Investigation of Multiphase Flow in Ashuganj-Bakhrabad-Demra Pipeline

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This thesis is dedicated to

My Parents

Late Sarwar Jahan

and

Hosne Ara Jahan

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ABSTRACT

Ashuganj-Bakhrabad-Demra pipeline is operated by Gas Transmission Company Limited (GTCL) and it carries almost 40% of the natural gas from North Eastern part to central region i.e. Dhaka region of Bangladesh. The pipeline length is 130 km and diameter varies from 30 inches to 20 inches. From operational point of view it is very important to study the flow behavior and pressure profile as it supply gas to many important installations like power plants, fertilizer factory and other industries. Multiphase flow variables are key parameters for studying the flow pattern in terms of Two-Phase flow. Before studying the two phase flow in this pipeline, measurement of condensate for each flow rates at different locations are important.

The objective of this work is to investigate multiphase flow in Ahuganj-Bakhrabad-Demra(ABD) pipeline with the data provided by GTCL. In this thesis, theoretical amount of condensate generated was calculated by the Flash and Heat Balance method by assuming that gas compositions were not changed significantly over a period of time. Then the predicted condensate values were compared with the measured values given by the GTCL and found Flash method calculated the volume of condensate with 25% error. A trend analysis was done and found that flash correlation predicts the measured trend better than other correlation Then measured condensate values collected from GTCL were used to calculate the Two-Phase flow parameters. Here Taitel and Duckler stratified equilibrium model for horizontal and near horizontal pipes was used. Flow patterns for different pipe segments were evaluated on the basis of Taitel and Duckler model Stratified Flow pattern was found in every section of the pipe. After establishing the flow pattern, suitable pressure predicting correlations were identified for Two Phase and Single phase flow from the literature.

Pressure calculation by various methods (Marching Algorithm, Weymouth, Panhandle A, Panhandle B and by Fekete Software) were done for each segments. Comparison between calculated pressure and measured pressure data gave a clear idea about application of the correlations and limiting factors. Error analysis was done for different correlation to find a suitable correlation for a segment of the pipe. No single correlation found best for the enter length of the pipe. Panhandle B was found the best performing pressure correlations for Segment-1, Segment-2 where absolute average error values for all the correlations were less than 5%.

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A-B-D Line = Ashuganj-Bakhlabad-Demra Line

AGMS = Ashuganj Gas Metering Station

BKB = Bakhrabad

d = Diameter, ft

GTCL = Gas Transmission Company Limited

l = Length, ft

P = Pressure, Psia

Q = Flow Rate, MMSCFD

Re = Reynold's Number

T = Temperature, ^{0}R

v = Velocity, ft/s

w = Mass of Fluid, lb

Z = Compressibility Factor

μ = Viscosity, cp

ω = Acentric Factor

γ = Specific Gravity of Fluid

ρ= Density, lb/ft^3

CHAPTER 1
INTRODUCTION

Transporting natural gas through pipeline is the easiest means to transfer it from one location to further distant locations. Pipelines carrying natural gas usually buried underground and operate under higher pressure. Other means of transporting natural gas are liquefied form that is known as LNG (Liquefied Natural Gas) and hydrate form where gas is allowed to mix with water to form hydrate. Hydrate form is still under experimental stage. LNG and Hydrate requires huge capital investment and moderate operating cost. In contrary, gas pipeline needs huge initial cost but operating cost is very low.

Since natural gas field was discovered in 1962, at first natural gas was used in Chattak Cement Factory by transporting it through pipeline. Later on it was used in power generation and fertilizer production. In recent years it is used in industries, domestic households and transport sector. Bangladesh already has an established pipeline network and Ashuganj-Bakhrabad-Demra (ABD) pipeline is the part of main trunk line, which operates at 1000 psig pressure. It mainly transports gas from northeastern part of Bangladesh to central and southeastern part of Bangladesh. At Bramnbaria, this line divides into two parts. One part goes to southeastern part i.e Comilla and Chittagong and other part goes to central part i.e. Dhaka, Gazipur and Narayanganj area. Total length of the ABD pipeline is almost 130 km and diameter varies from 30 to 20 inches. Most of the industries, power plant and fertilizer factory depend on this line. Any operation problems in this pipeline will severely affect the power sector and the industries.

When operating parameters i.e. flow rate, pressure and temperature changes or any operating condition change in the processing plants, liquid separate out from the gas stream because of multi- component nature of natural gas and its associated phase behavior. The separated out liquid might accumulate in the pipeline or carry over by the gas stream. If the accumulated liquid amount gets higher, multiphase flow might occur in the pipeline or section of the pipeline. This significant liquid volume flowing simultaneously with gas contributes to the overall pressure loss in pipelines. In such situations, the single- phase equations are inapplicable for design of a pipeline and prediction of pressure losses. Two-phase flow in pipes generally causes a significantly higher-pressure drop than the equivalent single-phase flow, even though the total mass flow rates are the same. Gas Transmission Company

Limited (GTCL) sometime having complains about excessive condensate in pipeline gas from her bulk customers like power plants, fertilizer factories and other industries, which cause them to shutdown the plants. These require a complete understanding of the multiphase flow behavior in pipeline and detail study about multiphase flow parameters, flow regimes and pressure profiles for a pipeline or segments of the pipe.

1.1 Objectives:

The objectives of the study are as follows:

a. Build sound knowledge about two phase flow parameters and different types of flow.
b. Identify possible condensate build-up and calculate build-up rate in ABD pipeline.
c. Investigate the effect of condensate in gas flow in terms of two phase flow relationships and validate the two phase flow correlations.
d. Investigate the effect of pressure loss calculation by two phase and single phase correlations.

1.2 Methodology:

Daily production data of the different gas fields and average input to the Ashuganj Metering Station (AGMS) data were collected from the GTCL. This daily report contains the operation variables like pressure, temperature and gas flow rates. As natural gas is multi-component in nature. one chromatographic test was done at Demra outlet point in the Petroleum and Mineral Resource Engineering (PMRE) laboratory. If there is any condensate separation-taking place, this will have seen in the difference of heavier components in chromatographic results. Condensate production data were collected from Titas Gas Transmission and Distribution Company Limited (TGTDCL). From flash calculation theoretical value of condensate was calculated and compared with the measured data. As the amount of condensate was determined, evaluating Two-Phase flow parameters checked its distribution in gas phase – flow regime. If the flow regime suggests two-phase flow then "The Marching Algorithm" calculated pressure drop of the pipeline. The single-phase equations like Weymouth. Panhandle A and Panhandle B also calculate pressure drop. If temperature data are available along the pipeline, a heat balance method could be a good check for condensate calculations. Later on a statistical analysis was carried out to find the best correlation for pressure drop and condensate calculation.

Ashuganj-Bakhrabad-Demra (ABD) line was divided into 3 (three) segments for calculation purposes. Table 1.1 gives them below.

Table 1.1: Description of ABD line.

No. of the segment	Name of the segment	Transmission Capacity, MMSCFD	Operating Maximum Pressure, Psig	Length, km	Diameter, inch
1	Ashuganj Gas Metering Station (AGMS) to Bangura junction	330	1000	40	30
2	Bangura junction to Bakhrabad	425	1000	18.5	30
3	Bakhrabad to Demra	250	400	68.72	20

From the above table, it is clear that; transmitted amount of gas exceeds the transmission capacity of the first two (2) segments, where transmitted amount of gas becomes one-eighth of the available capacity of gas in segment 3.

On the next page, Figure 1.1 shows the schematic view of the ABD pipeline, where major input and output were shown only. Pipeline was assumed straight line and undulation of pipeline was ignored.

