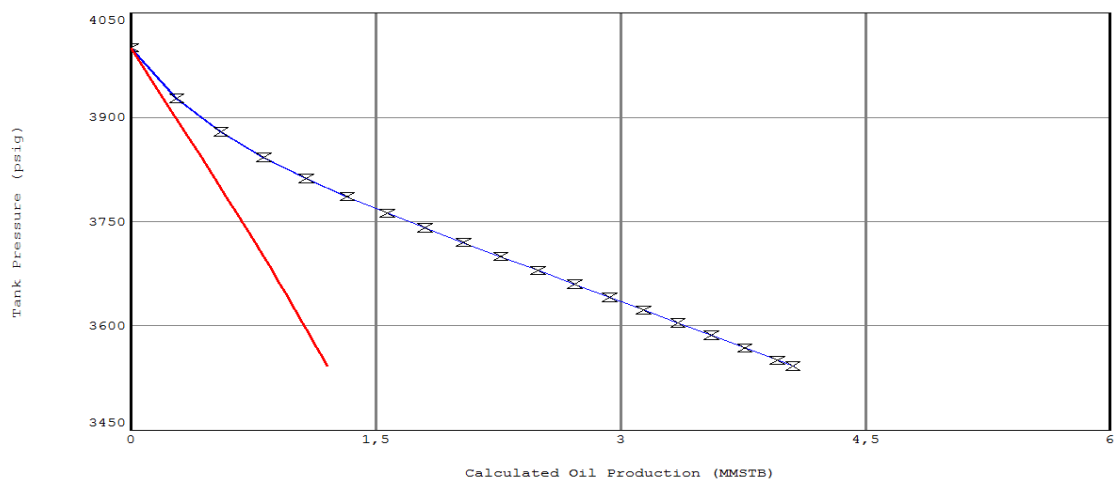Figure 5.12a History matching of production data for reservoir A

Figure 5.12b History matching of production data for reservoir B

After the matching, the STOIIP was estimated to be 376.95 MMsbbl for reservoir A and 238.95 MMsbbl for reservoir B.

Table 5.12 Information about the Reservoirs

Tank Name	History		Initial reservoir Pressure	Final reservoir Pressure	STOIIP
	Start	End			
	dd/mm/yyyy	dd/mm/yyyy	psia	psia	MMSTB
Reservoir A	01/01/2003	22/12/2010	4000	3752	375.95
Reservoir B	01/01/2008	01/12/2010	4000	3542.19	238.95

The tuned MBAL models of both reservoirs were then imported into GAP to represent the characteristics of the reservoirs.

5.1.3 Well Description

Fluid is produced from 6 wells. The main well data are shown in Table 5.13.

Table 5.13 Main Characteristics of the wells

Well Label	PI	Sand Control
Well1A	7.0348504	Gravel Pack
Well2A	7.1488697	Gravel Pack
Well3A	7.1486874	Gravel Pack
Well1B	0.685364	None
Well2B	0.685364	None
Well3B	0.721209	None

PROSPER was used to build the well models and to generate their IPR and VLP curves. The PROSPER output files for all six wells were then imported into GAP to represent the wells. GAP was then used to generate the wells' IPRs and VLPs. The red lines represent the IPRs and the blue lines represent the VLPs. These are shown on Figures 5.13 **a-f.**
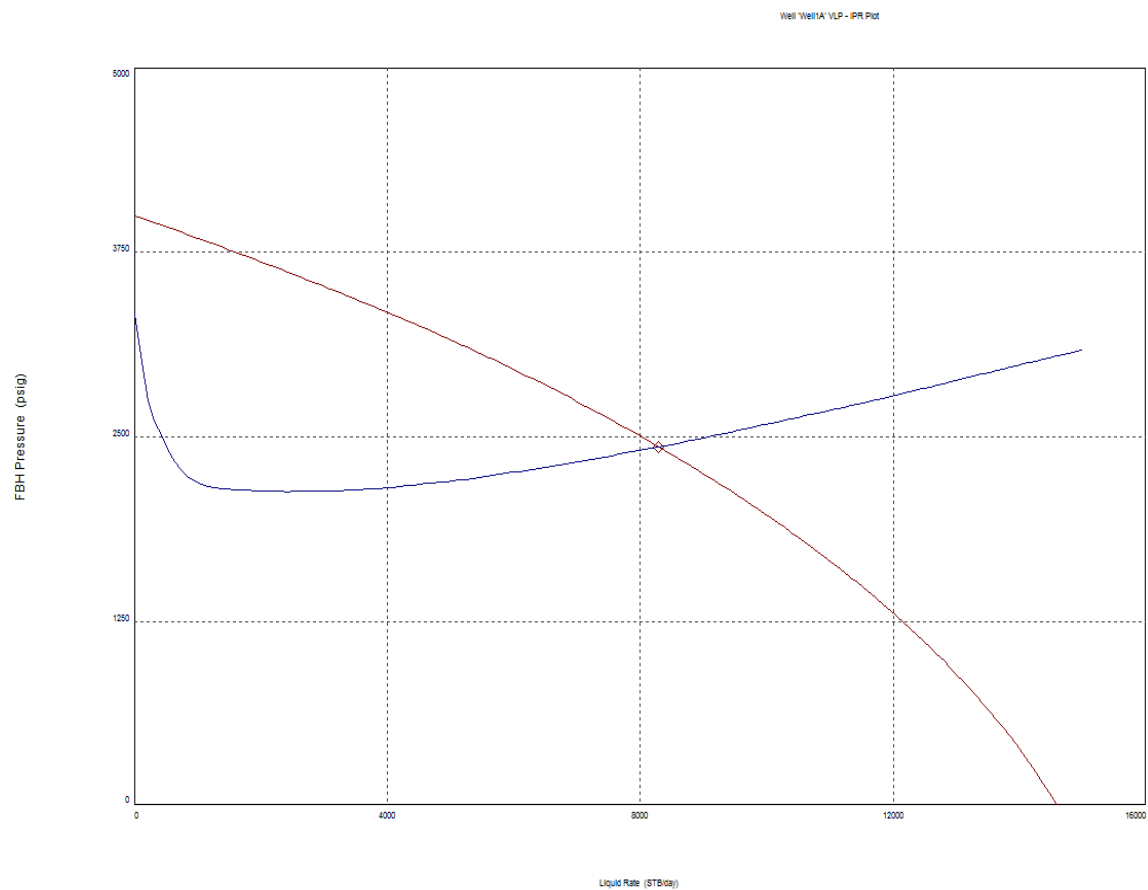
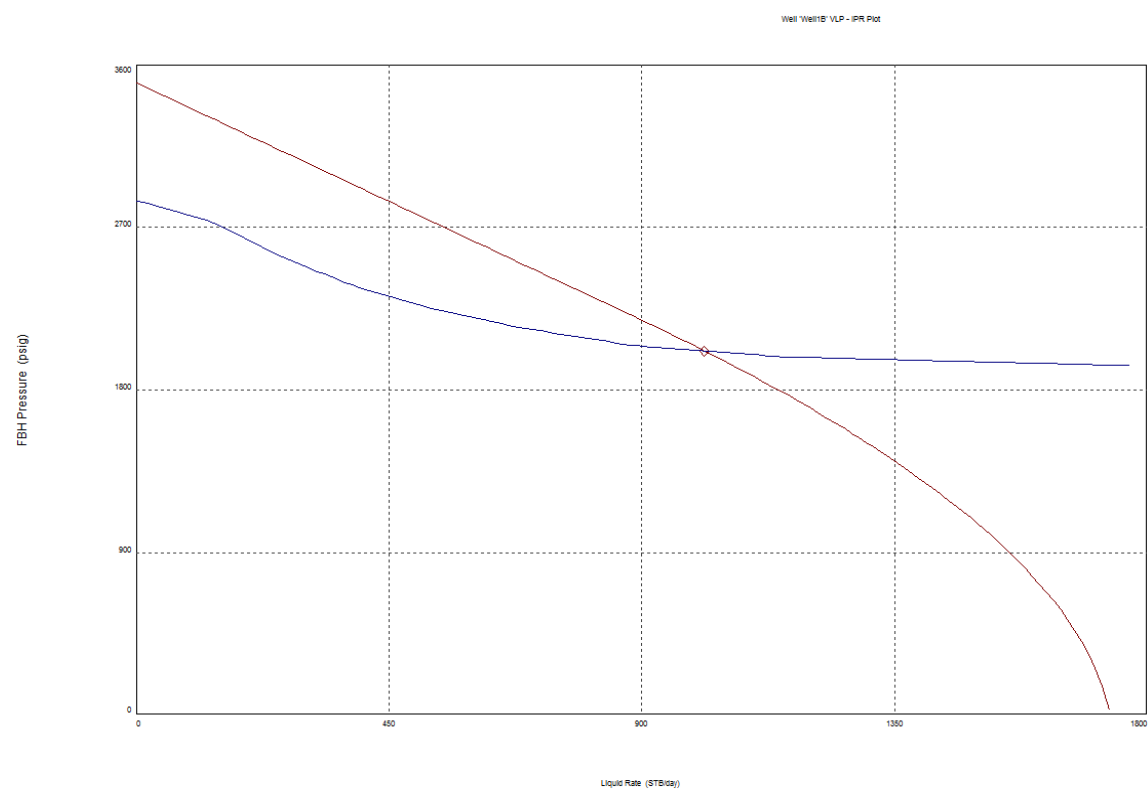Figure 5.13a IPR and VLP curve for Well1A

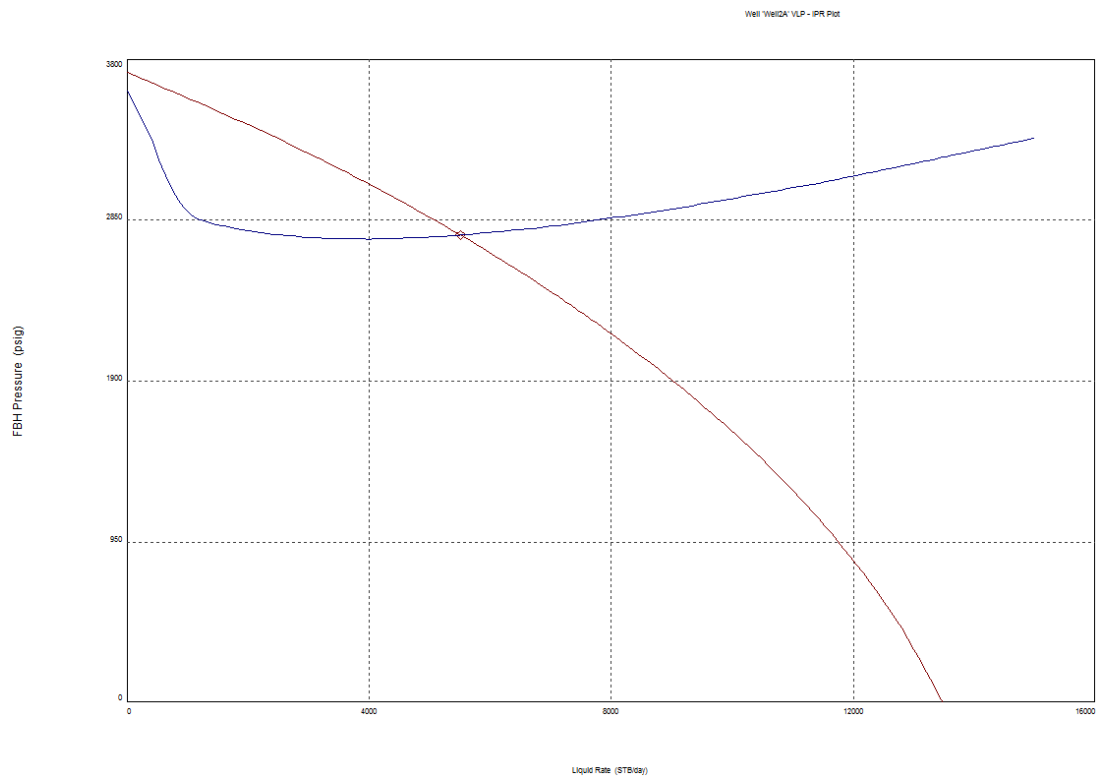Figure 5.13b IPR and VLP curve for Well1B
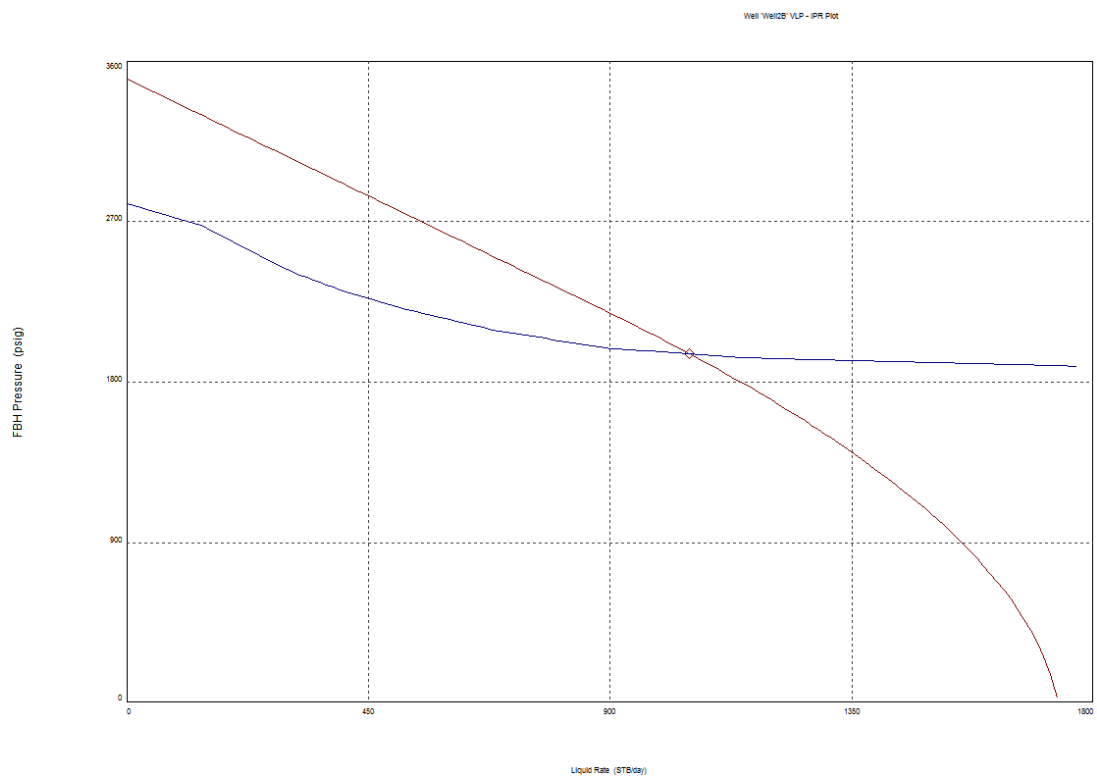
Figure 5.13c IPR and VLP curve for Well 2A

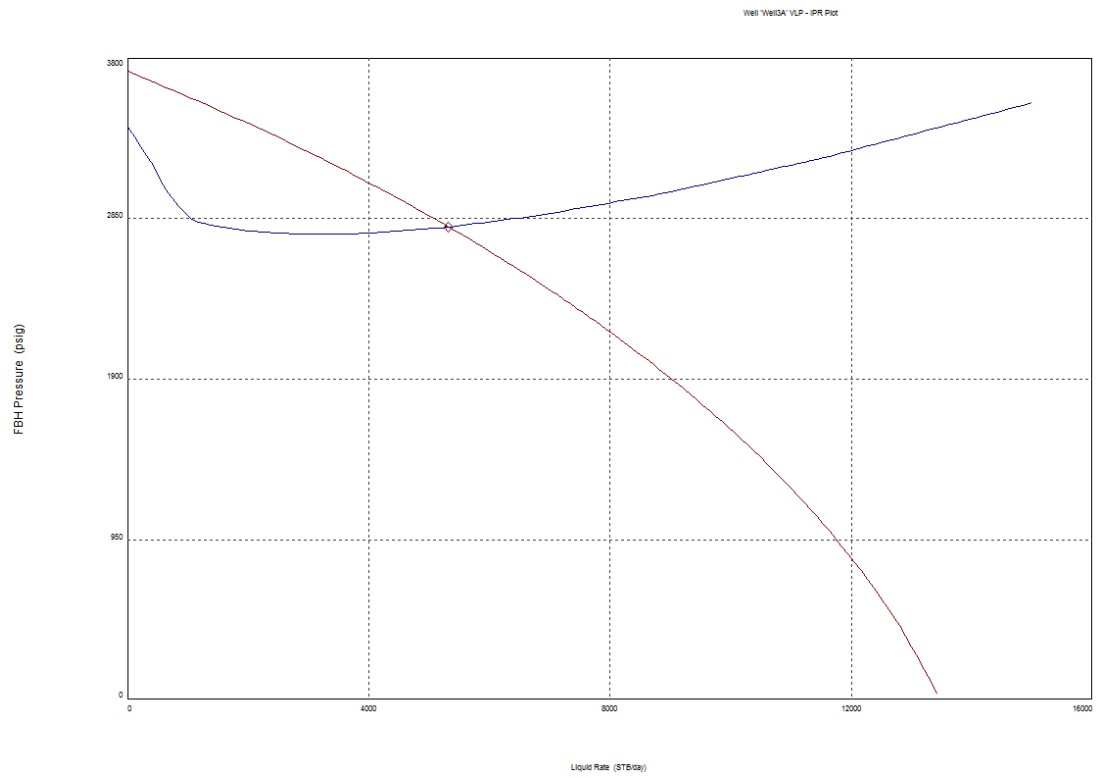Figure 5.13d IPR and VLP curve for Well 2B
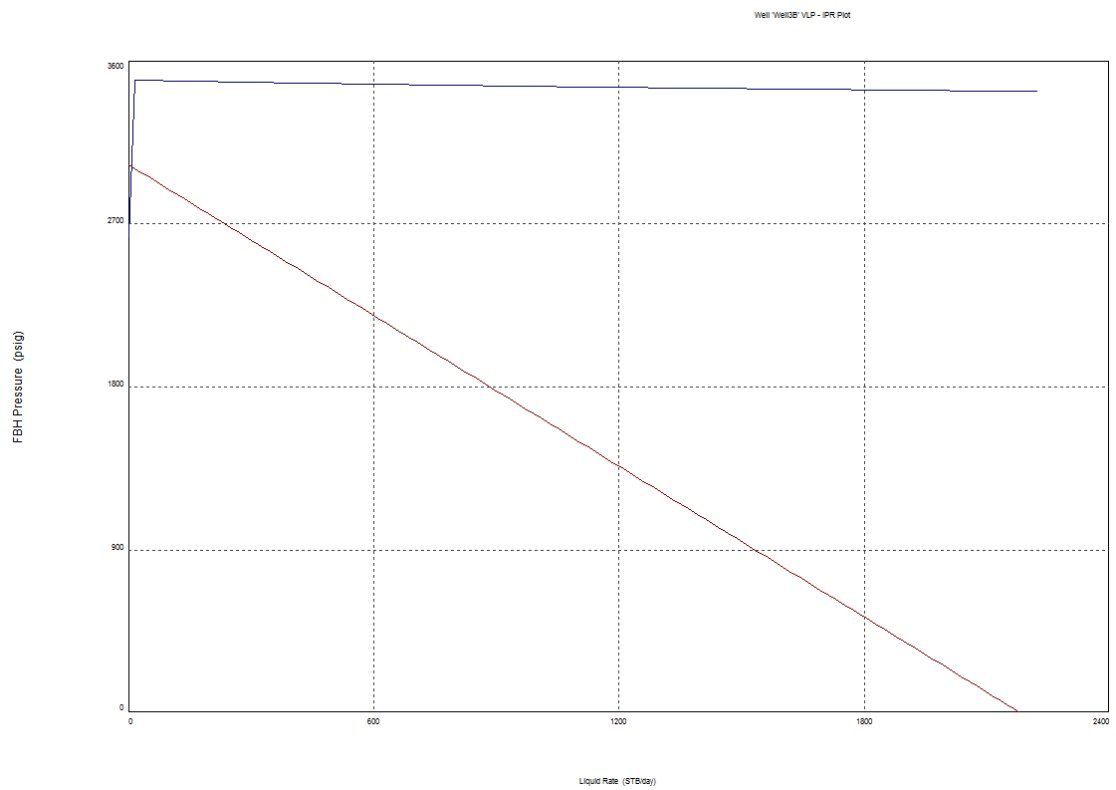
Figure 5.13e IPR and VLP curve for Well 3A

Figure 5.13f IPR and VLP curve for Well3B

Well Model Validation in GAP

This was done to ensure that, PROSPER well models that were imported into GAP, have been adequately matched with the field data, i.e. that the well models can reproduce the measured data from the field well tests. Well parameters such as reservoir pressure, manifold pressure, water cut, GOR and liquid rate were imported into GAP for the model validation. GAP was used then to generate IPR and VLP curves and calculate the bottom hole conditions for these test data.

The measured and the estimated values and the difference between the liquid rates are displayed in Table 5.14. From the results, it can be concluded that, there is an acceptable match with all the well tests.

Table 5.14 Comparison between measured and estimated rates for VLP/IPR wells

	Measured					Difference	Estimated				
Wells	Reservoir Pressure	Manifold Pressure	Liquid Rate	WCT	GOR	Liquid Rate	Liquid Rate	FBH Pressure	WCT	GOR	FWH Temperature
Well1A	3722	507	7021	9	800	25.490171	7046.49	2407.721	9	800	132.08606
Well1B	3500	205	1274	8	500	61.122599	1335.123	1571.232	8	500	71.398172
Well2A	3722	413	7850	8	800	-35.49277	7814.507	2213.572	8	800	138.13881
Well2B	3500	350	965	8	500	77.129744	1042.13	2019.517	8	500	64.827732
Well3A	3722	510	7790	9	800	8.9105659	7798.911	2219.815	9	800	137.55385
Well3B	3500	300	1170	8	500	98.207215	1268.207	1690.246	8	500	72.296797

The IPR and VLP curves generated by GAP for the test data are displayed on Figures 5.14 *a to f*. The IPR and VLP curves are represented by the red and blue lines respectively.
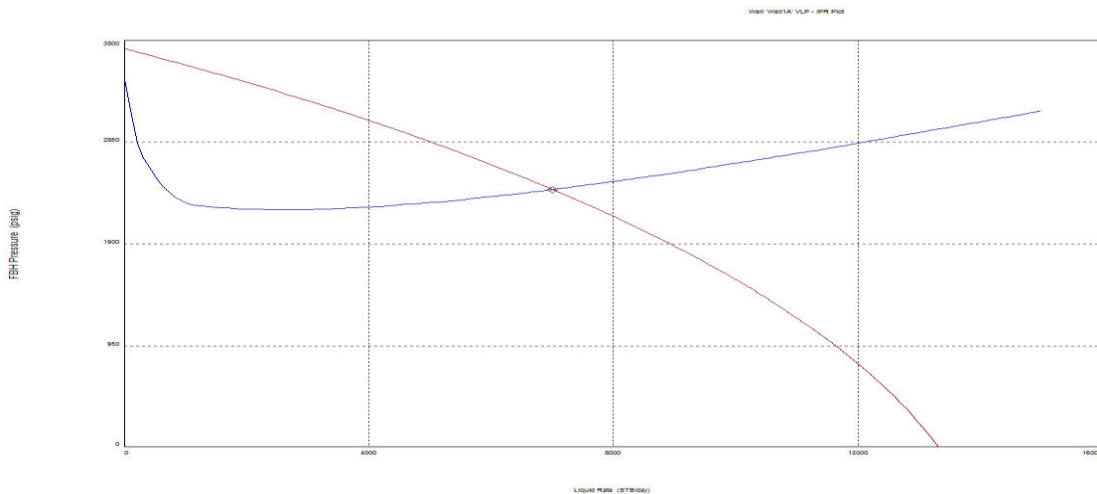

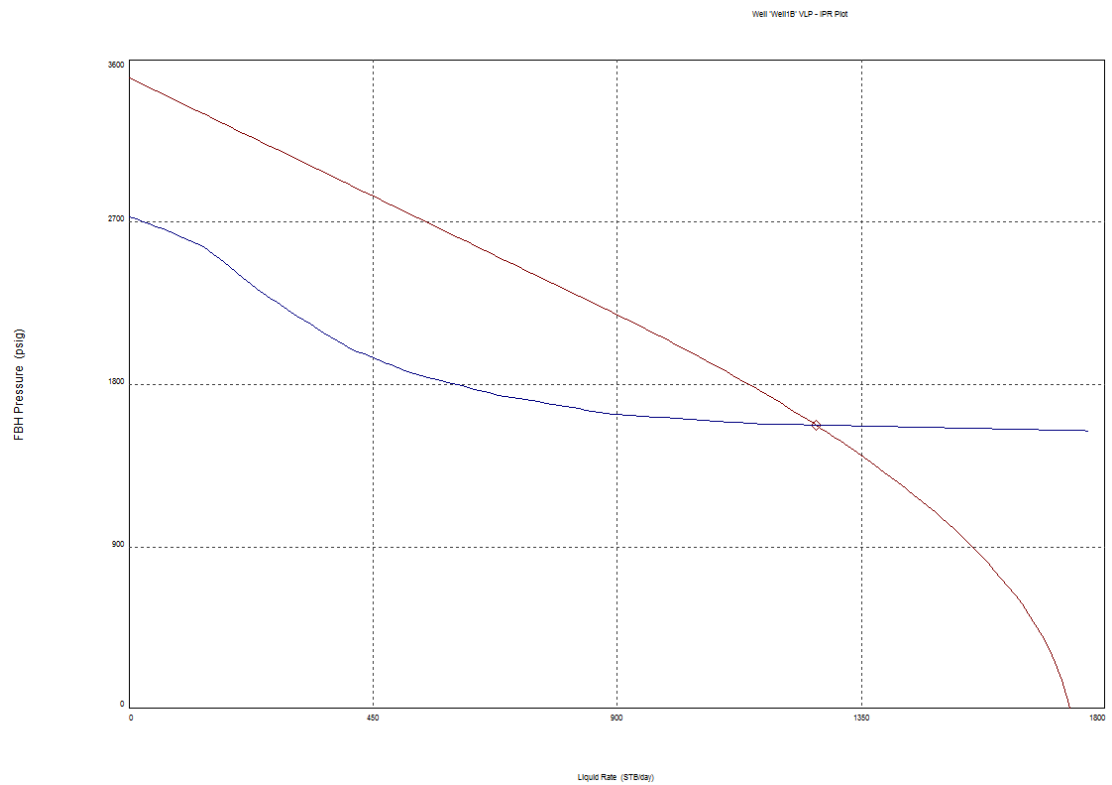Figure 5.14a IPR/VLP curve of test data for Well1A

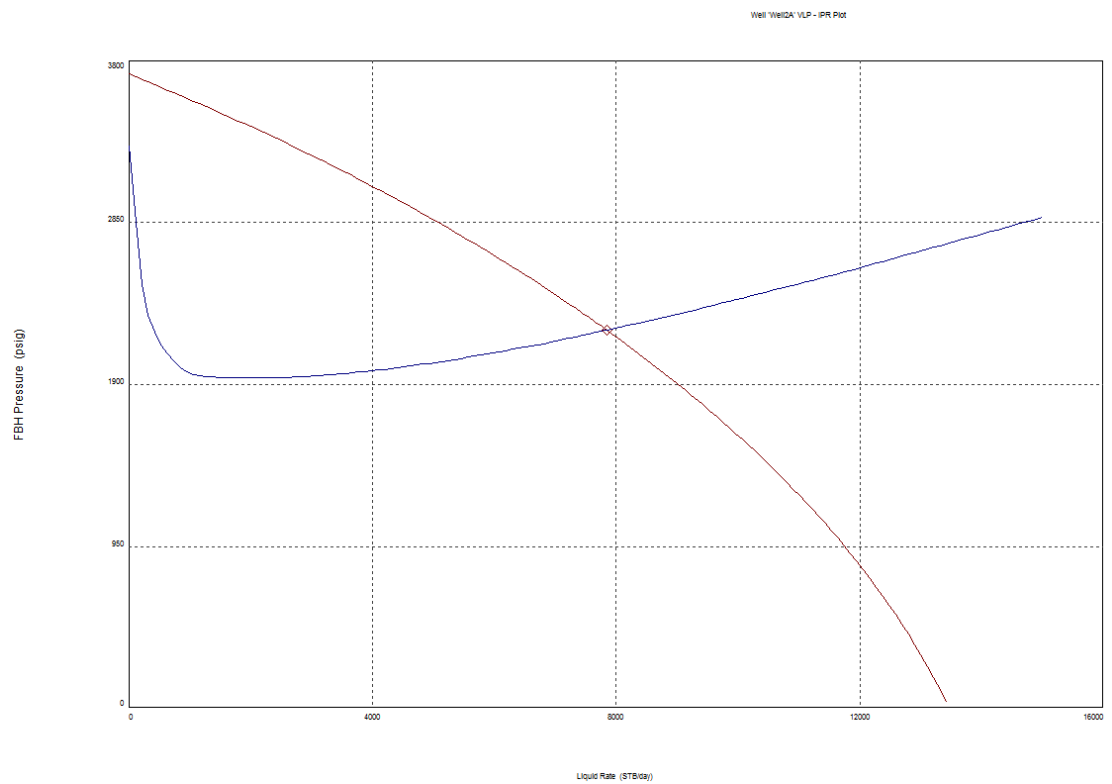Figure 5.14b IPR/VLP curve of test data for Well1B
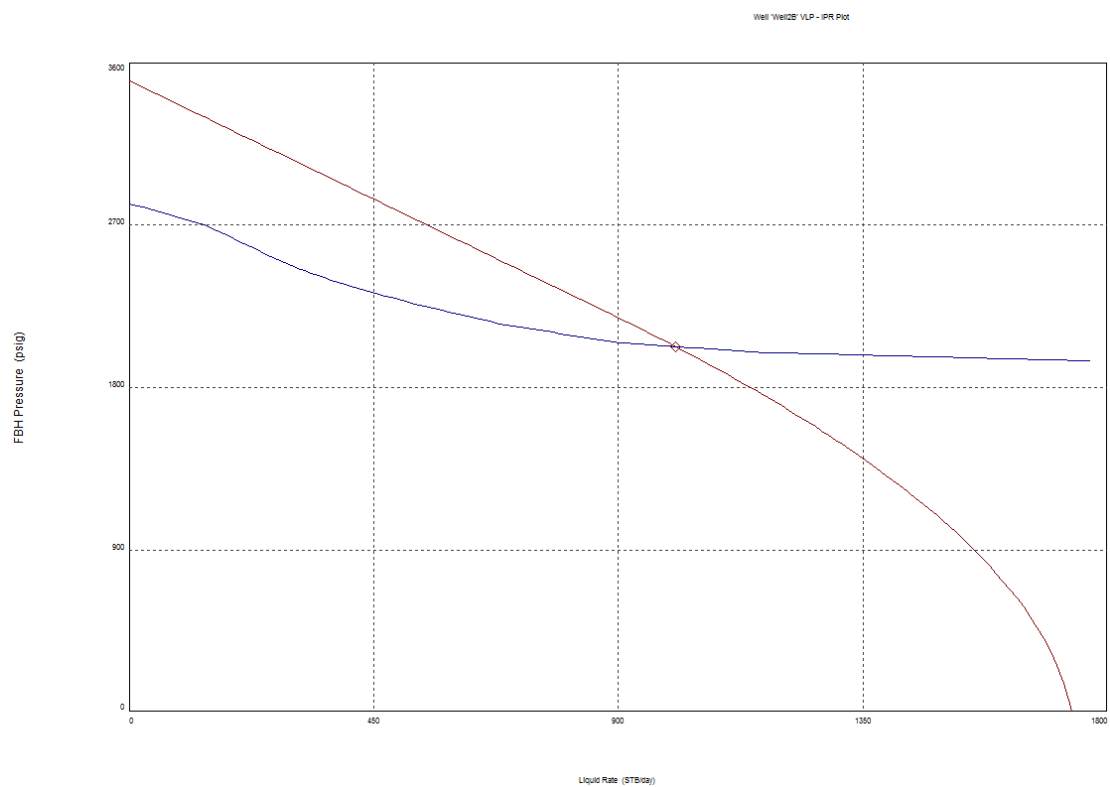
Figure 5.14c IPR/VLP curve of test data for Well2A