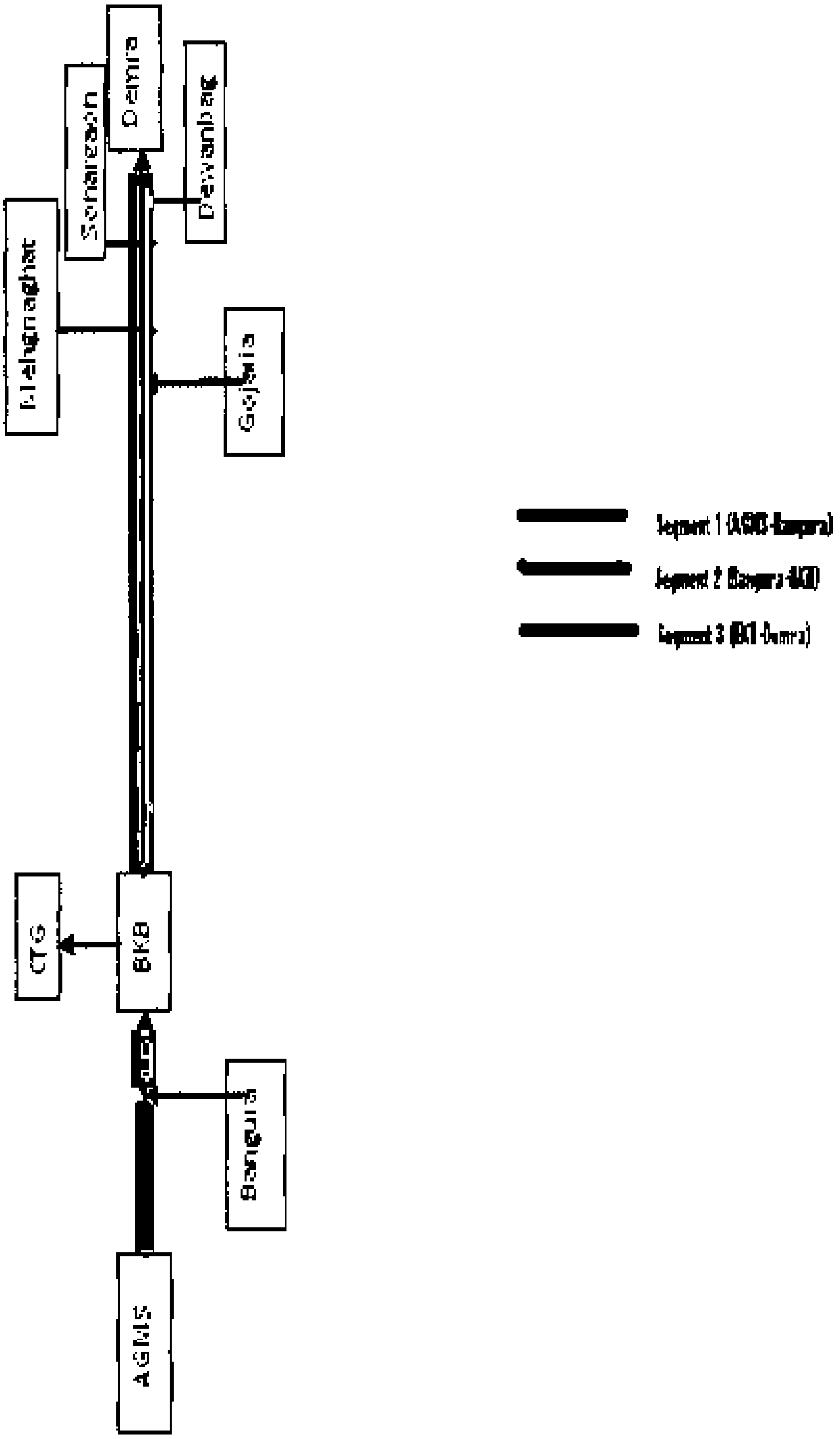

Fig 1.1: The Schematic Diagram of Ashuganj Bakhrabad and Demra pipeline.

CHAPTER 2
LITERATURE REVIEW

The two-phase flow occurs in a pipe when two phases flow thorough the pipe and the phases tend to separate because of their difference in density and viscosity. Shear stresses at the pipe wall would be different for each phase. Expansion of highly compressible gas phase occurs as it travels a long distance pipeline with decreasing pressure. As a result gas and liquid phases do not travel at the same speed; less viscous gas phase tends to move faster than liquid phase causing a phenomenon known as slippage.

The basis for fluid flow calculations in pipes is conservation of mass, momentum and energy. Application of these principles allows the calculation of changes in pressure and temperature with distance. The hydrodynamic behavior of two-phase systems is radically different from that of single-phase systems. Perhaps the most distinguishing aspect of multiphase flow is the variation in the physical distribution of the phases in the flow conduit, a characteristic known as flow pattern or flow regime. During multiphase flow through pipes, the flow pattern that exists depends on the relative magnitude of the forces that act on the fluids. Buoyancy, turbulence, inertia, and surface tension forces vary significantly with flow rates, pipe diameter, inclination angle, and fluid properties of the phases. Thus two-phase system not only exhibit representative true pressure loss, but also are subjected to mass-generation-induced forces and inter phase forces, all of which are completely absent in single-phase systems. Other system variables needed to define a two-phase system are liquid holdup, phase velocity, phase thermo physical properties, and the relative spatial distribution of the phases known as flow regime. [1]

Several different flow patterns may exist in a long pipeline if there are large pressure and temperature changes. Especially important parameter in two-phase flow is the variation in pressure gradient with flow pattern. Before calculating the flow pattern, one should know how much liquid is producing in the pipeline with specific condition or change in operation variables. Once the mass generation of liquid is known, it would be readily easier to analyze two-phase flow correlations. In the later sections, essential fluid parameters are discussed in details.

2.1 Equilibrium Constant:

For a multi-component system, such as petroleum fluids, the composition, pressure, and temperature uniquely define the system phase behavior. The equilibrium constant, K_i, of a component i is defined as the ratio of the mole fraction of the component in the gas phase, y_i, to the mole fraction of the same component in the liquid phase, x_i and expressed mathematically as follows:

$$K_i = \frac{y_i}{x_i} \tag{2.1}$$

For ideal solutions at low pressures (typically below 75 psia), an ideal equilibrium constant can be derived by combining Raoult's law and Dalton's law. This can be expressed mathematically for a fixed temperature as

$$K_i = \frac{y_i}{x_i} = \frac{P_{vi}}{P} \tag{2.2}$$

Where P_{vi}= vapor pressure of component i, psia, and P = total system pressure, psia.

The vapor pressure of any pure hydrocarbon is a unique function of temperature. Eq. 2.2 assumes the liquid phase behaves as an ideal solution following Dalton's law. For a real solution like hydrocarbon mixture, the equilibrium constants are not only functions of pressure and temperature but also of the composition of the hydrocarbon phases and assumptions used in deriving the Eq. 2.2 become unrealistic. A hydrocarbon / liquid mixture at any point in the pipe is in equilibrium with a vapor mixture at the local pressure and temperature. Thus the equilibrium constant or K value depend on all these parameters.

$$K = f(P, T, Z_i) \tag{2.3}$$

Where Z_i is the i phase composition.

These equilibrium constants depend on the fugacity of the each component in the each of the phases. Equal fugacity of a component in each phase implies zero net mass transfer of that component between the phases, resulting thermodynamic equilibrium of the phases. Thus at thermodynamic equilibrium, fugacity of every components in each phase are equal.

$$f_i^v = f_i^L \tag{2.4}$$

Where, f is the fugacity, v = vapor and L = liquid phase. From the definition of the fugacity, fugacity has a unit of pressure. Introducing vapor pressure to represent the escaping tendency of molecules from one phase into the other can modify fugacity. Fugacity coefficient, the ratio of fugacity to system pressure, of a mixture is thermodynamically vapor/ liquid equilibrium. The equilibrium constant can be found with

$$K_i = \frac{y_i}{x_i} = \frac{\phi^L}{\phi^v} \tag{2.5}$$

The Equation of State (EOS) defines the fugacity coefficients for each component in each phase.

2.2 Flash Calculations:

The multi-component flash of a liquid may be visualized as a simple distillation process using a single equilibrium stage. Flash calculations are based on simple overall and component molal balance. But multi component flash calculation is very different and more complex than the flash evaporation of single-component liquid because of its overall convergence criterion in determining the equilibrium constant. For a multi-component liquid, calculating the amounts of flashed vapor and residual liquid in equilibrium at a given temperature and pressure requires a trial-and-error iterative solution techniques. Such a calculation is commonly referred to as equilibrium flash calculation. It involves solving the *Rachford-Rice equation*:

$$f\left(\frac{V}{F}\right) = \sum_i^n (y_i - x_i) = \sum_i^n \frac{z_i(K_i - 1)}{(K_i - 1)\beta + 1} \tag{2.6}$$

Where:

- z_i is the mole fraction of component i in the feed liquid (assumed to be known);
- β is the fraction of feed that is vaporized, V/F;
- K_i is the equilibrium constant of component i.

The equilibrium constants K_i are in general functions of many parameters, though the most important is arguably temperature; they are defined as:

$$y_i = K_i x_i$$

Where:

- x_i is the mole fraction of component i in liquid phase;

- y_i is the mole fraction of component i in gas phase.