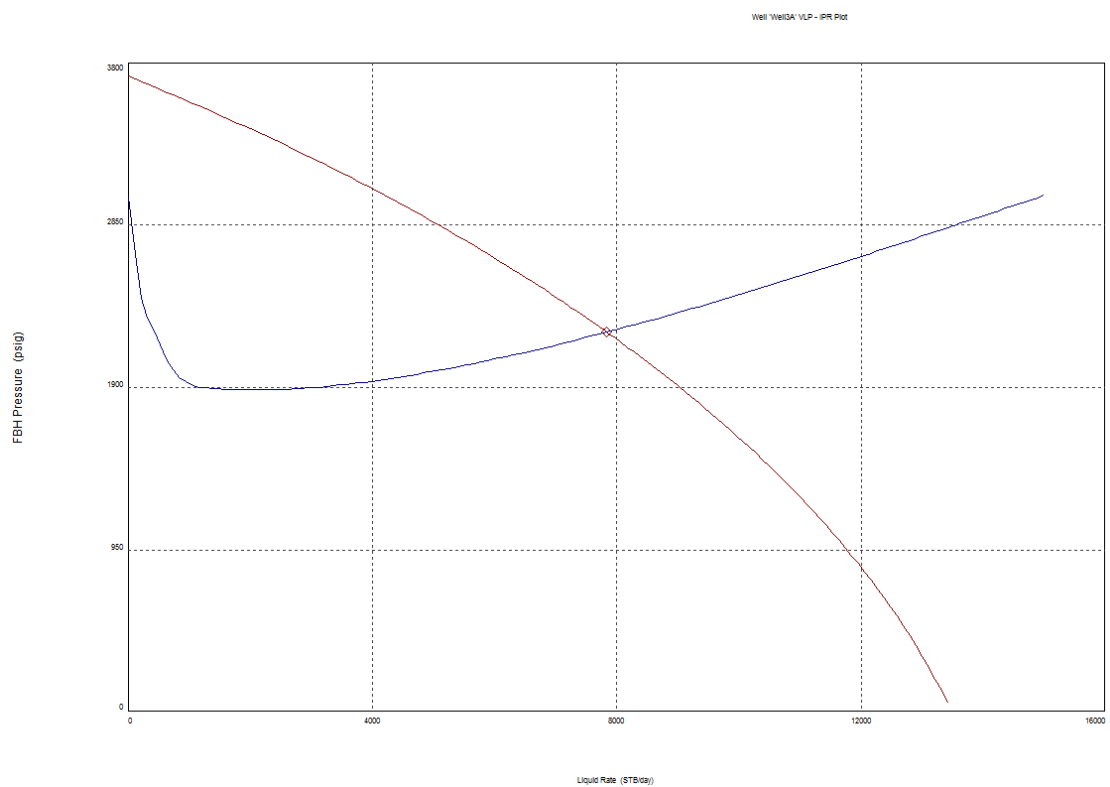Figure 5.14d IPR/VLP curve of test data for Well2B

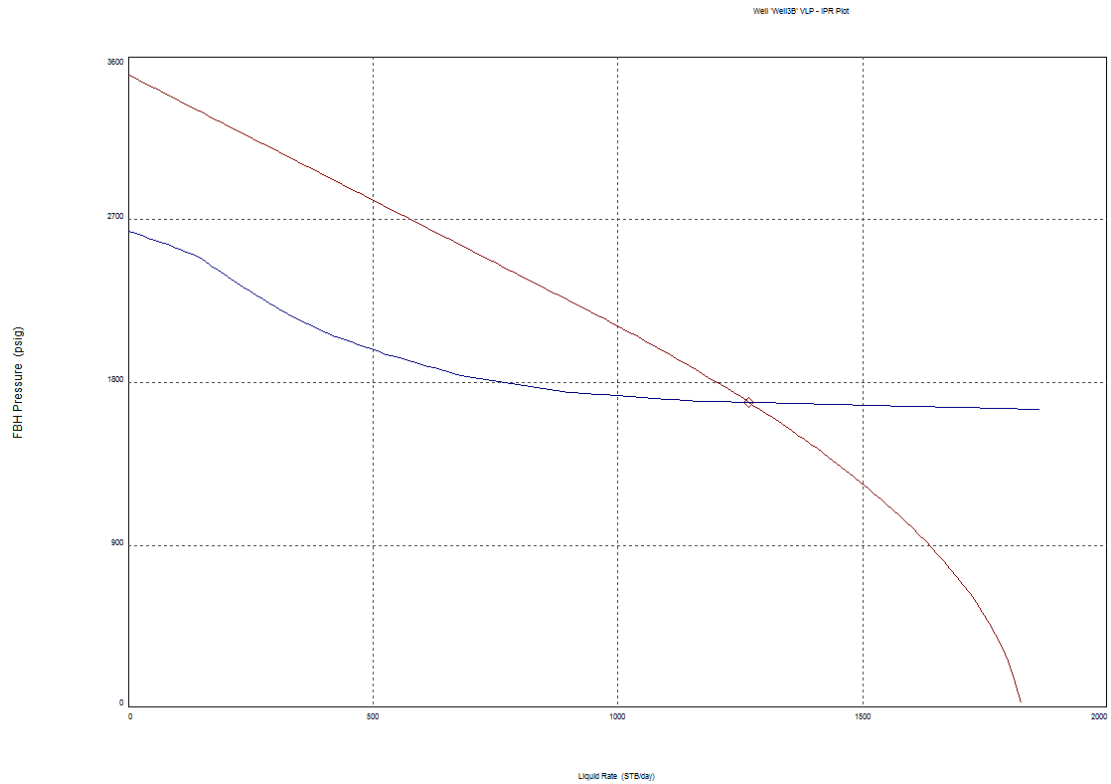Figure 5.14e IPR/VLP curve of test data for Well3A

Figure 5.14f IPR/VLP curve of test data for Well3B

5.1.4 PVT Data

Since the detailed data on the composition of the well effluents were not available, the Black Oil model was used for the estimation of the fluids' properties. Both reservoirs are oil reservoirs. The main characteristics of the reservoir oils are presented in Table 5.15.

Table 5.15 Characteristics of the Reservoir Oils

Properties	Reservoir A	Reservoir B
API (°)	37	39
GOR (scf/sbbl)	800	500
Average Reservoir Pressure (psig)	3750. 42	3542.19
Bubble point Pressure (psig)	3500	2200

5.1.5 Pipeline Description

In this project, we simulated two scenarios A and B using different flowline diameters (Table 5.16). Flowlines from wellheads to separators normally range between 2 inches to 20 inches in diameter [1]. Therefore, the selected diameters for both scenarios were within this range. In each scenario, the smaller diameters were

chosen for the flowlines connecting the wellheads to the manifolds while the larger diameters correspond to the flowline-connecting Manifold A to Manifold B, since it transports fluids from all the wells of reservoir A to Manifold B.
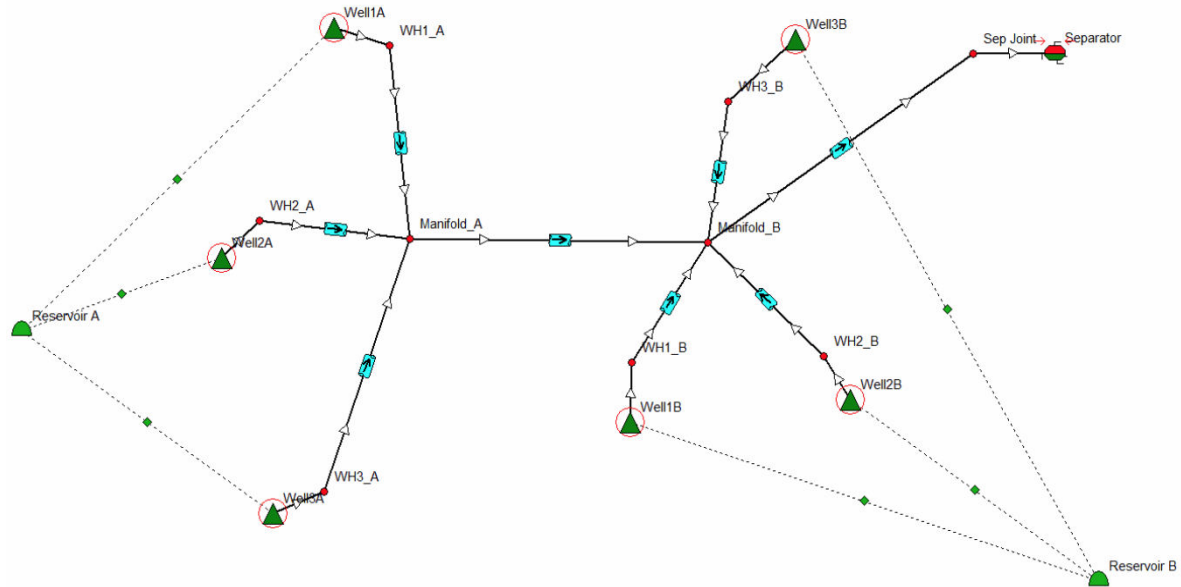

Figure 5.11 Surface network layout design of Loggie Mill offshore field with GAP

The flowline connecting Manifold B to the separator was always the largest in diameter because; it transports the comingled fluids from all six wells to the separator. Selected flowline diameters for both scenarios have 3 inches differences starting from the smallest to the largest. This is to observe if this 3 inches difference will effect any regular pattern on the production rates and pressure drop after simulation.

All flowlines have walls made of steel with an internal roughness of 0.0006 inches. The information about all the production flowlines, from the wellheads to the first separator are displayed in Table 5.16. The sea level is used as the datum level and the pipelines are defined from upstream to downstream in GAP.

The 12km long flowline, connecting manifold B with the separator joint (in yellow colour) has four different elevation profiles in Table 5.16. The downstream end of this flowline is the separator, which is located at the sea level with a reference depth of 0 feet while the upstream end is Manifold B, which is located at a depth of 525 feet. The first section of this flowline has an elevation from 525 ft to 480 ft (starting from the bottom of the table), the second from 480 ft to 515 ft, then from 515 ft to 525 ft, and finally from 525 ft to 0 ft which is the separator joint on the sea level. This elevation profile is graphically demonstrated on Figure 5.15 from GAP.

Table 5.16 Pipeline Data

| Pipe section	Pipe segment length	SCENARIOS		Upstream True Vertical Depth	Downstream True Vertical Depth	Overall Heat Transfer Coefficient
		Pipe ID				
		A	B			
	km	in	in	Feet	feet	Btu/h/ft2/F
WH1_A to Manifold_A	1	3	6	600	600	8
WH2_A to Manifold_A	0.8	3	6	600	600	8
WH3_A to Manifold_A	1.2	3	6	600	600	8
WH1_B to Manifold_B	0.5	3	6	525	525	8
WH2_B to Manifold_B	0.3	3	6	525	525	8
WH3_B to Manifold_B	0.4	3	6	525	525	8
Manifold_A to Manifold_B	7.5	6	9	600	525	8
Manifold_B to Sep Joint	3	9	12	525	0	8
Manifold_B to Sep Joint	4	9	12	515	525	8
Manifold_B to Sep Joint	3	9	12	480	515	8
Manifold_B to Sep Joint	2	9	12	525	480	8
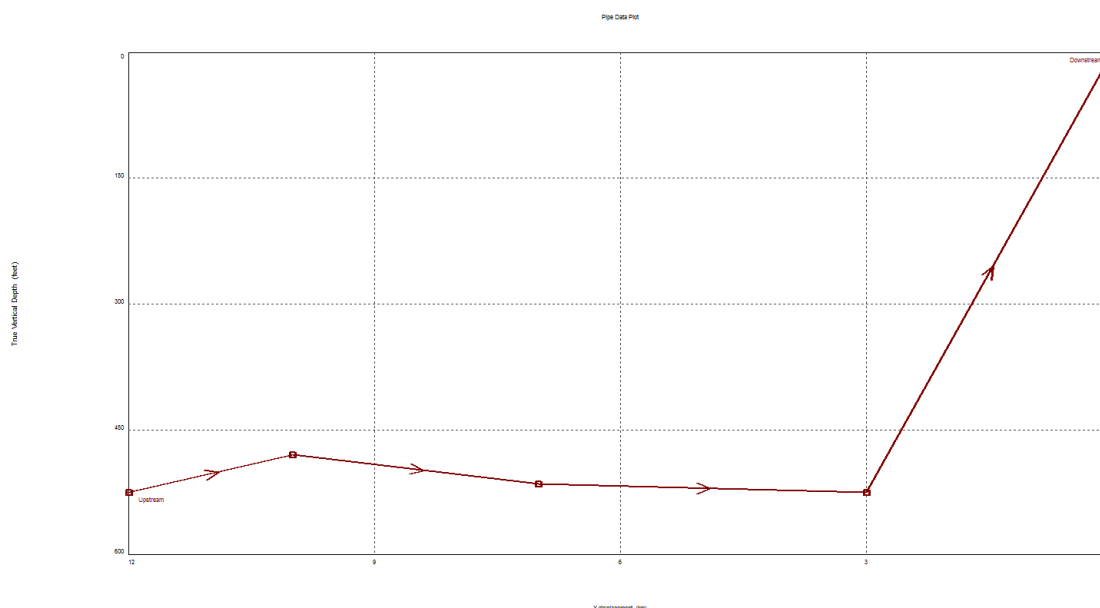
Figure 5.15 Pipeline elevation profile from Manifold_B to Separator joint
(generated from GAP).

Pipeline Matching

Field measurements for the flowline from Manifold_A to Manifold_B are presented in Table 5.17. The data were used to match the pipeline in GAP in order for GAP to reproduce the same pressure drop as it is in the field. Unfortunately, the date at which this data were measured was not available.

Table 5.17 Field Data for Pipeline Matching

Upstream Pressure	Upstream Temperature	Liquid Rate	Downstream pressure	Water Cut	Gas Oil Ratio	Oil Gravity	Gas Gravity	Water Salinity
psig	psig	stb/d	psig	%	scf/stb	^{o}API		ppm
480	90	17250	340	47	800	37	0.76	23000

Pipeline Environment Description

The environmental conditions of the subsea area, where the pipelines are located are presented on Table 5.18.

Table 5.18 Pipeline Environment

Surface Temperature	50 ^{o}F
Overall Heat Transfer Coefficient	8 BTU/h/ft
Oil Heat Capacity	0.53 BTU/lb/f (GAP default)
Gas Heat Capacity	0.51 BTU/lb/f (GAP default)
Water Heat Capacity	1 BTU/lb/f (GAP default)

The environmental information was also input into GAP. The flow in the pipelines is considered as multiphase and the pressure drop is calculated using the Petroleum Experts 5 correlation as it is recommended in the GAP manual.

5.2 Simulation

After the well data from PROSPER, reservoir data from MBAL, PVT and pipeline data were imported into GAP, the 2 scenarios were simulated for a period of 10 years from 01/12/2010 to 01/12/2020. GAP was set so as to choke the wells in case the separator constraint of 20,000 stb/day was exceeded during the simulation. Analysis of the simulation results is presented below.

5.3 Results

5.3.1 Cumulative oil production

At the end of the 10-year simulation period, scenario A yields a cumulative oil production of 27 MMstb and for scenario B, a cumulative oil production of 29.2 MMstb. The result is shown on figure 5.16

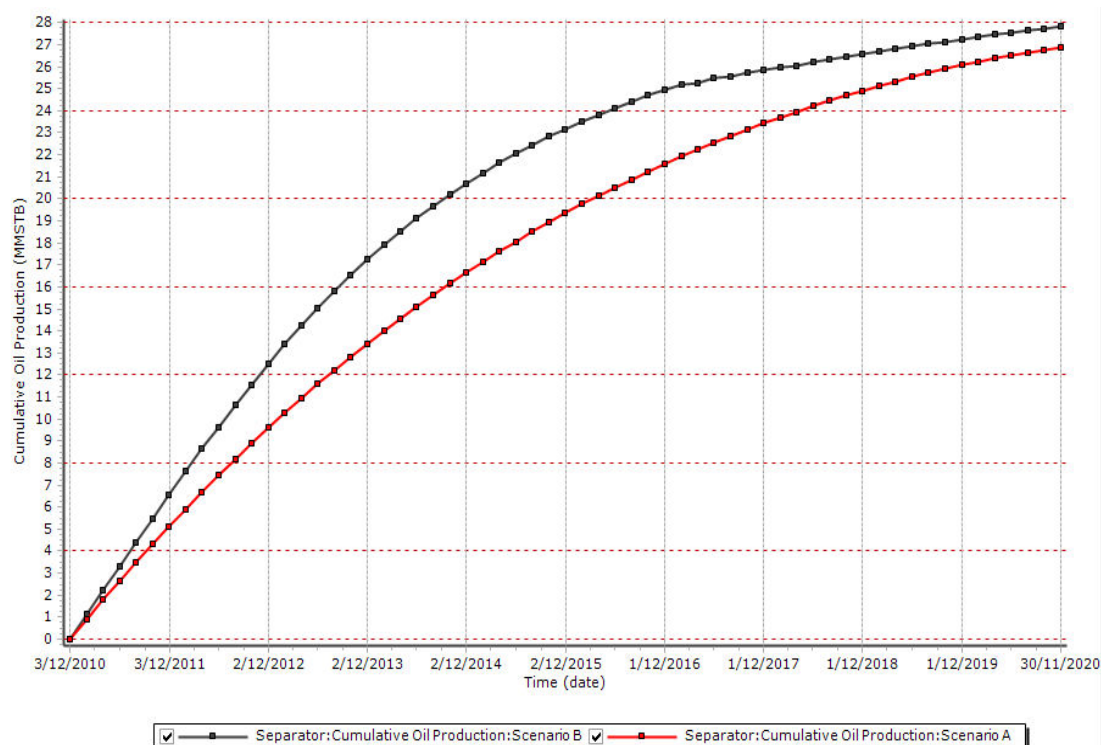Figure 5.16 Cumulative oil production of reservoir A and B from separator results.

5.3.2 Well performance

A graph of oil rate vs pressure is plotted for each well and for each scenario. The results are shown on Figures 5.17 *a* and *b*.
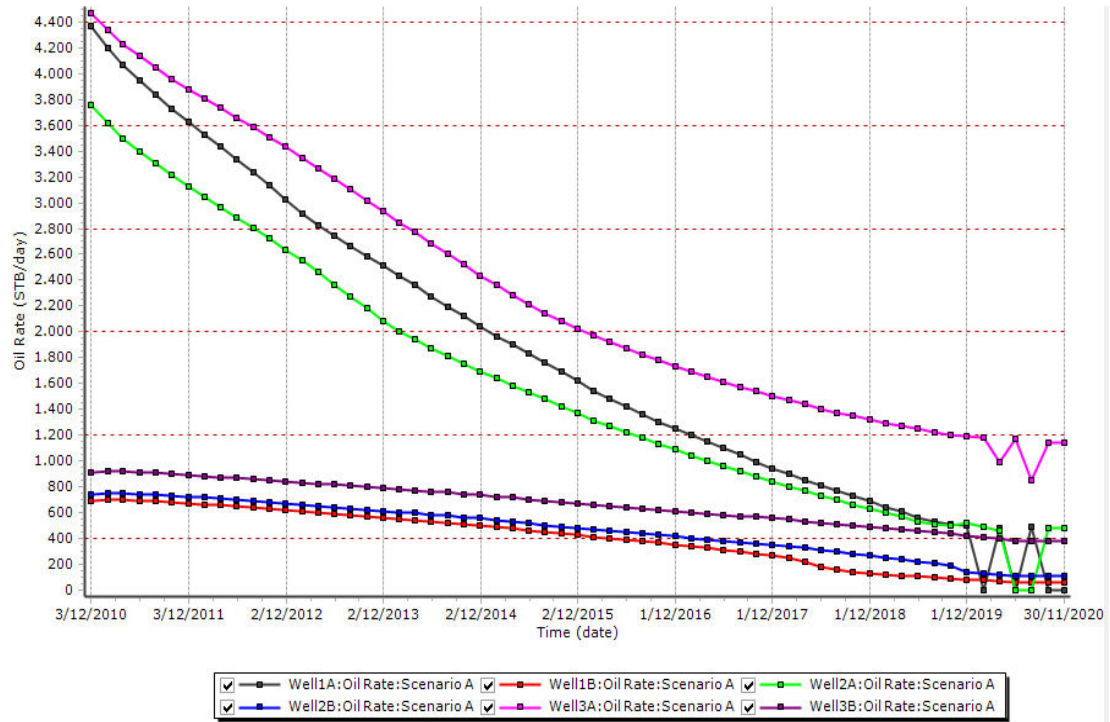

Figure 5.17a Well performance for scenario A
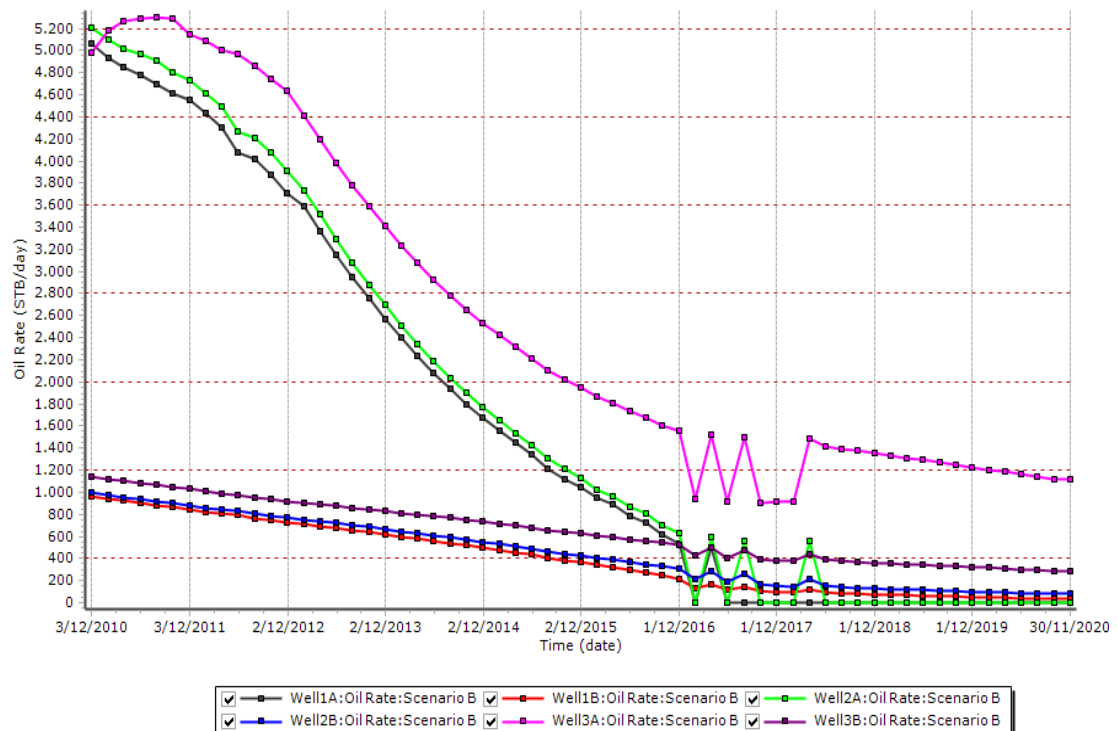
Figure 5.17b Well performance for scenario B

From Figure 5.17a, it can be seen that Well1A and Well2A die at the beginning of year 2020 for scenario A. Figure 5.17b also indicates that Well1A and Well2A die after 6 six years of production for scenario B. The well conditions at the date at which Well1A begins to die are analysed for scenario A in the following graphs.

Figure 5.18 Well performance analysis for Well1A (scenario A)

The date at which Well1A begins to die was estimated from Figure 5.18 to be 01/09/2019. The IPR and VLP curves at this date are plotted in figure 5.19.
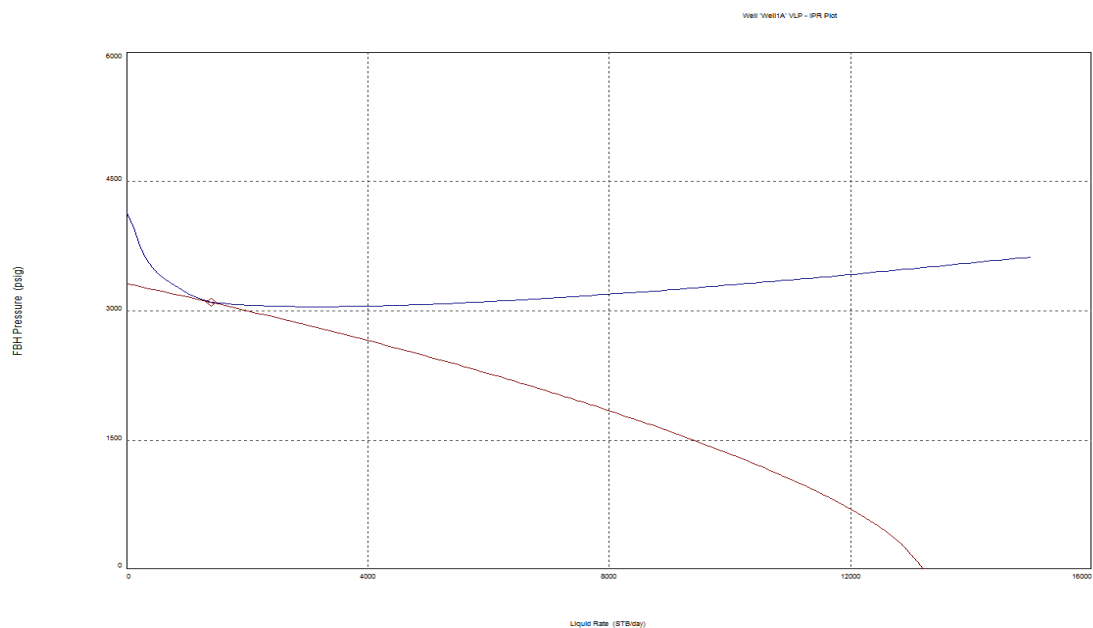
Figure 5.19 IPR and VLP curve for Well1A (scenario A) on 01/09/2019

From Figure 5.19, it is inferred that, the well is about to die, as the operating bottomhole pressure is the minimum required to lift the fluid to the surface.

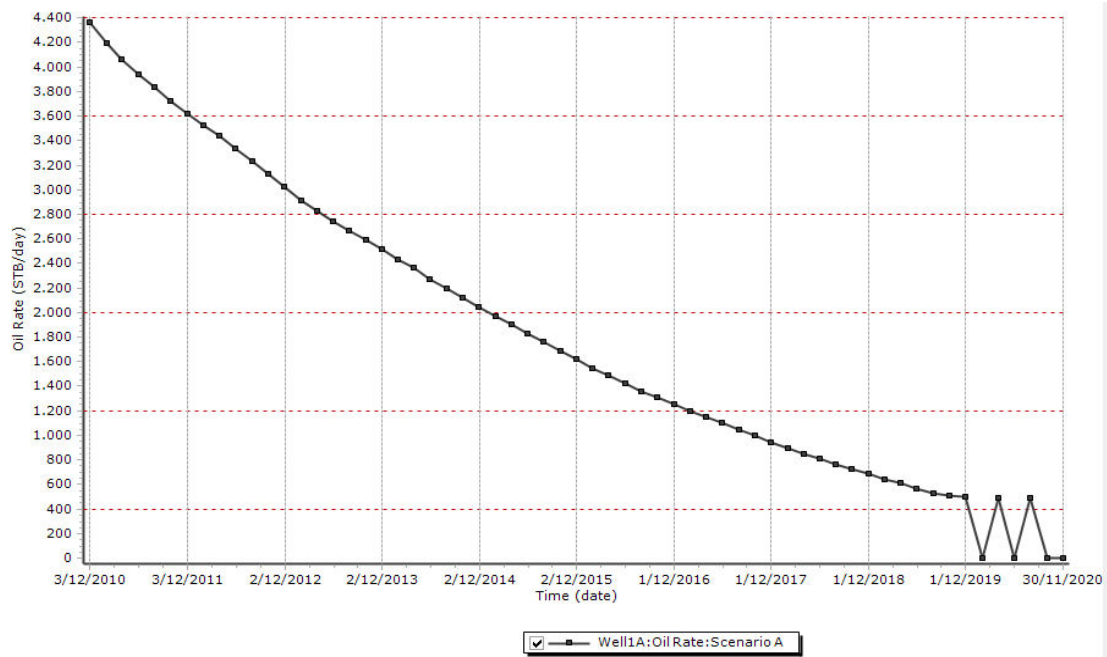To further demonstrate the sensitivity of Well1A, the manifold pressure on 01/09/2019 was increased from 424 psig to 450 psig and the IPR/VLP curves were plotted again as shown in Figure 5.20.
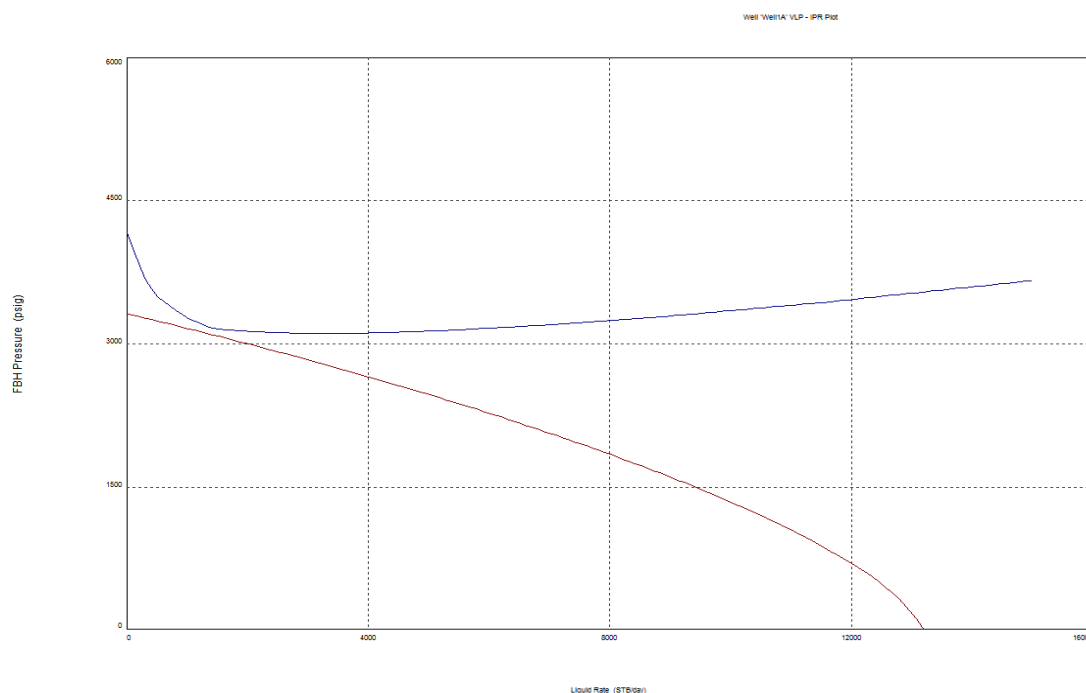
Figure 5.20 IPR/VLP curve for Well1A on 01/09/2019 after Manifold pressure was increased to 450 psig

It can be observed from the IPR/VLP curve in Figure 5.20 that the well has stopped flowing.

5.3.3 Reservoir Performance

From figures 5.21a and 5.21b, it can be seen that, reservoir A falls below the bubble point pressure of 3500 psig in both scenarios, whilst reservoir B undergoes a steady decline in pressure but it remains above the bubble point pressure of 2200 psig even after the 10 year simulation period.
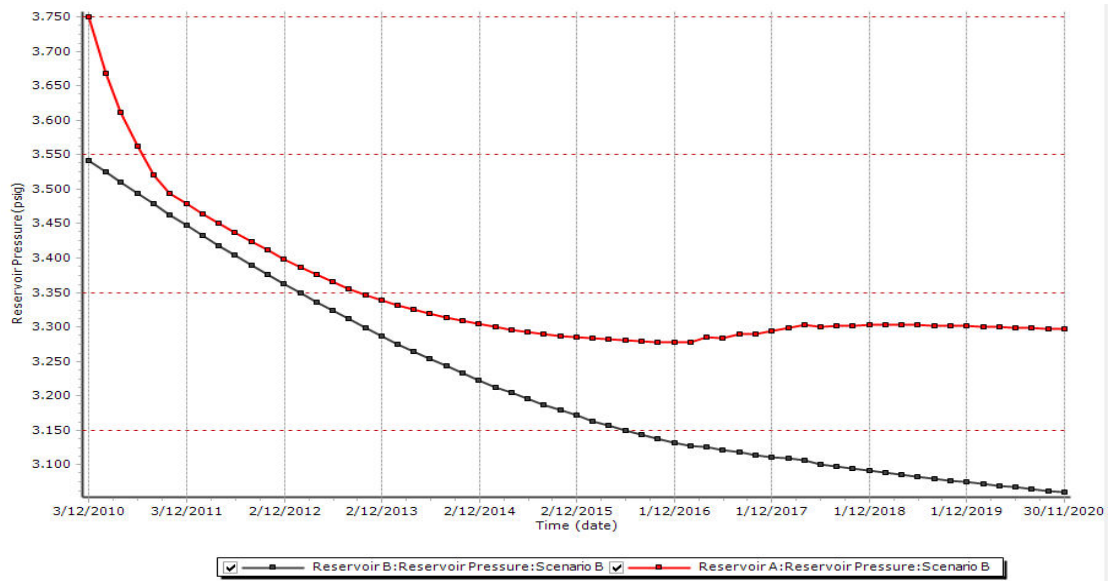
Figure 5.21a Reservoir pressure decline for scenario A
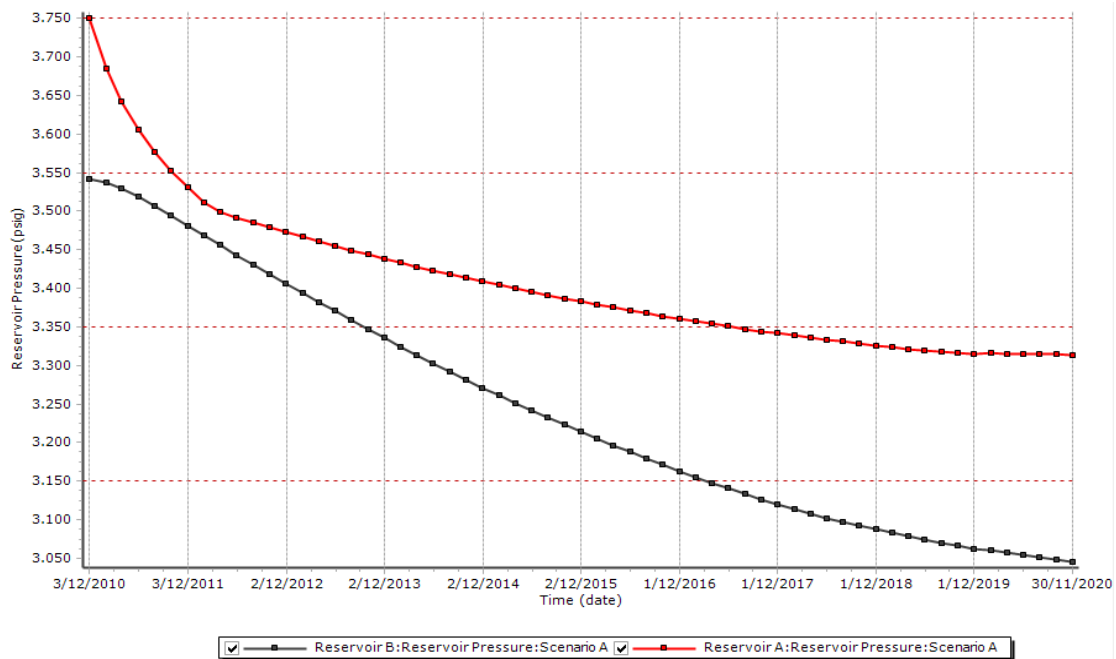
Figure 5.21b Reservoir pressure decline for scenario B