Once the Rachford-Rice equation has been solved for β, the compositions x_i and y_i can be immediately calculated as:

$$x_i = \frac{z_i}{1 + \beta(K_i - 1)}$$
$$y_i = K_i\, x_i.$$

The equilibrium flash of multi-component liquids is very widely used in petroleum refineries, petrochemical and chemical plants and natural gas processing plants. Flash calculation resulted in the molar distribution of a component in each phases [1]. Detailed calculation was given in the Appendix A. A.2.

2.3 Condensate Calculation by Heat Balance Method.

In terms of energy balance, transfer of energy takes place in the system from gaseous to liquid phase–through liquid as a wetting phase at the pipe wall to surroundings or vice versa. Transfer of energy occurs in the form of heat exchange or of acceleration. In gas pipeline energy transfer is considered in the form of heat. Gas pipelines are generally buried underground and wrapped with thick poly vinyl chloride (PVC) tape, which does not give total insulation to the pipe. During the flow of gas in the pipeline, frictional loss causes pressure drop due to roughness of the pipe and of the gas-liquid interface. Though gas viscosity is low but for a long pipe section, pressure drop term is quiet significant resulting changes in physical properties. As the pressure drops, temperature also drops which causes heavier gas molecules to condense. As the gas is transferred through a long pipeline, percentage of condensable molecules gets leaner if they present significantly in the gas stream. Chromatographic analysis of gas samples at two points of the pipeline supports this and showed a noticeable change of molar percentage for the component above Butane (C_4+).

As the pipeline is not totally insulated and representative temperature of the pipeline is difficult to get. Moreover pipeline does not have the provision to measure the temperature locally; it is convenient to use theoretical temperature for condensate calculation. Amount of condensate (molar flow rate) was also calculated with the help of the equation,

$$Q=\sum nH= n \sum Cp \, \Delta T$$

Where,

Q= Heat of Condensation, n = number of moles, H= Heat of Condensation, C_P= Specific Heat at Constant Pressure [2].

2.4 Flow Pattern Calculation:

When amount of condensate is calculated, it is essential to study the effect of condensate in terms of two-phase flow. In two-phase flow, liquid hold up and flow pattern are the most important parameters. Flow pattern shows the distribution of the phases in the system. Many empirical correlations have been developed to predict flow pattern considering physical properties, geometrical variables and operational parameters. All design variables like liquid holdup, pressure gradient, heat and mass transfer coefficient, residence time distribution etc. strongly depend on flow pattern. Most simplistic unified flow pattern approach proposed by Shoham (1982) was followed. In early time, liquid hold up is calculated by the considering no slippage between phases. The no slip approach did not recognize two phases rather than consider two phases a single homogeneous phase. Later on introducing the superficial velocities recognizes the slippage introduced in the hold up calculation and relative phase velocity. Among all the correlations, the Taitel and Dukler [3] model is more applicable for steady state, fully developed Newtonian flow in horizontal and slightly inclined pipes, namely ±10°. Transient, entrance and exit effects can cause deviation from the prediction. Flow pattern prediction calculation was given in Appendix A, A.3.

2.5 Pressure Calculation:

Pressure calculation for single-phase flow is based on the conservation of the mass and momentum. Similar concept is used to calculate the two-phase pressure drop considering slip velocity of the phases. As two phases are involved, the pressure calculations are more complicated than single phase one. Liquid hold up term and flow patterns are frequently used to predict the pressure drop as correctly as one could. In early days, numerous flow patterns were used considering the no slip and slip. Now an acceptable set of flow pattern maps are used which has been developed for vertical, horizontal and upward and downward inclination angles. For horizontal pipe Duckler et al., Beggs and Brill and Taitel and Duckler models are extensively used. In this thesis, Taitel and Duckler model is used to predict the flow pattern boundary and pressure gradient. Pipeline is divided into several segments and Marching Algorithm [3] technique is used for pressure drop calculation. In two-phase flow where slip velocity does not occur i.e. stratified flow any single phase models like Weymouth, Panhandle A, Panhandle B etc. can be used to calculate the pressure gradient [4]. In two-phase flow, equations become complex and numerical analysis techniques like Iteration/ Marching Algorithm is useful. In commercial software like Fekete, Panhandle – B is used for pressure calculations.

Table 2.5.1: Assumptions of Various Pressure Equations.

No. of Pressure Equations	Name of Pressure Equations	Assumptions
1	Marching Algorithm/Iteration	a. Two-phase liquid flow with compressible gas phase under isothermal conditions. b. Total pipeline was divided into 20 (twenty) segments.
2	Weymouth	a. No mechanical work. b. Isothermal Steady –State flow. c. Constant Gas Compressibility Factor. d. No undulation. e. Negligible kinetic energy change. f. Fully turbulent flow in pipe with diameters around NPS 36.
3	Panhandle A	a. Pipe diameter from NPS 6 to NPS 24. b. Reynolds number greater than 300,000 with partially turbulent flow.
4	Panhandle B	Long pipelines with diameter greater than NPS 24.

No. of Pressure Equations	Name of Pressure Equations	Assumptions
5	Fekete Software (Piper)	a. Single phase flow using Panhandle B as governing equation. b. As necessary data for Sonargaon and Gojaria take-off points were absent, flow rate and pressure at these points were merged with Dewanbag take-off point.

Weymouth Equation :

$$q_{sc} = 5.6353821 \left(\frac{T_{sc}}{P_{sc}}\right)\left(\frac{(P_1^2 - P_2^2)d^5}{\gamma_g Z_{av} TfL}\right)^{0.5}$$

Here, q_{sc} in MSCFD.

Panhandle A Equation :

$$q_{sc} = 32.6491 \left(\frac{T_{sc}}{P_{sc}}\right)^{1.07881}\left(\frac{P_1^2 - P_2^2}{Z_{av} TfL}\right)^{0.5394}\left(\frac{1}{\gamma_g}\right)^{0.4606}\frac{d^{2.61821}}{\mu_g^{0.07881}}$$

Here, q_{sc} in ft^3/s.

Panhandle B Equation :

$$q_{sc} = 109.364 \left(\frac{T_{sc}}{P_{sc}}\right)^{1.02}\left(\frac{P_1^2 - P_2^2}{Z_{av} TfL}\right)^{0.51}\left(\frac{1}{\gamma_g}\right)^{0.49}\frac{d^{2.53}}{\mu_g^{0.02}}$$

Here. q_{sc} in ft^3/s.

More on pressure calculation and basic assumptions are given in Appendix A, A.4.

2.6 Statistical Analyses:

Any correlation that is presented in this study should be checked statistically in order to obtain a quantitative measurement about the accuracy of the prediction. Some basic statistical parameters used for correlation performance evaluation are average percentage relative error

(APRE), average absolute percentage relative error (AAPRE) and standard error of estimate (SEE).

Average Percentage Relative Error (APRE):

APRE is a measure of the relative deviation of the predicted values from the experimental values in the percentage. The equation is given as follows,

$$\text{APRE} = \left(\frac{1}{n}\right) \text{x} \sum \left[\frac{\{(Ypred - Y\,exp)x100\}}{Y\,exp}\right]$$

The smaller the error is the more evenly distributed the positive and negative differences between predicted and experimental values.

Average Absolute Percentage Relative Error (AAPRE):

AAPRE is the absolute measure of the relative deviation of the predicted values from the experimental values in the percentage. The equation is given as follows,

$$\text{AAPRE} = \left(\frac{1}{n}\right) \text{x} \sum \left| \left[\frac{\{(Ypred - Y\,exp)x100\}}{Y\,exp}\right] \right|$$

Standard Error of Estimate (SEE):

SEE is the square root of the mean square error, which is the variance of the true residuals. It is expressed as

$$\text{SEE} = \sigma = \left(\frac{1}{n-v-1}\right) \text{x} \left[\sum (Y\,pred - Y\,exp)^2\right]^{1/2}$$

Where,

V= number of independent variables

n = number of data points

n-v-1 = degree of freedom in multiple regression

(Y pred – Yexp) = true residual

A small value of standard error of estimate indicates the small deviation.

Chapter 3

Calculations

In calculations daily gas production data were used supplied by the GTCL and Condensate production data at Demra city gate station were collected from Titas Gas Transmission and Distribution Company Limited (TGTDCL). As A-B-D pipeline is the main trunk line, numerous intakes and off takes are present in the system. In calculations pipeline network was simplified with considering all intakes and off takes into few major points. Effects of valves, strainer and geometry like elbow, undulation etc were ignored. As A-B-D pipeline operates at high pressure, properties of gas at in-situ condition were determined at the operating pressure and temperature. After that equilibrium constant (K) was determined and with this equilibrium constant value vapor to feed ratio was determined.