5.3.4 Reservoir Pressure Support

Since reservoir A fell below the bubble point pressure in both scenarios, GAP was instructed to start water injection into reservoir A to maintain the pressure at a target of 3670 psig above its bubble point pressure of 3500 psig. The effect of water injection into reservoir A and the respective producing wells are shown in Figures 5.22a and 5.22b.
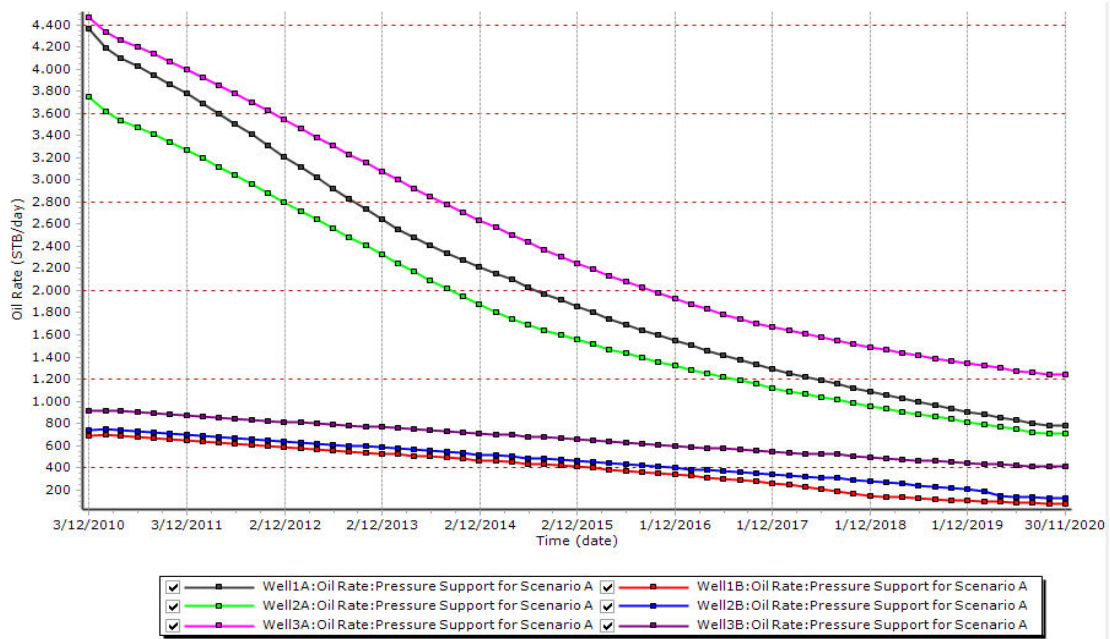
Figure 5.22a Pressure support for reservoir A (scenario A)
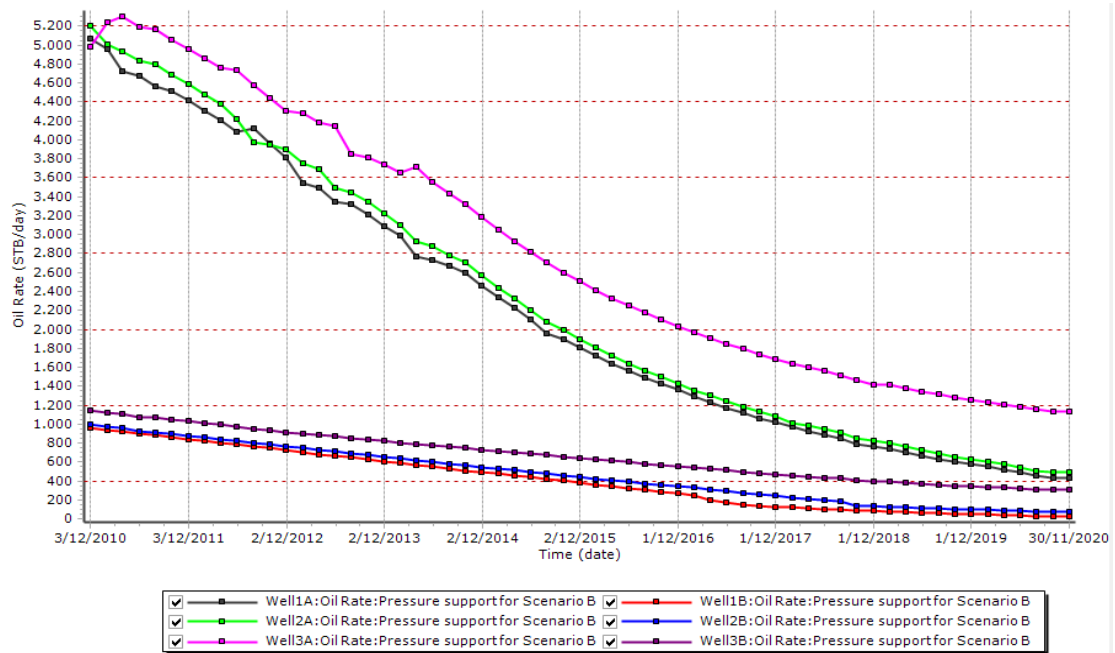
Figure 5.22b Pressure support for reservoir A (scenario B)

The additional pressure support for reservoir A prevents Well1A and Well1B from dying at the time period of the simulation and resulted in a significant increase in the cumulative oil production from 27 MMstb to 29.2 MMstb for scenario A and from 28 MMstb to 33MMstb for scenario B. This is shown on Figures 5.23 *a* and *b*.
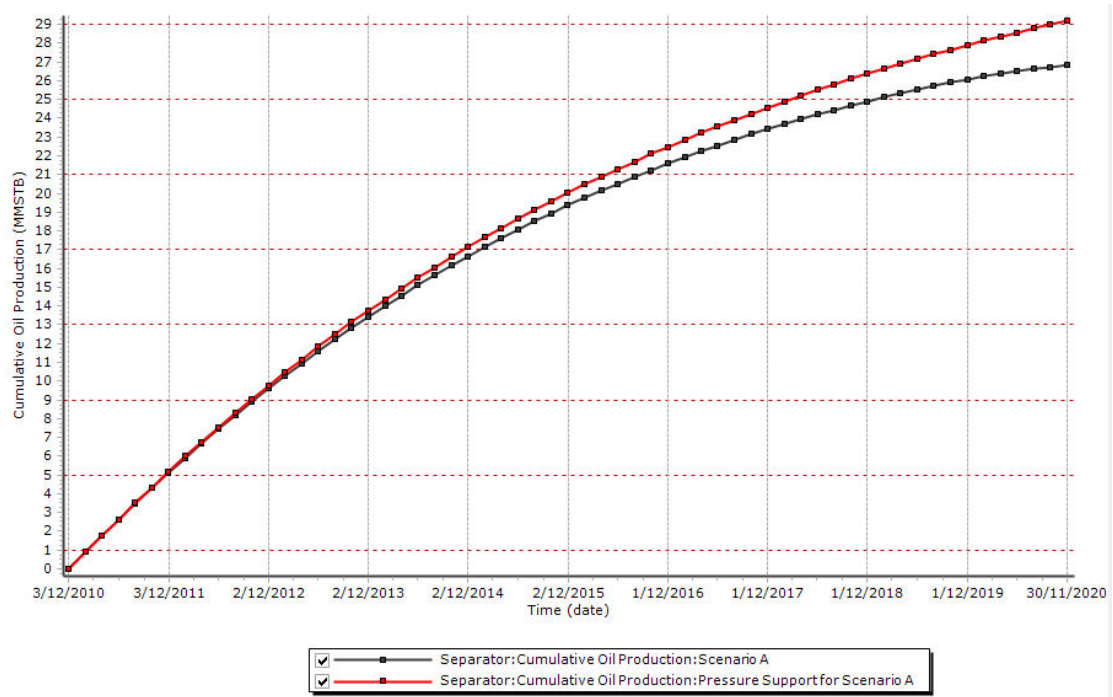
Figure 5.23a Pressure support for reservoir A (scenario A)
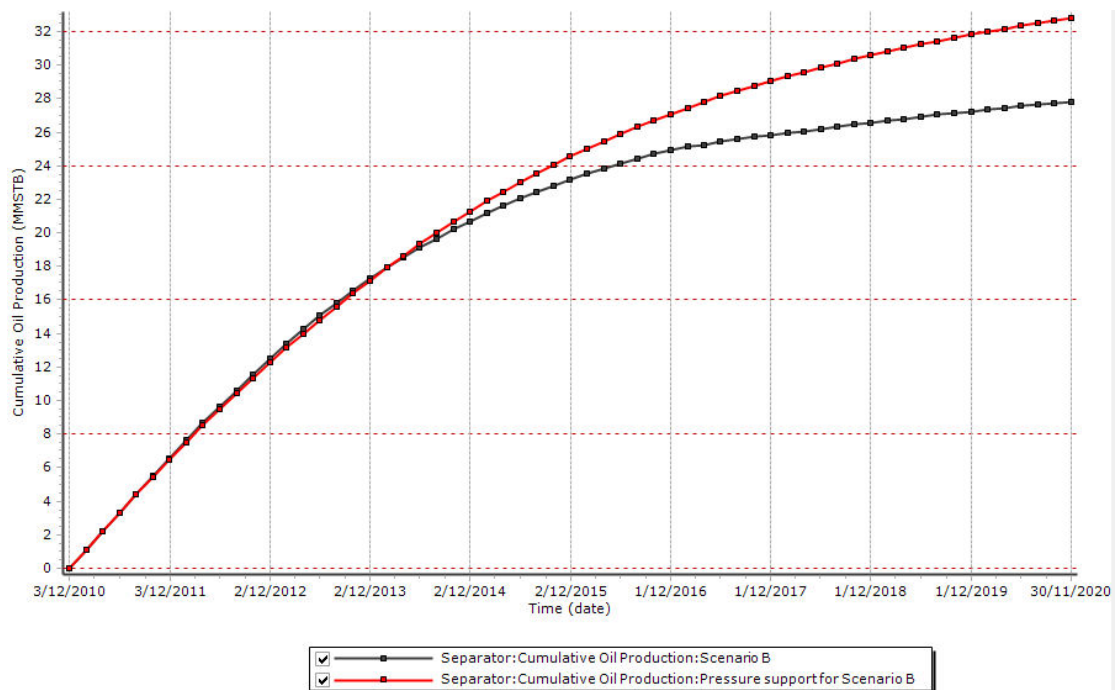
Figure 5.23b Pressure support for reservoir A (scenario B)

Figures 5.24 *a* and *b* show the average water quantity that was injected into reservoir A in order to maintain the pressure at the predefined value of 3670 psig.
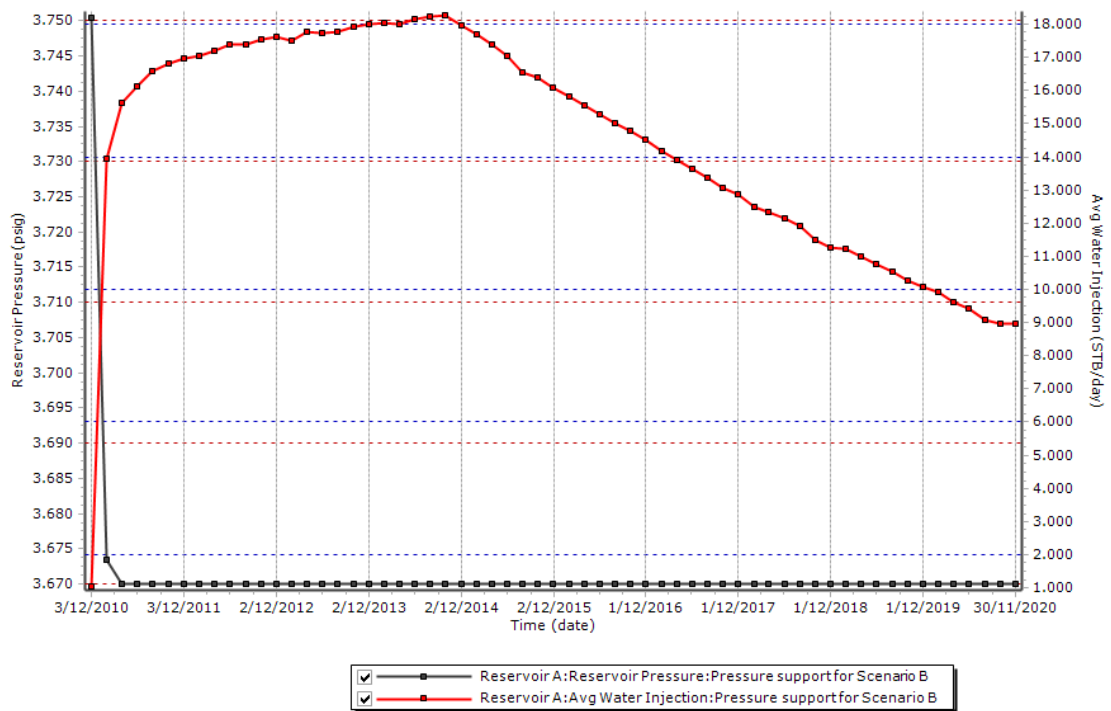
Figure 5.24a Reservoir A water injection and pressure maintenance (scenario A)
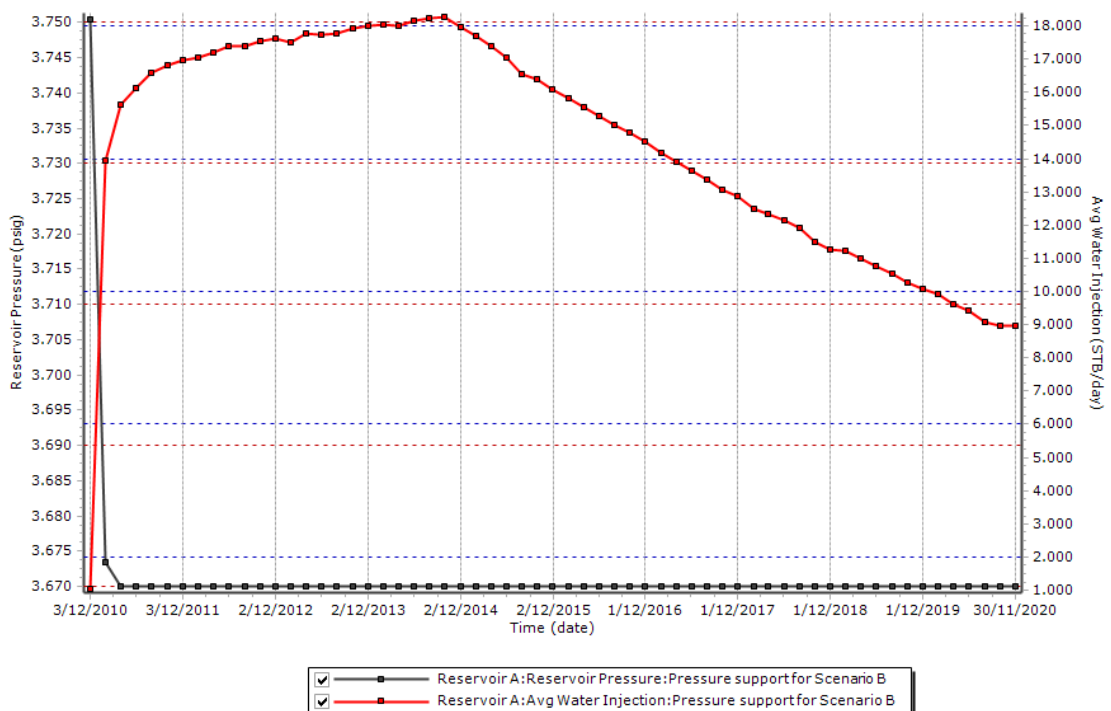
Figure 5.24b Reservoir A water injection and pressure maintenance (scenario B)

Figure 5.24 can be used to screen various water injection schemes for the Loggie Mill Block's offshore field. After the final injection scheme is chosen, and by following the rates of water injected per day by GAP into reservoir A, the engineer

can also inject the same amount of water per day into reservoir A to maintain its pressure at the predetermined value above its bubble point.