3.1 Physical Properties Calculations

The physical properties are the important parameters for different calculations. For compressible fluids physical parameters are determined at operating pressure and temperature. Different properties [1] are given below: -

3.1.1 Gas Solubility / Solution Gas Ratio

Solution gas ratio is the function of the pressure, temperature, specific gravity of oil and gas. Al-Marhoun et al. presented the correlations as follows:

$$R_s = (a \gamma_g b \gamma_o c T^d p)^e$$

Where,

γ_g = Gas specific gravity,
γ_o= Condensate specific gravity
T =temperature, °R ,
P = pressure, psia
a= 185.843208,
b=1.87784
c= - 3.1437,
d = -1.32657
e = 1.398441

So, $R_s = 11.417$

3.1.2. Condensate and Gas Viscosity

The viscosity of the crude oil with dissolved gas is an important parameter in the pressure-loss calculation in pipes. Condensate viscosity of dead oil is calculated by empirical correlations. Beal's correlation for viscosity is used here.

At first condensate API gravity was calculated from condensate specific gravity. Condensate Specific Gravity, $\gamma_g = 0.793$ and

Condensate gravity, °API $= \{\frac{141.5}{0.793} - 131.5\} = 46.9363$

From Beal Correlation,

$$\mu_{od} = \{0.32 + (\frac{16000000}{\gamma API^{4.53}})\}(\frac{360}{T-260})^{a}$$

Where $a = 10^{(0.43 + \frac{8.33}{\gamma API})}$

So, dead oil viscosity = 3.257415605

Saturated Condensate viscosity was calculated from Beggs and Robinson Correlation.

μ_o is found by,

$\mu_o = [10.715(Rs + 100)^{-0.515}]\, \mu_{od}^{\ b}$

Where, $b = 5.44(Rs + 150)^{-0.338}$

$\mu_o = 1.810101435$

Gas viscosity was calculated by Lee et al. correlations

$$\mu_g = 10^{-4} K \exp\left[X\left(\frac{\rho_g}{62.4}\right)^{Y}\right]$$

$$\text{Where } K = \frac{(9.4 + 0.02M_g)T^{1.5}}{(209 + 19M_gT)};$$

$$X = 3.5 + (986/T) + 0.01M_g$$

$$Y = 2.4 - 0.2X$$

ρ_g is the gas density at the system pressure and temperature, lbm/ft³; T= Reservoir temperature, °R and Mg = apparent gas molecular weight of the gas mixture.

3.1.3. Gas density

Gas density in lbm/ft³ can be determined easily by combining the real gas law with the definition of specific gravity.

$$\rho_g = 2.7\,\gamma_g\,\frac{p}{zT}$$

$$=8.473043181\ \text{lb/ft}^3$$

$$= 135.847104\ \text{kg/m}^3$$

3.1.4. Gas deviation factor

Natural gas is a real gas and it deviates from the ideal gas. It is an important parameter for calculating the gas density and gas formation volume factor. Standing correlations to determine the pseudo-critical pressure and temperature were used to calculate gas deviation factor. For gas condensate system

$$T_{pc} = 187 + 330\gamma_g - 71.5\gamma_g^{\ 2}$$

$$P_{pc} = 706 - 51.7\gamma_g - 11.1\gamma_g^{\ 2}$$

3.2 Flash Calculations

Basis for flash calculation is overall and component molal balances. Suppose F is the number of moles in the feed, L is the number of moles in liquid phase and V is the number of moles in the vapor phase

$$F = L + V$$

$$\frac{L}{F} = 1 - \frac{V}{F}$$

For ith component,

$$z_i F = x_i L + y_i V$$

For liquid phase

$$x_i = \frac{z_i F - y_i V}{L} = \frac{z_i - x_i K_i \frac{V}{F}}{1 - \frac{V}{F}}$$

$$x_i = \frac{z_i}{1 + \frac{V}{F}(K_i - 1)}$$

For vapor phase

$$y_i = \frac{z_i K_i}{1 + \frac{V}{F}(K_i - 1)}$$

From the definition of the mole fractions

$$\sum_{i=1}^{n} x_i = \sum_{i=1}^{n} y_i = 1$$

$$\sum_{i=1}^{n} (y_i - x_i) = 0$$

$$\sum_{i=1}^{n} (y_i - x_i) = \sum_{i=1}^{n} \frac{z_i (K_i - 1)}{(K_i - 1)\frac{V}{F} + 1} = f\left(\frac{V}{F}\right) = 0$$

Equilibrium constant K values depend on the phase and composition. Any convergence solution is applicable to find the solution [1]. More details are given in Appendix A.
For Ashuganj section the V/F value is 0.95 and other values are given in result section.

3.3 Heat Balance

Whenever gas follows through a pipeline, it looses pressure due to frictional and elevation effect with pipe wall. This loss in pressure causes temperature drop, which will initiate condensation of heavier molecules. As the pipelines are buried to ground, they exchanges heat with the surroundings. But the gas flows in a very high velocity which gives a little retention time to reach equilibrium.

3.3.1 Condensate Calculation by Heat Balance:

As gas pipeline is not totally insulated and buried underground, heat is exchanging from system to surroundings. Any temperature drop due to pressure drop and other operational reasons are compensated by the surroundings. As it is an instantaneous process, measurement of local parameters are difficult and the rate of heat transfer from surroundings to the pipe wall was not measured or unknown. With high velocity of gas, theoretically equilibrium value of the temperature can be used to calculate the total heat transferred. As the temperature drops, some condensable heavier molecules condense which liberate heat of condensation until the system becomes equilibrium. Therefore temperature of the gas in pipeline increased slightly. Heat calculation is given below for Segment 1 (AGMS-Bangura): -

Total heat taken by the gas stream $Q = n \sum C_p \Delta T$, (Date 01.02.08)

Heat given by condensable components

Q = Number of moles x Heat of Condensation

$= \sum nH$

$= -39.5$ Btu/lb

$\Delta T = (12.3374-12.8659)\ ^{\circ}C = -0.5285\ ^{\circ}C = 459.4715\ ^{\circ}F$

$\sum Cp = 3.2762$ Btu/lb.F

So, $Cp\,\Delta T = 1505.32053$ Btu/lb

And, n = 0.02623803 lb mole/ lb mole

Basis = 100 lb mole

n= 2.623803 lb mole

Molecular Weight, M = 110

Mass, m = (100 x 2.623803) lb

= 262.3803 lb

Density = 50 lb/ ft^3

Volume, $V = \frac{m}{\rho} = 5.772366557\ ft^3 = 163.455$Lt

Similarly, for other 2 (two) segments, amount of condensate is found. Total amount is the sum of three. It is 659.0175 Lt [5].

3.4 Flow Pattern Calculation

Once the amount of condensate is figured out, it is essential to know the flow pattern of the fluid. Flow patterns are grouped on the basis of flow i.e. horizontal or vertical. As the pipeline is almost horizontal with little undulations due to river crossing or terrain change, flow patterns for horizontal flow are checked. In flow pattern calculations, it starts from the stratified flow and several flow criteria and boundary conditions are checked to find accurate flow pattern. The flow pattern calculation is given below: -

From flash calculation for segment 1(AGMS-Bangura), V/F is 0.95

Diameter = 28.874 in = 0.7334 m Pressure= 819 psig=833.7 psia

Temperature=14.8184 °C=287.8184 °R

Flow = 321.3 MMSCFD = 55.7160176 m^3/s

In-situ Flow=0.98177884 m^3/s

In-situ Liquid Flow, $q_L = (0.01 \times 0.98177884)\ m^3/s = 0.0098177884\ m^3/s$

In-situ Gas Flow, q_G = (0.98177884-.0098177884) m^3/s = 0.97196105 m^3/s

Liquid Density, ρ_L = 793 kg/m^3 Gas Density, ρ_G =131.34 kg/m^3

Liquid viscosity, μ_L = 1.754 cp Gas viscosity, μ_G = 0.016 cp

1. Calculate superficial velocities and superficial Reynolds numbers:

   $A_P = \frac{\pi}{4} d^2 = 0.422447\ m^2$ $V_{SL} = \frac{q_L}{A_P} = 0.02324$ m/s

   $V_{SG} = \frac{q_G}{A_P} = 2.300788$ m/s $Re_{SL} = \frac{DV\rho}{\mu} = 7.705941$ (laminar flow)

   $Re_{SG} = 13851.42$ (turbulent flow)

2. Calculate the Lockhart- Martinelli parameter and the equilibrium liquid level in the pipe:

   $X^2 = 0.002758$ And, $X = 0.052513$ and for horizontal pipe Y=0.