5.3.5 Well Results

The well operating conditions as predicted in GAP for both scenarios are presented in the Tables 5.19 *a* and *b*.

Table 5.19a Well Results for Scenario A

Wells	Reservoir Pressure	Bottomhole Pressure	Wellhead Pressure	Wellhead Temperature	Average Liquid Rate	Average Water Cut	Drawdown	dp Choke
	psig	psig	psig	^{o}F	stb/day	%	psig	psig
Well1A	3750.43	2793.19	748.42	119.15	5484.8	8.28	957.24	220.92
Well2A	3750.43	2765.22	751.39	122.07	5427.5	8.28	985.18	225.23
Well3A	3750.43	2805.03	950.9	119.37	5615.9	8.28	945.4	421
Well1B	3542.19	1993.5	331.29	66.15	1121.4	8.01	1548.69	0
Well2B	3542.19	1943.02	331.22	66.72	1110.4	8.01	1599.17	0
Well3B	3542.19	1789.1	331.28	71.59	1240.1	8.01	1753.09	0

From Table 5.19a, it can be seen that, producing wells of reservoir A are choked back more heavily. This is because; they produce at high flow rates and therefore present high associated water cut. As a result, they were choked back to reduce the water cut to obey the separator capacity constraint of 20,000 stb/d.

On the other hand, the wells producing from reservoir B have lower flow rates with respect to the ones of reservoir A. This might be associated with their low PI (which is less than 1) and the absence of sand control at the perforations. Their low PI values and the absence of sand control can be verified in the *well description* section.

Table 5.19b Well Results for Scenario B

Wells	Reservoir Pressure	Bottomhole Pressure	Wellhead Pressure	Wellhead Temperature	Average Liquid Rate	Average Water Cut	Drawdown	dp Choke
	psig	psig	psig	^{o}F	stb/day	%	psig	psig
Well1A	4000.1	2432.82	512.22	135.78	8289.8	0	1567.19	0
Well2A	3722.01	2353.27	502.03	135.03	7309.8	8	1368.74	0
Well3A	3722.01	2984.18	1064.66	108.97	4407.5	9	737.83	568.83
Well1B	3500	3500	333.04	60	0	_	0	200000
Well2B	3500	3500	333.05	60	0	_	0	200000
Well3B	3026	2973.79	333.08	41	0	_	52.21	200000

For scenario B, only wells draining reservoir A contribute to production and Well3A is the only one amongst them that was choked back by GAP. Well1B and Well2B do not produce because, there is no pressure drawdown to cause flow from the reservoir into the wellbore since the reservoir and the bottomhole pressures are the same (3500 psig). Well3B does not produce because; the drawdown pressure is too small to cause any significant production and has a high choke backpressure.

5.3.6 Pipeline Results

The GAP predictions of the pipeline for both scenarios after the 10 year period are presented in Tables 5.20 *a* and *b*. It can be seen from Table 5.20b that, there is no production from pipelines connected to the wells draining from reservoir B for scenario B and the reason is already explained above. Table 5.20b also indicates that, there is negligible to no pressure drop in these pipelines and therefore there is no flow.

Figure 5.20a Pipeline results for Scenario A

Pipe segment	Average oil rate	Upstr. Press.	Downstr. Press.	Upstr. Temp.	Downstr. Temp.	ΔP	dP Accel.	dP Frict.	dP Grav.	Max. mix. vel.	Water Cut
	stb/d	psi	psi	°F	°F	psi	psi	psi	psi	ft/s	%
WH1_A to Manifold_A	5484.8	527.5	519.17	119.15	77.34	8.34	0.00	8.34	0.00	6.844	8.28
WH2_A to Manifold_A	5427.5	526.15	519.17	122.07	84.9	6.99	0.00	0.00	0.00	7.102	8.28
WH3_A to Manifold_A	5615.9	528.95	519.17	119.37	72.51	9.78	0.00	9.78	0.00	6.773	8.28
WH1_B to Manifold_B	1121.4	331.29	331.1	66.15	51.48	0.18	0.00	0.18	0.00	1.245	8.01
WH2_B to Manifold_B	1110.4	331.22	331.1	66.72	54.04	0.11	0.00	0.11	0.00	1.263	8.01
WH3_B to Manifold_B	1240.1	331.28	331.1	71.59	53.89	0.18	0.00	0.18	0.00	1.404	8.01
Man._A to Manifold_B	16528.2	519.17	331.1	78.25	52.59	188.09	0.00	157.62	30.47	14.93	8.23
Manifold_B to Sep Joint	20000.1	331.1	200	52.77	50	131.1	0.00	70.25	60.75	12.143	8.23

Scenario B produces more oil than scenario A due to larger pipe diameters.

The pressure drop in the pipelines is not consistent with the 3 inches difference in diameter and therefore does not follow any regular pattern. For example, the diameter of the flowline "*WH1_A to Maniflod_A*" for scenario A is 3 inches and the same flowline for scenario B has a 6 inch diameter. It is expected that, for the

same flow rate, the pressure drop will be higher in scenario A than scenario B. From the results, there is a pressure drop of 22.94 psia for scenario B and 8.34 psia for scenario A, even though; scenario B has a larger diameter. This is because scenario B has a higher liquid flowrate of 8289.8 stb/day and a mixture velocity of up to 11.974 ft/s, whilst scenario A has a liquid flowrate of only 5030.7 stb/day and a velocity up to only 6.844 ft/s. Therefore, it can be concluded that, the higher flowrate and higher mixture velocity caused a higher pressure drop in scenario B than A for WellA, even though scenario B has a larger diameter.

Figure 5.20b Pipeline results for Scenario B

Label	Average oil rate	Upstr. Press.	Downstr. Press.	Upstr. Press.	Downstr. Press.	ΔP	dP Accel.	dP Frict.	dP Grav.	Max. mix. vel.	Water Cut
	stb/d	psi	psi	^oF	^oF	psi	psi	psi	psi	ft/s	%
WH1_A to Manifold_A	8289.8	512.22	489.29	135.78	94.16	22.94	0	22.94	0	11.974	0
WH2_A to Manifold_A	6725	502.03	489.28	135.03	98.65	12.75	0	12.75	0	10.048	8
WH3_A to Manifold_A	4010.8	496.03	489.29	108.97	64.88	6.75	0	6.75	0	5.646	9
WH1_B to Manifold_B	0	333.04	333.08	60	60	-0.04	_	_	_	0	_
WH2_B to Manifold_B	0	333.05	333.08	60	60	-0.03	_	_	_	0	_
WH2_B to Manifold_B	0	333.08	333.08	60	50	0	_	_	_	0	_
Man._A to Man._B	20007.1	429.86	333.08	89.34	52.05	156.2	0	145.42	10.77	14.68	4.91
Manifold_B to Sep Joint	20007.1	333.08	2000	52.05	50	133.08	0	68.33	64.75	12.141	4.91

In addition, the temperature generally decreases from upstream to downstream for all pipelines. This is because; the effluents of the wells have higher temperature than the surrounding pipeline environment temperature of 50 ^oF. Therefore, heat is lost to the environment by convection through contact with the walls of the pipe as the fluids are transported through the pipe.

5.3.7 Separator Results

The separator was set to a pressure of 200 psia and a temperature of 50 ^{o}F in GAP. It was designed to support 20,000 stb of liquid per day. The predicted results of the separator after the simulation for both scenarios are presented in Table 5.21.

Table 5.21 Separator

Separator	Average Liquid rate	Average Oil Rate	Average Gas Rate	Separated Gas Rate	Average Water Rate	Separated Water Rate	Water Cut	GOR	Number of active wells
	stb/d	stb/d	MMscf/d	MMscf/d	stb/d	stb/d	%	Scf/std	
Scenario A	20000.1	18353.3	13.724	12.524	1646.7	1646.7	8.23	747.79	6
Scenario B	20007.1	19025.6	15.221	13.998	981.5	981.5	4.91	800	3

It can be verified from Table 5.21 that, scenario B yields a higher oil production per day with almost half the water cut of scenario A. Both scenarios comply with the separator capacity constraint of 20,000 stb/day.

It is also clear that, the separator did not separate at least about 8.74% and 8.01% of the gas produced for scenario A and B respectively. For example, the average gas rate for scenario A is 13.724 MMscf/d and the average gas separated is 12.524 MMscf/d. Therefore, there might be a requirement for a second separator in order to separate the gas carry over. All produced water is separated.

The number of active wells producing to the separator at the end of the 10-year period is consistent with the previous analysis of the well production rates.

5.3.8 Flow Assurance Results

Below is the analysis of the flow assurance results of the flowline from Manifold B to the separator that was predicted by GAP.

Predicted Flow Regime

GAP was used to generate a Taitel-Duckler flow regime map for the flowline connecting Manifold B to the separator on 01/12/2010 (the simulation start date). The Taitel-Duckler flow map is used to visually inspect the expected flow regimes for any flow condition in a pipeline. The maps are shown on Figure 5.25a for scenario A and Figure 5.25b for scenario B.
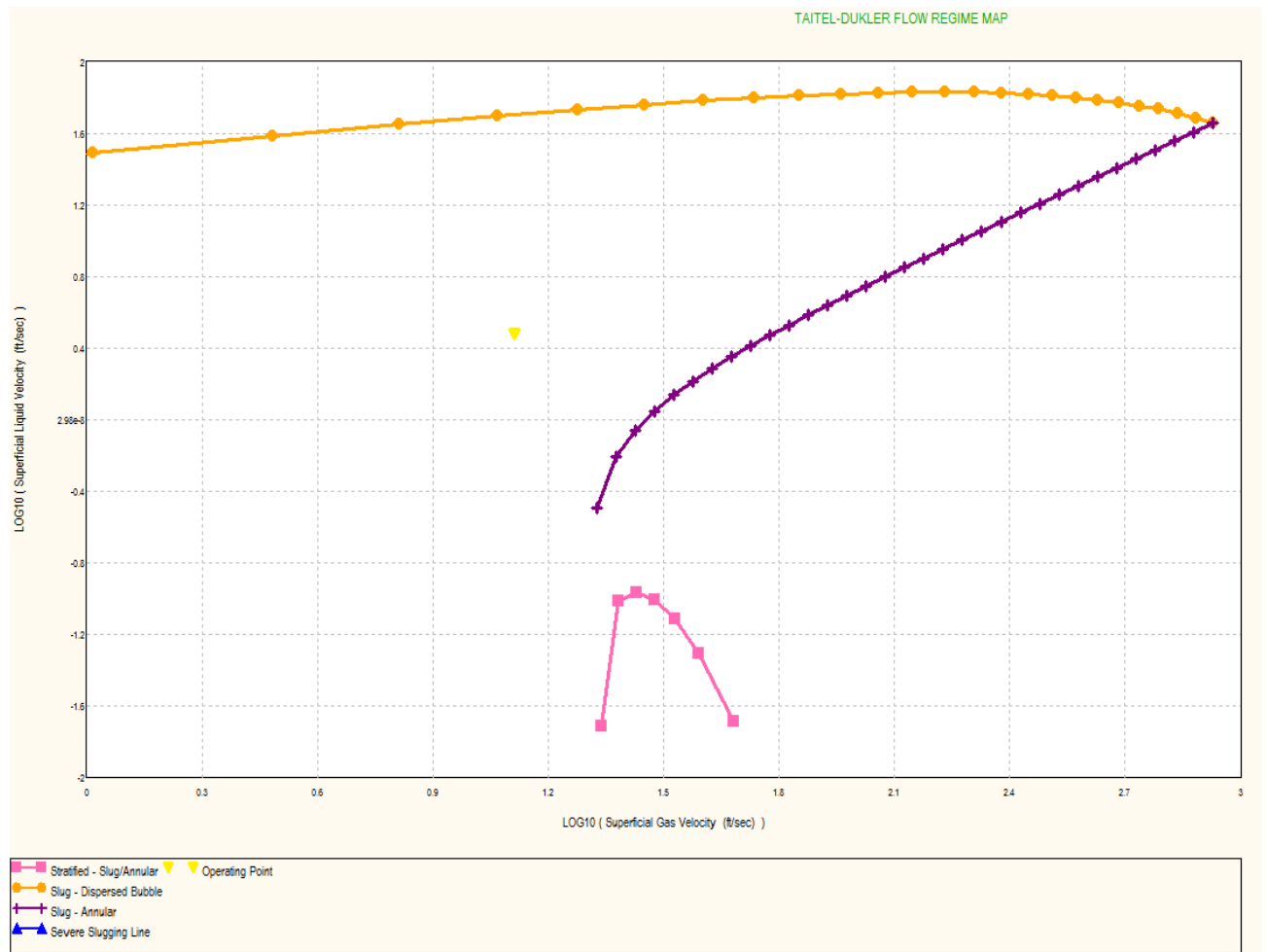
Figure 5.25**a** Flow regime map for scenario A.

From the Taitel-Dukler flow map, it can be seen that GAP predicted a slug flow regime for both scenarios on the simulation start date of 01/12/2010. This is expected because of the flowline's elevation profile as described earlier in the "*pipeline description*" section. The operation point is denoted by the yellow dot on the map.

The superficial liquid and gas velocities for scenario A is 3.0263 ft/s and 12.8633 ft/s respectively. The superficial liquid and gas velocities predicted for scenario B is 1.7088 ft/s and 7.47503 ft/s respectively.

For scenario A, the mean slug length is 231 ft and the mean slug frequency ranges from 89 to 1676 hours; whereas for scenario B, the mean slug length is 237ft and the mean slug frequency ranges from 23 to 29 hours.