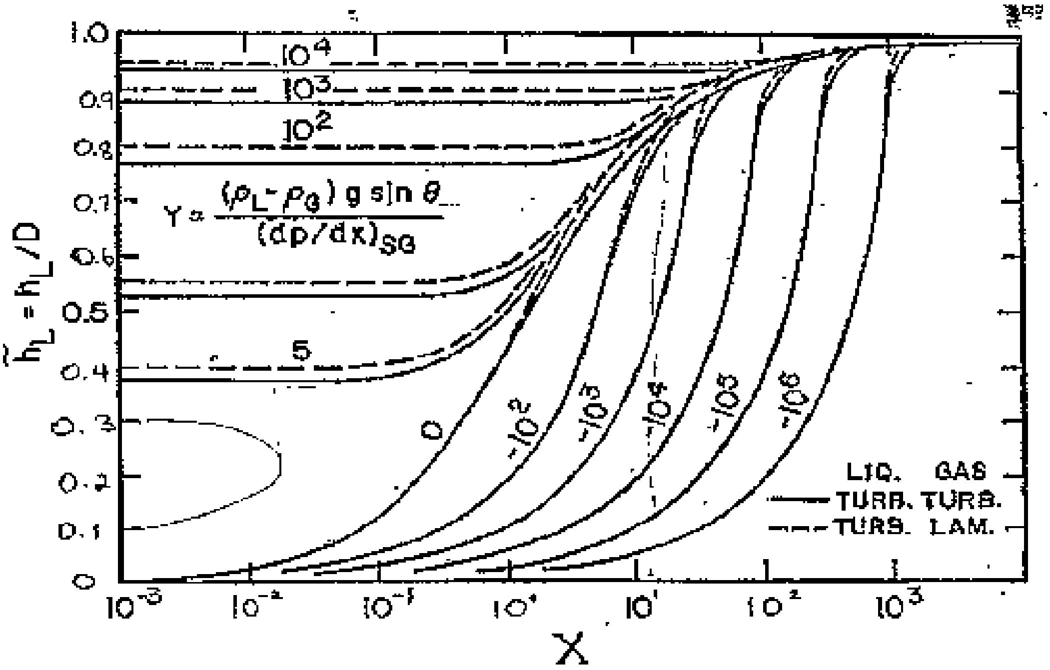

Figure 3.1: Equilibrium Liquid Level in Stratified Flow.

From Figure 3.1. dimensionless liquid hold up factor is $\tilde{h}_L = 0.05$, and liquid holdup factor is $h_L = 0.03667$.

3. Calculate dimensionless (tilde) variables:

   $\cos^{-1}(2\tilde{h}_L - 1) = 2.690566$ $\tilde{S}_I = \sqrt{(1-(2\tilde{h}_L - 1))} = 0.43589$

   $\tilde{S}_G = \cos^{-1}(2\tilde{h}_L - 1) = 2.690566$ $\tilde{S}_L = \Pi - \tilde{S}_G = 0.451034$

   $\tilde{A}_G = 0.770717$ $\tilde{A}_L = 0.014683$

   $\tilde{V}_L = 53.48929$ $\tilde{V}_G = 1.019052$

   $\tilde{D}_L = 0.130219$ $\tilde{D}_G = 0.986058$

4. Check stratified to non-stratified transition criterion:

The criteria for stratified to non-stratified flow is

$F^2 [\frac{(\tilde{V}_G^2 x \tilde{S}_i)}{(1-2\tilde{h}_L)^2 x \tilde{A}_G}] > 1$

$F = \sqrt{(\frac{\rho G}{\rho L - \rho G})} x (\frac{vSG}{\sqrt{(d g \cos\theta)}})$

$= 0.382167$

$0.095045773 < 1$

So, flow is stable and stratified flow exists.

5. Check for stratified smooth to stratified-wavy transition criterion:

The criteria for stratified smooth to stratified wavy is

$K \geq 2/[(\sqrt{(\tilde{V}_L x s)}) x \tilde{V}_G]$

Where s = 0.01 and

$K = F x \sqrt{(Re_{LS})} = 1.060879$

$2/[(\sqrt{(\tilde{V}_L x s)}) x \tilde{V}_G] = 2.683493$

$K \geq 2.683493$

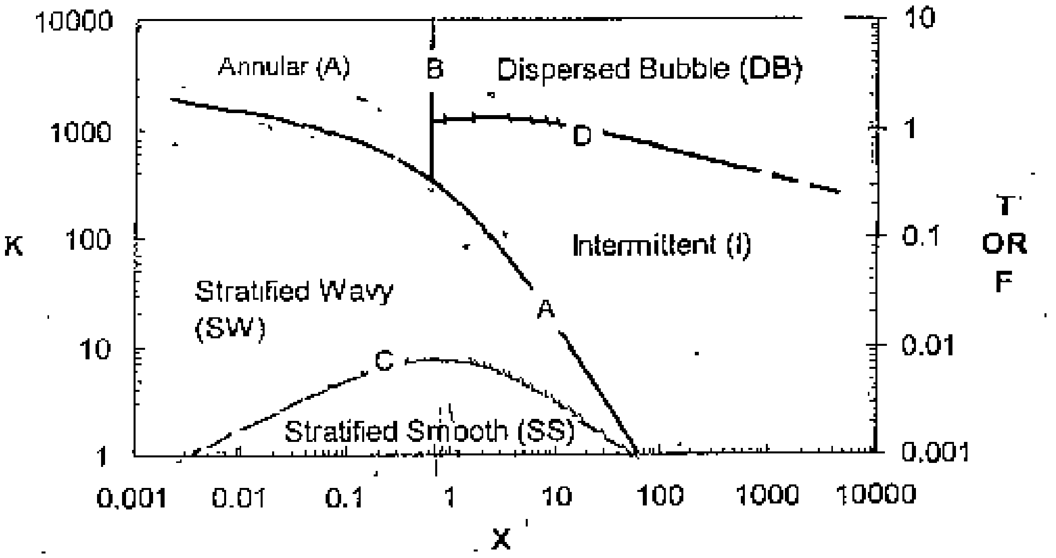

Figure 3.2: Generalized Flow Pattern Map for Horizontal Flow.

From Figure 3.2, it is clear that the flow is stratified smooth.

3.5 Pressure Calculation:

The frictional pressure gradient in the Two-phase flow strongly depends on fluid flow pattern. In literature there are numbers of correlations and graphs developed with air water system for 3 to 6 inch pipes by two-phase flow researchers. Though the gas-condensate system is different from air-water system, same flow pattern maps were used to identify the flow pattern. As the flow pattern for the gas-condensate system always remains in the stratified region and condensate amount is very low compared with pipe size, a single phase pressure correlations are used here for pressure/pressure drop calculations. But from the two phase flow side marching algorithm techniques used with the help of Two-phase flow parameters. In this case, the pipeline is divided into several of sections and an iterative procedure is used to find the solution. In this thesis, results from simple excel application are used to compute the pressure data. Calculation steps are shown for the marching algorithm below:-

3.5.1 Marching Algorithm Method (Iteration) for segment 1 (AGMS-Bangura):

Gas flow rate (from daily report) = 321.3 MMSCFD=105.3032733 ft^3/s

Liquid flow rate = (0.01 x 105.3031733) ft^3/s = 1.053031733 ft^3/s

Gas density (by calculation) =127.8384833 kg/ m^3

Liquid density = 793 kg/ m^3

P in = 777 psig = 791.7 psia

L= 40 km = 131233.6 ft

Diameter = 28.874 in = 0.7334 ft

$\Delta L = \frac{L}{n} = \frac{131233.6}{1}$ = 131233.6 ft

Pgrad = 0.0002 psia / ft (given)

Avg. T = 514.68 °R

$P_{i+1\,(G)} = \{P\text{ in} - P.\text{grad} \times \Delta L\}$ = 529.2328084 psia

Avg. P. = $\frac{P in + P_{i+1\,(G)}}{2}$ = 660.4664042 psia

Reynold's Number and all physical properties are determined in this avg. T. and P.