This information can be vital for the sizing of slug catchers and/or the production separator in order to prevent liquid loading of the separator from the start

date of the simulation on 01/12/2010 onwards. The results are presented in tabular form in Table 5.22.
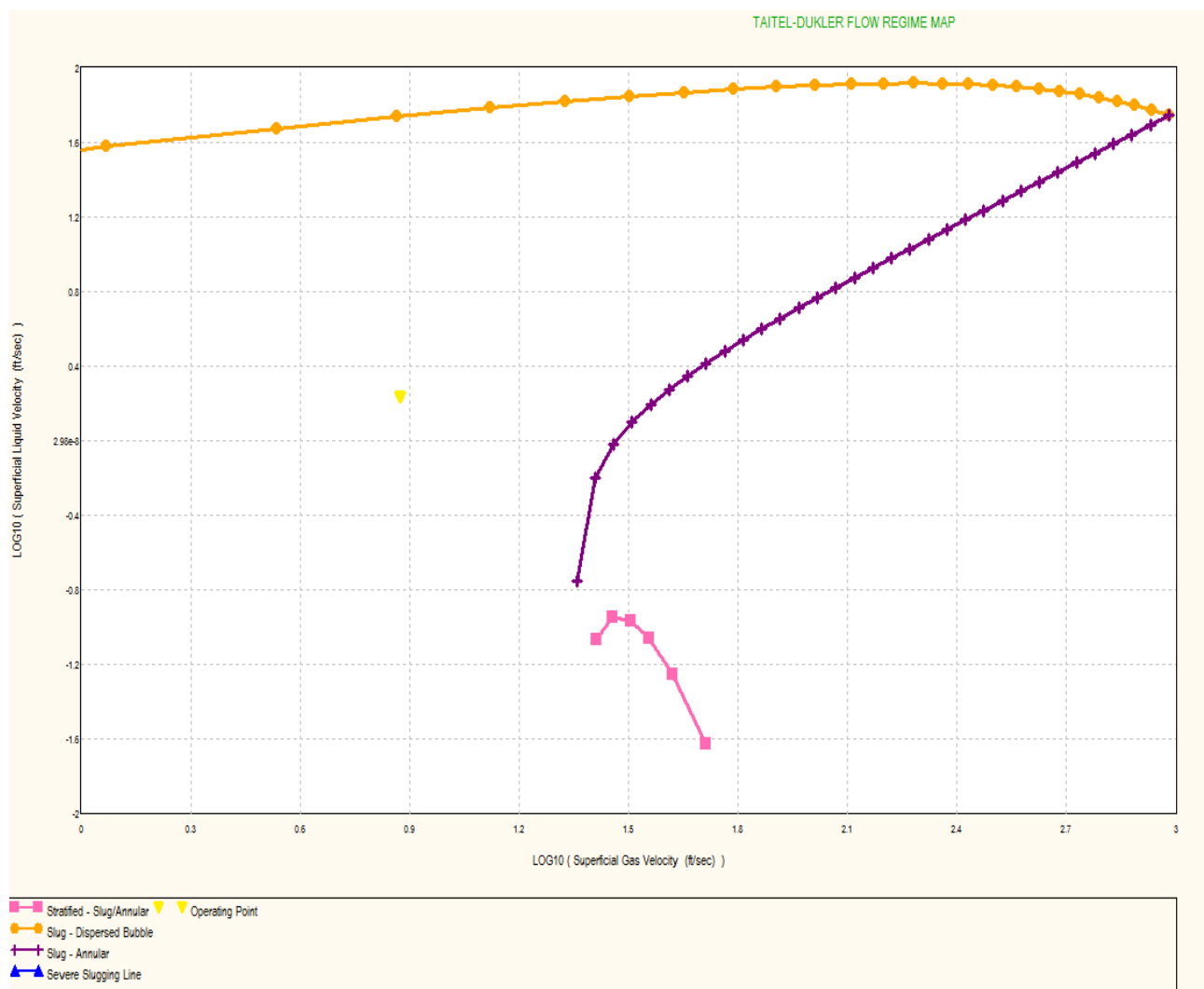
Figure 5.25**b** Flow regime map for scenario B.

Table 5.22 Slugging Conditions for 01/01/2010

Scenario	Mean Slug length	Mean slug Frequency	Flow Regime	Superficial liquid velocity	Superficial Gas velocity
	ft	hr		ft/s	ft/s
Scenario A	231	89 to 1676	Slug	3.0263	12.8633
Scenario B	337	23 to 29	Slug	1.7088	7.47503

The mean slug frequency data shows that, slugs will form in the flowline connecting Manifold B to the separator more frequently for scenario B than scenario A.

GAP can also be used to predict the length and frequency of slugs at any date within the simulation period in case there would be a potential need to upgrade the

separation facility on this proposed date. In this project, the date was chosen to be 01/02/2015, five years after the start of the simulation. The predicted slug length and frequency on this date are presented in Table 5.23.

Table 5.23 Slugging Conditions for 01/02/2015

Scenario	Mean Slug length	Mean slug Frequency
	ft	hr
Scenario A	205	39 to 42
Scenario B	315	22 to 24

The results show that, for scenario A, by 01/02/2015, the slug length would be reduced by 26 ft and its frequency range would increase to 22 to 24 hours whereas, for scenario B, there will be a reduction in slug length by 12 ft and a slight increase in frequency range between 22 to 24 hours. This information can be used to design an upgrade of the production facility and/or slug catcher from 01/02/2015 onwards.

Other Flow Assurance Results

Since the detailed data on the composition of the well effluents were not available, the Black Oil model was used for the estimation of the fluids' properties. For this reason, other flow assurance issues such as those of wax, asphaltene accumulation and of hydrate and scale formation were not calculated by GAP.

Conclusion

- From the results of the simulation, scenario B yielded the highest cumulative production of oil with a total of 33MMstb after pressure support, which is 3.8MMstb, more than that produced by scenario A.

- In monetary terms, if we consider the average oil price (at the time of writing) of $55.30 per barrel, then the gross revenue from production with scenario B will be $1,825M which is about $210M more than the revenue that would be generated from production with scenario A.

- In oil business, the cumulative oil production is not the only determinant of the choice of pipe diameter. Certainly, larger diameters always have higher production and lower pressure drops. However, if the flowrates and velocities are very high comparable to that of the smaller diameter pipes, then the larger diameter pipes can exhibit higher pressure drops as in the case of this project.

- Larger pipe diameters come along with higher cost of design of the pipes due to more materials needed for construction and possible construction limitations like; available space, more support required etc.

- Apart from cumulative oil production, some of the factors taken into consideration when choosing pipe diameter include maximum allowable speed and pressure drop, possible slug flow, water hammer effects, noise etc.

- In this project, scenario B produced the highest cumulative oil recovery as well as the lowest water cut after the 10-year simulation period. Therefore, the choice of pipe diameter for the offshore field at the Loggie Mill Block will be of scenario B.

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APPENDIX A

Figure 3.7 Resistance coefficients
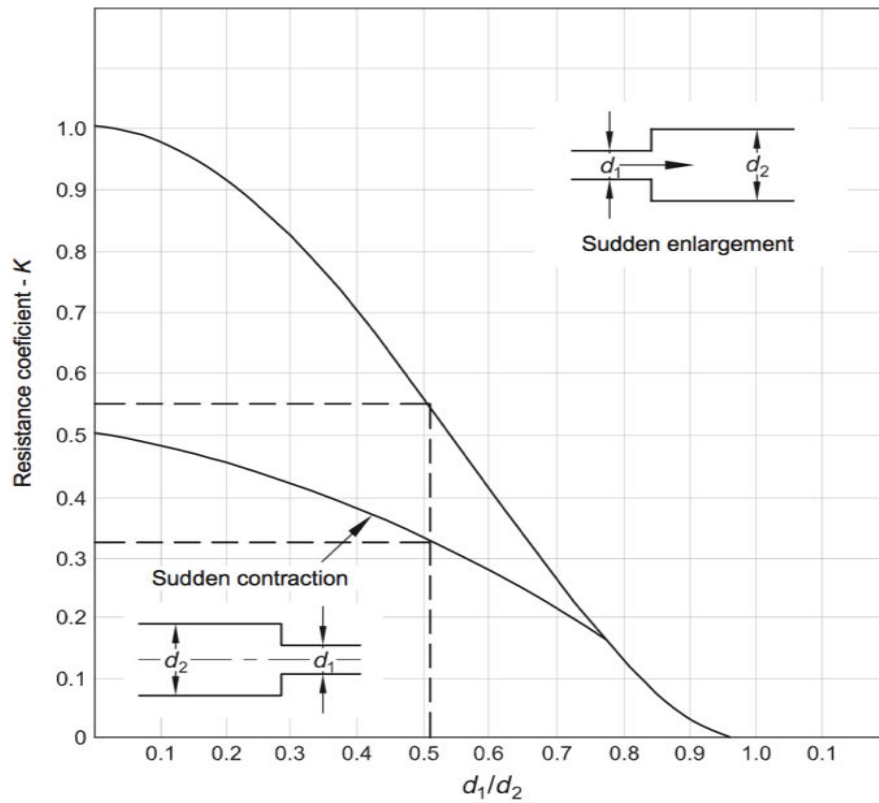
Figure 3.7a Resistance coefficients for sudden contractions and enlargements
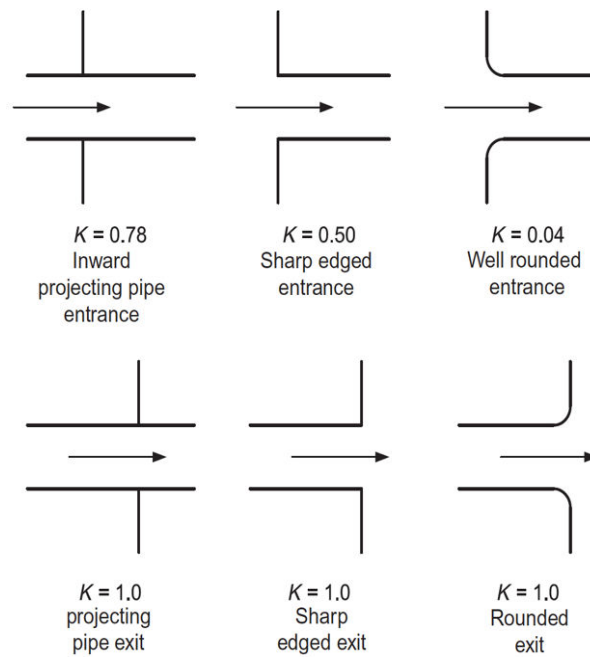
Figure 3.7b Resistance coefficients for pipe entrances and exits.

(Courtesy of Paragon Engineering)

Figure 3.8 Representative Resistance Coefficients

PIPE FRICTION DATA FOR CLEAN COMMERCIAL STEEL PIPE
WITH FLOW IN ZONE OF COMPLETE TURBULENCE

Nominal Size	$\frac{1}{2}$"	$\frac{3}{4}$"	1"	$1\frac{1}{4}$"	$1\frac{1}{2}$"	2"	$2\frac{1}{2}$,3"	4"	5"	6"	8-10"	12-16"	18-24"
Friction Factor (fr)	.027	.025	.023	.022	.021	.019	.018	.017	.016	.015	.014	.013	.012

FORMULAS FOR CALCULATING "K" FACTORS *
FOR VALVES AND FITTINGS WITH REDUCED FOAT

* Formula 1

$$K_2 = \frac{0.8 \left(\sin\frac{\theta}{2}\right) (1-\beta^2)}{\beta^4} = \frac{K_2}{\beta^4}$$

* Formula 2

$$K_2 = \frac{0.5 (1-\beta^4) \sqrt{\sin\frac{\theta}{2}}}{\beta^4} = \frac{K_1}{\beta^4}$$

* Formula 3

$$K_2 = \frac{2.6 \left(\sin\frac{\theta}{2}\right) (1-\beta^2)}{\beta^4} = \frac{K_1}{\beta^4}$$

* Formula 4

$$K_2 = \frac{(1-\beta^4)^2}{\beta^4} = \frac{K_1}{\beta^4}$$

* Formula 5

$$K_2 = \frac{K_1}{\beta^4} + \text{Formula 1} + \text{Formula 2}$$

$$K_2 = \frac{K_1 + \sin\frac{\theta}{2} \left[0.8(1-\beta^4)+2.6(1-\beta^4)^2\right]}{\beta^4}$$

* Use " K " Furnished By Valve Or Fitting Supplier When Available.

* Formula 6

$$K_2 = \frac{K_1}{\beta^4} + \text{Formula 1} + \text{Formula 2}$$

$$K_2 = \frac{K_1 + 0.5 \sqrt{\sin\frac{\theta}{2}} (1-\beta^2) + (1-\beta^2)^2}{\beta^4}$$

* Formula 7

$$K_2 = \frac{K_1}{\beta^4} + \beta (\text{Formula 2} + \text{Formula 4}) \text{ When } -180° \ 0$$

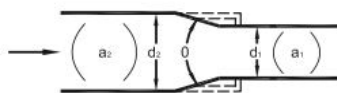$$K_2 = \frac{K_1 + \beta \left[0.5 (1-\beta^2) + (1-\beta^2)^2\right]}{\beta^4}$$

$$\beta = \frac{d_1}{d_2}$$

$$\beta^2 = \left(\frac{d_1}{d_2}\right)^2 = \frac{d_1}{d_2}$$

Subscript 1 Defines Dimensions
End Coefficient With Reference To
The Smaller Diameter.
Subscript 2 Refers To The Larger
Diameter.

SUDDEN AND GRADUAL CONTRACTION

If : $0 \lessgtr 45°$ .... K_2 = Formula 1

$45° < 0 \lessgtr 180°$ ..... K_2 = Formula 2

SUDDEN AND GRADUAL ENLARGEMENT

If : $0 \lessgtr 45°$ .... = Formula 3

$45° < 0 \lessgtr 180°$ ..... = Formula 4

Figure 3.8 Representative resistant coefficients   (*Continued*)

Gate Valves
Wedge Disc, Double Disc, Or Plug Type

If : $\beta = 1.0 = 0$.............................$K_1 = 8fr$
$\beta < 1$ And $0 \gtrless + 45°$..............$K_1 = $ Formula 5
$\beta < 1$ And $45° < 0 \gtrless 180°$.....$k_1 = $ Formula 6

Swing Check Valves

$K = 100fr$
Minimum Pipe Velocity
(fps) For Full Disc Lift
$= 35 \quad \sqrt{V}$

$K = 50fr$
Minimum Pipe Velocity
(fps) For Full Disc Lift
$= 60 \quad \sqrt{V}$ Except

U/L Insert $= 100 \quad \sqrt{V}$

Globe And Angle Valves

If : $\beta = 1....K_1 = 340 \; fr$

If : $\beta = 1....K_1 = 55 \; fr$

If : $\beta = 1...K_1 = 150 \; fr$ If : $\beta = 1....K_1 = 55 \; fr$

All Globe And Angle Valves,
Whether Reduced Seat Ot Throttled
If : $\beta < 1...K_2 = $ Formula 7

Lift Check Valves

If : $\beta = 1...K_1 = 600fr$
$\beta < 1...K_2 = $ Formula 7
Minimum Pipe Velocity (fps) For Full Disc Lift
$= 40 \; \beta^2\sqrt{V}$

If : $\beta = 1...K_1 = 600fr$
$\beta < 1...K_2 = $ Formula 7
Minimum Pipe Velocity (fps) For Full Disc Lift
$= 140 \; \beta^2\sqrt{V}$

Tilting Disc Check Valves

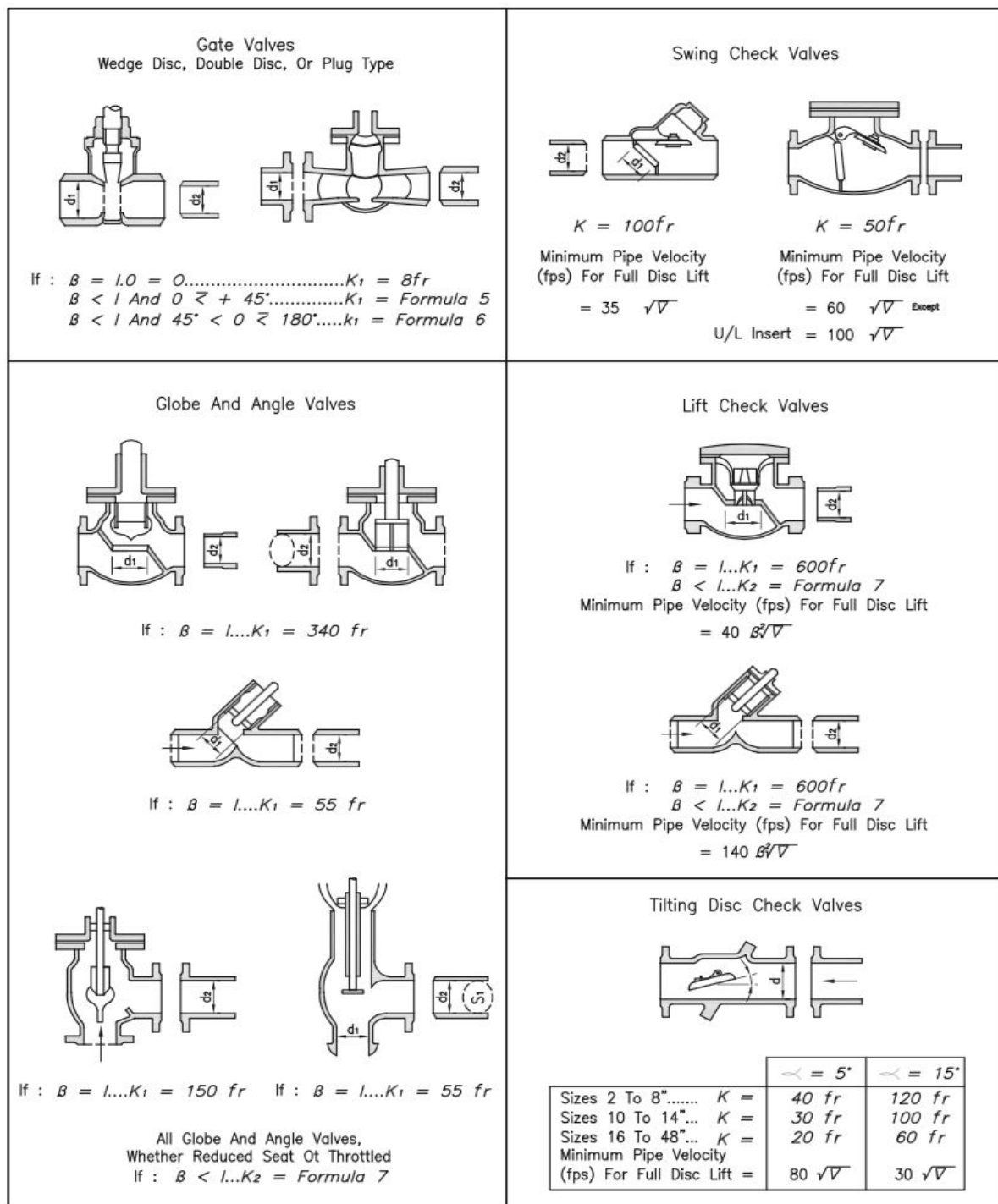|	$\sphericalangle = 5°$	$\sphericalangle = 15°$
Sizes 2 To 8"....... $K =$	$40 \; fr$	$120 \; fr$
Sizes 10 To 14"... $K =$	$30 \; fr$	$100 \; fr$
Sizes 16 To 48"... $K =$	$20 \; fr$	$60 \; fr$
Minimum Pipe Velocity (fps) For Full Disc Lift $=$	$80 \; \sqrt{V}$	$30 \; \sqrt{V}$