Re_{SL} = 12.49673 (laminar flow, $c_L=c_G=16$, $m=n=1$)

Re_{SG} = 23900.75 (turbulent flow, $c_l=c_G=0.046$, $m=n=0.2$)

$f_G = 0.046\,(Re_{SG})^{-0.2}$

$f_L = 16(Re_{SL})^{-1}$

$-\left(\frac{dp}{dl}\right)_{SG} = \frac{2}{d} f_G\, \rho_G\, v_{SG}\, 2$

=33.940525 Psia/ft

$-(\frac{dp}{dl})_{SL} = \frac{2}{d} f_l \; \rho_L \, v_{SL}\, 2$

$= 4.49063694$

$X = \sqrt{[\; \frac{-(dp/dl)_{SG}}{-(dp/dl)_{SL}} \;]} = 0.363743$

$H_L = 1 - (1 + X^{0.8})^{-0.378}$

$\Phi_L^{\,2} = (1 \div \frac{C}{X} + \frac{1}{X^2})$, C is a constant which depends on flow.

$= 41.54839$

$(\frac{-dp}{dl})_F = \Phi_L^{\,2} \;x\; (-\frac{dP}{dL})_{SL}$

$= 186.5787196$ Psia/ft

$-(\frac{dp}{dl})_A = G \;x\; (X) \;x\; (\frac{dV_g}{dP}) = -\,0.0004936 \frac{dP}{dL}$

Where,

$G = \frac{WL + WG}{Ap}$

$X = \frac{WG}{WL + WG}$

$\frac{dV_g}{dP} = \{ \frac{-16.74}{8314 x 298 x \rho G x \rho G} \}$

$\frac{-dP}{dL} = 186.5787196 - 0.0004936 \frac{dP}{dL}$

Or, $\frac{-dP}{dL} = 186.48667$ psia/ft

$P_{i+1\,(C)} = Pi - (\frac{-dP}{dL})_i \, dL_i$

$= 24472525$ psia

Compare,

$\frac{\{P_{i+1}(G) - P_{i+1}(C)\}}{P_{i+1}(C)} \le \epsilon$

By iteration process, final P = 578.0518 Psia. Convergence criteria was met if two final values were same number or within convergence criteria.

3.5.2 Weymouth Equation for segment 1(AGMS-Bangura):

Gas Flow Rate, Q = 170 MMSCFD =1967.6 ft^3/s (Date: 03.02.08)

P_1 = 819 psig = 833.7 psia

Diameter, d = 30 inch = 2.54 ft

$\gamma g = 0.575595$

$Z_{av} = 0.89$

Temperature, T = 14.9039 °C = 518.827 R

$P_{sc} = 14.7$ psia

Length, L = 40 km = 131233.6 ft

$T_{sc} = 520$ R

Area, A = 4.90875 ft^2

Velocity, $v = \frac{Q}{A} = 400.833$ ft/s

$Re = (dv\rho)/\mu = 5416032.658$ (smooth pipe)

Friction Factor, f = 0.0128 (from curve)

So, $(P_1^2 - P_2^2) = 13369.61127$

Or, $P_2^2 = 681686.08$

Or, $P_2 = 825.64$ psia (measured value is 854.7 psia).

3.5.3 Panhandle A Equation for segment 1(AGMS-Bangura):

Gas Flow Rate, Q = 294.95 MMSCFD =3413.74 ft^3/s (Date: 01.02.08)

$P_1 = 777$ psig = 791.7 psia

Diameter, d = 30 inch = 2.54 ft

$\gamma g = 0.575595$

$Z_{av} = 0.88$

Temperature, T = 14.9039 °C = 518.827 R

$P_{sc} = 14.7$ psia

Length, L = 40 km = 131233.6 ft

$T_{sc} = 520$ R

$\mu_g = 0.0110125$ cp

So, $P_2 = 717.1473$ psia (measured value is 784.7 psia).

3.5.4 Panhandle B Equation for segment 1(AGMS-Bangura):

Gas Flow Rate, Q = 294.95 MMSCFD =3413.74 ft^3/s (Date: 01.02.08)

$P_1 = 777$ psig = 791.7 psia

Diameter, d = 30 inch = 2.54 ft

$\gamma g = 0.575595$

$Z_{av} = 0.88$

Temperature, T = 14.9039 °C = 518.827 R

$P_{sc} = 14.7$ psia

Length, L = 40 km = 131233.6 ft

$T_{sc} = 520$ R

$\mu_g = 0.0110125$ cp

Area, A = 4.90875 ft^2

Velocity, $v = \frac{Q}{A} = 757.575$ ft/s

$Re = \frac{dv\rho}{\mu} = 10236301.72$ (smooth pipe)

So, $P_2 = 789.473$ psia (measured value is 784.7 psia).

3.6 Error Calculation:

Average Percentage Relative Error (APRE), Absolute Average Percentage Relative Error (AAPRE) and Standard Error of Estimate (SEE) were calculated for condensate calculation and pressure calculation.

All these values are calculated for Segment 1 (Ashuganj- Bangura) by using Panhandle A equation at 01.02.08.

$$APRE = (\frac{1}{n}) \times \sum | \frac{\{(Ypred - Y exp) x100\}}{Y exp}]$$

$$= 13.49 \%$$

$$AAPRE = (\frac{1}{n}) \times \sum | [\{\frac{(Ypred - Y exp) x100}{Y exp}\}] |$$

$$= 17.64 \%$$

$$SEE = \sigma = (\frac{1}{n-r-1}) \times [\sum (Y pred - Y exp)^2]^{1/2}$$

$$= 52.26$$

Chapter 4
Results and Discussion

4.1 Physical Properties Calculation:

In Two Phase flow, physical properties at different flow conditions are very important parameters for flow pattern and pressure drop calculations. As the pressure varies along the length of the pipeline, physical properties also change. Here condensate and gas physical properties are given in the Table 4.1 for segment 1 (AGMS-Bangura) at operating pressure and temperature conditions.

Table 4.1: Physical properties for segment 1(AGMS-Bangura).

Date	Gas Solubility, (SCF/bbl)	Oil Viscosity, (cp)	Gas Density, (Kg/m^3)	Gas Viscosity, (cp)
01.02.08	102.343	1.874	121.84	0.1564
02.02.08	105.9117	1.754	131.34	0.01593
03.02.08	109.6922	1.803	134.4066	0.01607
05.02.08	104.9982	1.874	130.1358	0.01577
06.02.08	105.209	1.9364	130.1172	0.01573
07.02.08	104.9385	1.92	129.936	0.01573
08.02.08	102.6985	1.933	127.957	0.0156
09.02.08	105.7041	1.775	131.083	0.0124
10.02.08	104.4752	1.8922	128.515	0.0158
11.02.08	103.7461	1.8975	127.873	0.011886
12.02.08	103.5722	1.87485	128.904	0.0157
13.02.08	105.0785	1.879	130.189	0.015772
14.02.08	106.03108	1.87935	131.25	0.015834
15.02.08	106.3198	1.856143	131.3357	0.0158564
16.02.08	112.5922	1.7283	137.148	0.016302
17.02.08	106.9364	1.84443	131.905	0.015898
18.02.08	107.3815	1.8347	132.32	0.01323
19.02.08	106.181	1.85999	131.2038	0.01584
20.02.08	106.645	1.81658	131.7511	0.0159098
21.02.08	104.692	1.8507	129.931	0.01571

Fluid physical properties for segment 2 and Segment 3 are given in the Appendix B.

4.2 Condensate Calculations:

Condensate volume was calculated in this thesis by two methods i.e. by flash calculation and heat balance method. Condensate volume was calculated for the entire pipe by these two methods and calculated values were compared with the measured values for the segment 3.

4.2.1 Flash Method:

Flash calculation predicts how many moles are in gaseous phase of total moles in feed. The number of moles of liquid is converted to the liquid volume by multiplying with liquid specific molar volume. With the help of equations as stated earlier, flash calculation was performed to different sections of the pipe and results are given below Table 4.2.

Table 4.2 Flash Calculation (V/F) results.

Name of segment	V/F
Ashuganj	0.95
BKB	0.992
Demra	0.99

Once the ratio of vapor phase moles to total moles (V/F) is known then liquid moles are calculated and liquid volume is determined.

4.2.2 Heat Balance Method:

In heat balance method, the physics lies in condensation as follows. When pressure in gas pipeline drops due to friction or elevation the temperature also drops. If temperature and pressure drops, the equilibrium also changes to a new point. If equilibrium changes the vapor and liquid volume also changes. In ideal case, condensation process liberates heat itself and gas stream takes the heat and reaches to equilibrium. As the underground gas pipe lines are not fully insulted it exchange heat with the surroundings. In Table 4.3, calculated values of condensate by flash method and heat balance method were given for segment 1 and segment 2. As these two segments did not have any measured condensate vales, comparison between measured and calculated values could not be made. For segment 3 (Bakhrabad- Demra) condensate data were collected from the Titas Gas Transmission and Distribution Company Ltd.

Table 4.3: Calculated condensate values by flash and heat balance method.

Date	Gas Flow Rate (MMSCFD)		Flash Method (Liter/day)		Heat Balance (Liter/day)	
	Segment 1	Segment 2	Segment 1	Segment 2	Segment 1	Segment 2
01.02.08	317.05	67.19	8879.473	2025.670	134.56	248.35
03.02.08	170	66.43	4528.232	1722.935	134.157	229.814
05.02.08	295.8	67.72	7927.362	5389.867	163.517	245.929
06.02.08	323	97.34	8299.085	2675.675	134.564	554.457
07.02.08	317.05	67.19	8259.186	1695.298	133.236	249.32
08.02.08	316.2	69.64	8382.438	1942.595	137.643	275.878
09.02.08	277.1	70.4	7845.587	880.126	131.745	226.03
10.02.08	309.4	70.63	8248.696	2878.317	139.155	261.06
11.02.08	321.3	82.23	8608.961	2350.153	139.155	261.06
12.02.08	306	70.3	8334.460	1539.494	135.785	249.108
13.02.08	317.9	70.24	8486.160	2286.696	136.128	256.364
14.02.08	317.05	70.22	8345.820	1927.389	137.6	265.727
15.02.08	307.7	70.19	8220.761	1707.233	136.108	254.2
17.02.08	283.05	70.24	7214.176	2381.697	135.378	245.456
18.02.08	290.7	70.24	7566.816	2169.434	135.095	236.472
19.02.08	298.35	70.23	7781.248	2147.093	136.635	250.325
20.02.08	295.8	70.17	7963.638	1893.837	133.057	232.9
21.02.08	297.5	70.16	8076.363	2023.374	137.5	246.19

Comparison of the calculated condensate values were made with the measured values. All these values are given in Table 4.4.

Table 4.4: Comparison of calculated condensate values for Segment 3 (BKB-Demra).

Date	Gas Flow Rate (MMSCFD)	Measured (Liter/day)	Flash Method (Liter/day)	Heat Balance Method (Liter/day)
01.02.08	210.031	800	1362.529	554.457
03.02.08	14.72	1000	85.38	562.533
05.02.08	171.525	1400	1159.874	559.278
06.02.08	204.367	600	1336.343	554.457
07.02.08	223.091	1400	1426.297	608.2
08.02.08	208.314	2000	1282.264	547.809
09.02.08	178.845	1800	1248.757	565.11
10.02.08	211.86	1200	1356.831	551.7
11.02.08	222.005	1200	1450.597	551.7
12.02.08	177.502	1200	1184.399	557.416
13.02.08	189.595	1400	1224.985	553.08
14.02.08	201.237	1200	1278.065	551.211
15.02.08	201.7	2000	1307.681	554.78
17.02.08	169.953	2000	1114.556	593.067
18.02.08	157.045	2000	1071.447	560.034
19.02.08	165.559	2000	1090.956	605.174
20.02.08	167.084	1500	1140.162	560.876
21.02.08	198.811	2400	1318.668	596.5

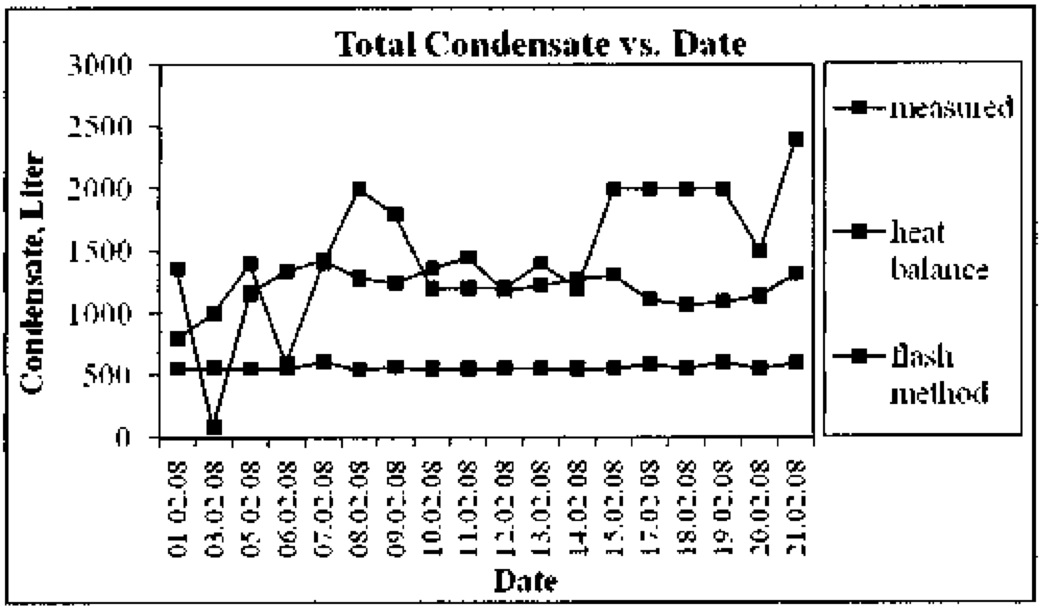

Figure 4.1: Total Condensate (liter) vs. Date for Segment 3 (BKB-Demra).

Measured values and calculated condensate values are shown in Figure 4.1. From the figure 4.1, condensate values from flash calculations are changing from 1000 to 1400 liters where as measured condensate values vary from 800 to 2300 liters. In heat balance method, calculated values of condensate is almost 300 to 400 liters and values are almost constant i.e no variation with the gas flow rates. Average Percentage Relative Error (APRE), Absolute Percentage Relative Error (AAPRE) and Standard Error of Estimate (SEE) were calculated for flash and heat balance method and presented in the Table 4.5.

Table 4.5: Condensate error calculation for Segment 3 (BKB-Demra).

Date	Condensate, Liter			APRE		AAPRE	
	Measured	Heat Balance	Flash method	Heat balance	Flash. method	Heat balance	Flash method
01.02.08	800	554.457	1362.529	-30.693	70.3162	30.6929	70.3162
03.02.08	1000	562.533	85.38	-43.747	-91.462	43.7467	91.462
05.02.08	1400	559.278	1159.874	-60.052	-17.152	60.0516	17.1518
06.02.08	600	554.457	1336.34	-7.5905	122.724	7.5905	122.724
07.02.08	1400	608.2	1426.297	-56.557	1.87837	56.5571	1.87837
08.02.08	2000	547.809	1282.263	-72.61	-35.887	72.6096	35.8868
09.02.08	1800	565.11	1248.756	-68.605	-30.625	68.605	30.6246
10.02.08	1200	551.7	1356.831	-54.025	13.0693	54.025	13.0693
11.02.08	1200	551.7	1450.596	-54.025	20.8831	54.025	20.8831
12.02.08	1200	557.416	1184.398	-53.549	-1.3001	53.5487	1.30009
13.02.08	1400	553.08	1224.985	-60.494	-12.501	60.4943	12.5011
14.02.08	1200	551.211	1278.065	-54.066	6.50543	54.0658	6.50543
15.02.08	2000	554.78	1307.681	-72.261	-34.616	72.261	34.6159
17.02.08	2000	593.067	1114.555	-70.347	-44.272	70.3467	44.2722
18.02.08	2000	560.034	1071.446	-71.998	-46.428	71.9983	46.4277
19.02.08	2000	605.174	1090.956	-69.741	-45.452	69.7413	45.4522
20.02.08	1500	560.876	1140.162	-62.608	-23.989	62.6083	23.9892
21.02.08	2400	596.5	1318.668	-75.146	-45.055	75.1458	45.0555
				APRE		AAPRE	
				-57.673	-10.74	57.673	36.895
						SEE	
						164.949	97.623

The AAPRE and SEE values for heat balance and flash methods were 57.673% & 164.949 and 36.895 % & 92.623 respectively. APRE values by flash and heat balance methods were presented in the Fig-4.2. From the statistical parameter analysis, it was found that flash method predicted the condensate values more closely than heat balance methods for

Segment-3. But this comparison was not done for Segments -1 & 2 due to lack of the measured condensate data.