Figure 3.8 Representative resistant coefficients   (*Continued*)

Stop-Check Valves (Globe And Angle Types)

If :
$\beta = 1...K_1 = 400fr$
$\beta < 1...K_2 = $ Formula 7

Minimum Pipe Velocity For Full Disc Lift $= 55 \, \beta^2\sqrt{V}$

If :
$\beta = 1...K_1 = 200fr$
$\beta < 1...K_2 = $ Formula 7

Minimum Pipe Velocity For Full Disc Lift $= 55 \, \beta^2\sqrt{V}$

Foot Valves With Strainer

Peppet Disc
$K = 420fr$
Minimum Pipe Velocity (fps) For Full Disc Lift $= 15 \, \sqrt{V}$

Hinged Disc
$K = 75fr$
Minimum Pipe Velocity (fps) For Full Disc Lift $= 25 \, \sqrt{V}$

If :
$\beta = 1...K_1 = 300fr$
$\beta < 1...K_2 = $ Formula 7

If :
$\beta = 1...K_1 = 350fr$
$\beta < 1...K_2 = $ Formula 7

Mnimum Pipe Velocity (fps) For Full Disc Lift $= 60 \, \beta^2\sqrt{V}$

Ball Valves

If :
$\beta = 1.0 = 0..........K_1 = 3fr$
$\beta < 1$ And $0 = 45°.........K_2 = $ Formula 5
$\beta < 1$ And $45° < 0 = 180°...K_2 = $ Formula 6

If :
$\beta = 1...K_1 = 55fr$
$\beta < 1...K_2 = $ Formula 7

If :
$\beta = 1...K_1 = 55fr$
$\beta < 1...K_2 = $ Formula 7

Mnimum Pipe Velocity (fps) For Full Disc Lift $= 140 \, \beta^2\sqrt{V}$

Butterfly Valves

Size 2 To 8" ... $K = 45fr$
Size 10 To 14" ... $K = 35fr$
Size 16 To 24" ... $K = 25fr$

Figure 3.8 Representative resistant coefficients   (*Continued*)

99

Plug Valve And Cocks

Straight—Way   3—Way

View X—Y

If : $\beta = l_1$
$K_1 = 18fr$

If : $\beta = l_1$
$K_1 = 30fr$

If : $\beta = l_1$
$K_1 = 90fr$

If : $\beta < l ... K_2 = $ Formula 6

Mitre Bends

⊥	K
0°	2 fr
15°	4 fr
30°	8 fr
45°	15 fr
60°	25 fr
75°	40 fr
90°	60 fr

90° Pipe Bends And Flanged Or Butt—Welding 90° Elbows

r/d	K	r/d	K
1	20 fr	8	24 fr
1.5	14 fr	10	30 fr
2	12 fr	12	34 fr
3	12 fr	14	38 fr
4	14 fr	16	42 fr
6	17 fr	20	50 fr

The Rsistunce Coefficience, .For Pipe Bends Other Than 90° May Be Determined Follows :

$$K_a = (n-1)\left(0.25\ r\ fr_d + 0.5K\right) + K$$

n = Number Of 90° Bends
K = Resistance Coeficience For one 90° Bends (Per Table)

Close Pattern Return Bends

$K = 50\ fr$

Standard Elbows

90°   45°

$K_1 = 30fr$ $K_1 = 16fr$

Standard Tees

Flow Thru Run....... $K_1 = 20fr$
Flow Thru Branch... $K_1 = 60fr$

Pipe Entrance

Inward Projecting   Flush

r/d	K
0.00°	0.5
0.02°	0.28
0.04°	0.24
0.06°	0.15
0.10°	0.09
0.15 & Up	0.04

Sharp—Edged

$K = 0.78$

For K.
See Table

Pipe Exit

Projecting   Sharp—Edged   Rounded

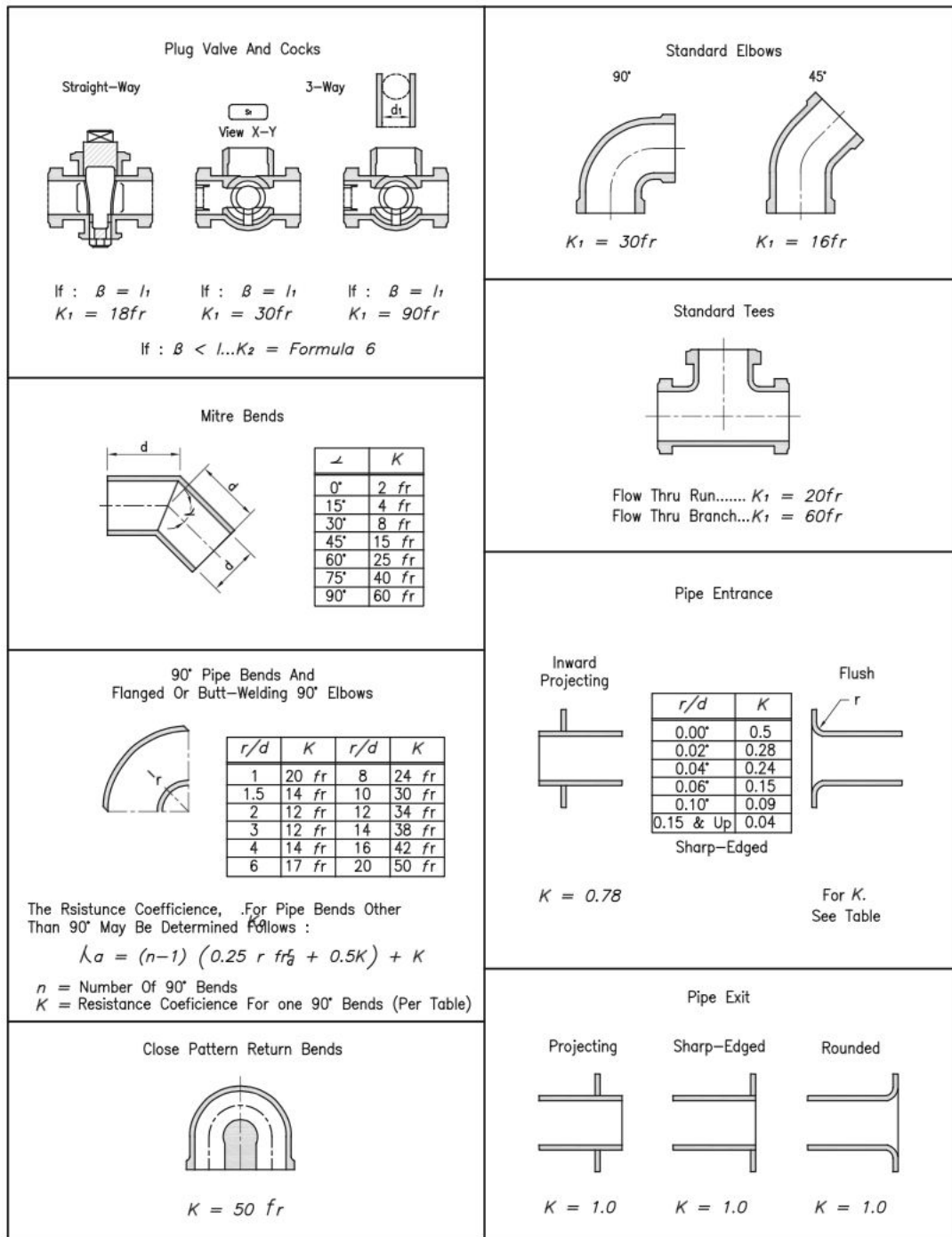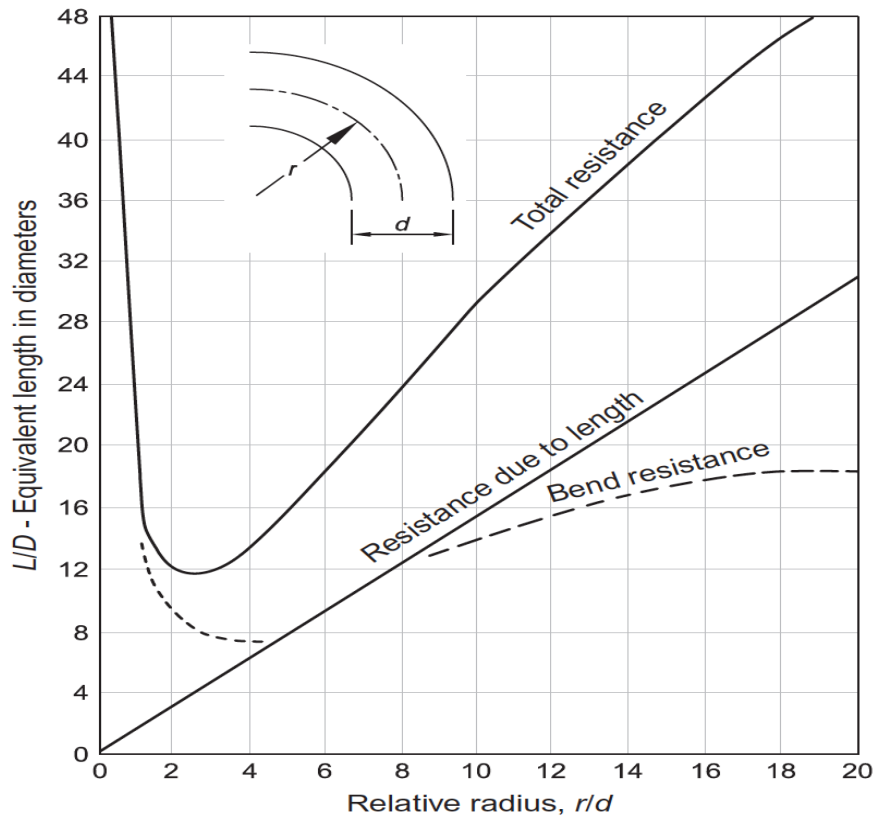$K = 1.0$ $K = 1.0$ $K = 1.0$

Figure 3.8 Representative resistant coefficients   (*Continued)*

Table 3.3 Equivalent length of valves and fittings

Nominal pipe size in.	Globe valve of ball check valve	Angle valve	Swing check valve	Plug valve	Gate or ball valve	45° all Weld thrd	Short red. all Weld thrd	Long red. all Weld thrd	Branch of tee Weld thrd	Run of tee Weld thrd	90° miter bends 2 miter	3 miter	4 miter	Enlargement Sudden d/D=1/4	Enlargement Sudden d/D=1/2	Enlargement Sudden d/D=3/4	Enlargement Std. red d/D=1/2	Enlargement Std. red d/D=3/4	Contraction Sudden d/D=1/4	Contraction Sudden d/D=1/2	Contraction Sudden d/D=3/4	Contraction Std. red d/D=1/2	Contraction Std. red d/D=3/4
1½	55	26	13	7	1	1 2	3 5	2 3	8	2 3				5	3	1	5	1	3	2	1	1	-
2	70	33	17	14	2	2 3	4 5	3 4	9	3 4				7	4	1	6	1	3	3	1	1	-
2½	80	40	20	11	2	2-	5-	3-	10 11	3-				8	5	2	8	2	4	3	2	2	-
3	100	50	25	17	2	2	6	4	12	4				10	6	2	10	2	5	4	2	2	-
4	130	65	32	30	3	3	7	5		5				12	8	3	14	3	6	5	3	3	-
6	200	100	48	70	4	4	11	8	14 19	8				18	12	4	19	4	9	7	4	4	1
8	260	125	64	120	6	6	15	9	28	9				25	16	5	24	5	12	9	5	5	2
10	330	160	80	170	7	7	18	12		12				31	20	7	28	7	15	12	6	6	2
12	400	190	95	170	9	9	22	14	37 47	14	28	21	20	37	24	8	-	8	18	14	7	7	-
14	450	210	105	80	10	10	26	16	55	16	32	24	22	42	26	9	-	-	20	16	8	-	2
16	500	240	120	145	11	11	29	18		18	38	27	24	47	30	10	-	-	24	18	9	-	-
18	550	280	140	160	12	12	33	20	62 72	20	42	30	28	53	35	11	-	-	26	20	10	-	-
20	650	300	155	210	14	14	36	23	82	23	46	33	32	60	38	13	-	-	30	23	11	-	-
22	688	335	170	225	15	15	40	25		25	52	36	34	65	42	14	-	-	32	25	12	-	-
24	750	370	185	254	16	16	44	27	90 100	27	56	39	36	70	46	15	-	-	35	27	13	-	-
30	-	-	-	312	21	21	55	40	110	40	70	51	44									-	
36	-	-	-		25	25	66	47		47	84	60	52									-	
42	-	-	-		30	30	77	55	140 170	55	98	69	64										
48	-	-	-		35	35	88	65	200	65	112	81	72										
54	-	-	-		40	40	99	70		70	126	90	80										
60					45	45	110	80	220 250 160	70													

Courtesy of GPSA Engineering Data Book.

Figure 3.9 Equivalent lengths of 90° bends.

Courtesy of Crane Technical Paper No. 410.
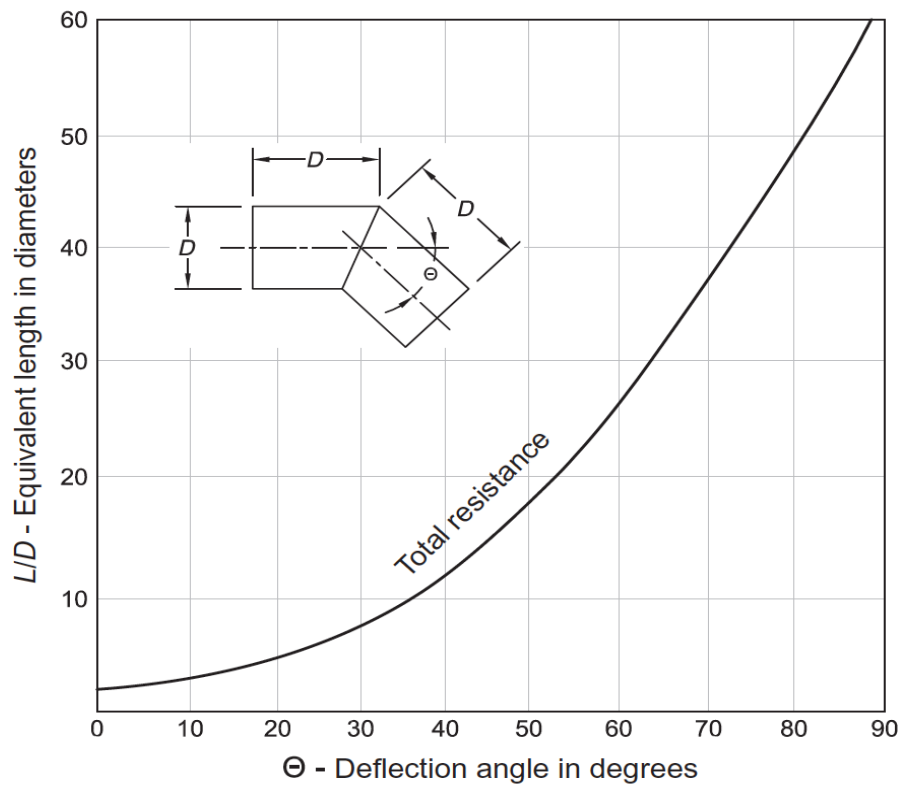
Figure 3.10 Equivalent length of miter bends.

Courtesy of Crane Technical Paper No. 410.