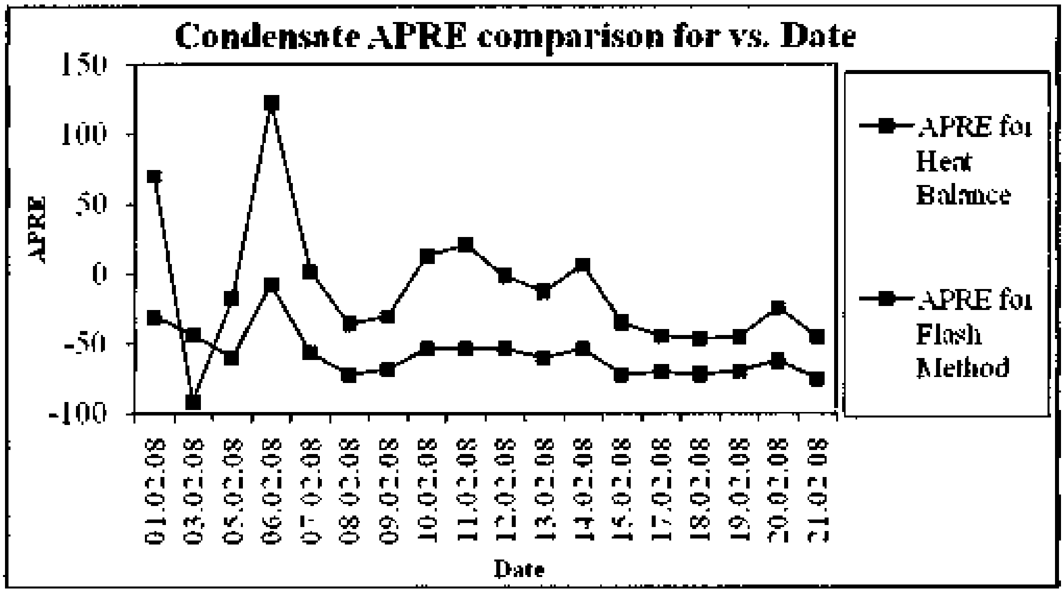

Figure 4.2: APRE comparison for condensate (Calculated by Heat Balance and Flash Method) vs. Date.

Trend analysis defines the change in output with the change of any input parameter. Here Trend analysis was done for three values i.e. measured values, flash method and heat balance methods with time. All calculated trend analysis values were presented in the Table 4.6. The graph for trend analysis was given in the Figure 4.3. From the Fig 4.3, trend analysis of measured values gave an almost sinusoidal curve with high amplitudes. The flash method trend also illustrated the same sinusoidal wave like shape but with low amplitude.

Table 4.6: Condensate Trend Analysis for Segment 3 (BKB-Demra).

Date	Measured Liter	Heat balance Liter	Flash Method Liter	Δ Measured/ Δ day	Δ Heat balance/ Aday	Δ Flash method/ Δ day
01.02.08	800	554.457	1362.53	0	0	0
03.02.08	1000	562.533	85.38	200	8.076	-1277.1
05.02.08	1400	559.278	1160	400	-3.255	1074.494
06.02.08	600	554.457	1336.4	-800	-4.821	176.4692
07.02.08	1400	608.2	1426.3	800	53.743	89.95367
08.02.08	2000	547.809	1282.26	600	-60.391	-144.033
09.02.08	1800	565.11	1248.75	-200	17.301	-33.5071
10.02.08	1200	551.7	1356.83	-600	-13.41	108.0747
11.02.08	1200	551.7	1450.6	0	0	93.76552

Date	Measured Liter	Heat balance Liter	Flash Method Liter	Δ Measured/ Δ day	Δ Heat balance/ Δday	Δ Flash method/ Δ day
12.02.08	1200	557.416	1184.4	0	5.716	-266.198
13.02.08	1400	553.08	1225	200	-4.336	40.58631
14.02.08	1200	551.211	1278	-200	-1.869	53.07992
15.02.08	2000	554.78	1307.7	800	3.569	29.61631
17.02.08	2000	593.067	1114.5	0	38.287	-193.126
18.02.08	2000	560.034	1071.5	0	-33.033	-43.1091
19.02.08	2000	605.174	1091	0	45.14	19.50951
20.02.08	1500	560.876	1140.1	-500	-44.298	49.20598
21.02.08	2400	596.5	1318.7	900	35.624	178.5058

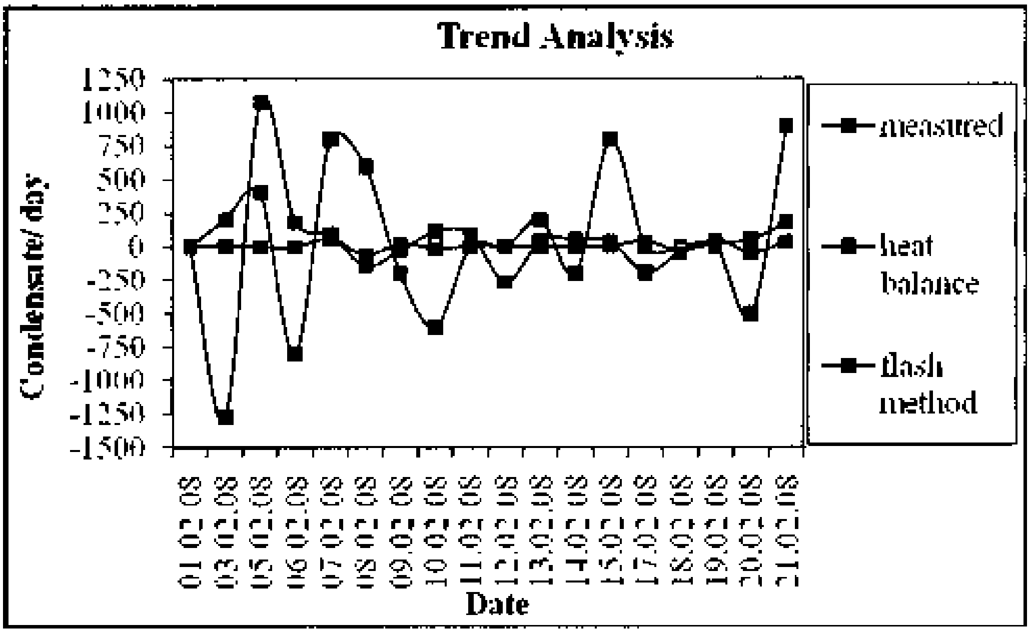

Figure 4.3: Condensate Trend Analysis for Segment 3 (BKB-Demra).

From the trend analysis it can be seen that flash method is trying to flow the actual trend in somewhat extent. It shows that the flash method gave better result than heat balance method. Sinusoidal wave in the measured trend suggested two things- firstly, Change in flow conditions and secondly the pipe was no longer perfectly horizontal rather it had some undulations. In flash method, trend analysis's amplitudes of the wave are very small, this happens due to the change in the flow only because undulation of the pipe cannot be considered in the flash calculation.

4.3 Flow Pattern Calculation:

At first, superficial velocities and superficial Reynolds numbers are calculated. Then, the Lockhart- Martinelli parameter (X) was calculated. The equilibrium liquid level h_L, in the pipe was found from the figure 3.1. Once, the equilibrium liquid level, h_L was found, then other parameters like, K or F was calculated. With the parameters K and F and Figure 3.2, flow pattern was established for the all sections of the pipeline and results were presented in Table 4.7. Detailed calculation of flow pattern prediction was given in the appendix B.

Table 4.7: Flow Pattern Results.

Section	Name	Type of Flow Pattern
1	AGMS-Bangura	Stratified-smooth
2	Bangura-BKB	Stratified-smooth
3	BKB-Demra	Stratified-smooth

As the flow pattern in all segments were Stratified- smooth, so from the assumption of the stratified flow we can say that at interface both liquid and gas phase has the same velocity. Here all flow patterns turned out as Stratified – smooth because pipe diameter for every section was very large compared to amount of the liquid generated and pipeline was assumed as perfectly horizontal one. In this report pipeline was considered horizontal because of undulation data of the pipeline were not found.

4.4 Pressure Calculation:

All pressure values were calculated by assuming single-phase fluid with Weymouth, Panhandle A, Panhandle B and Fekete Software (Piper, using Panhandle B) correlations. In pressure calculations, Reynolds's Number and all physical properties are determined in average temperature and pressure. Initial pressure was taken as the inlet pressure of pipe and then outlet pressures were calculated by above mentioned correlations. At first a guess value of outlet pressure was taken and fluid physical properties were calculated on the basis of the average value of pressures. Then iteration was done until the convergence criterion was achieved. If a convergence criterion met, final value may become more representative and that final value was reported. In some cases, if two subsequent iterations gave same number, it was also taken as final value. In Marching algorithm pipeline has to be divided into numerous segments where fluid properties do not change significantly within the segments and iteration of pressure calculations for each segments were done. The Marching algorithm technique yields better results with good computing knowledge and applications. In this thesis, Microsoft Excel application was used instead of the computer programming.

Here Piper software from Fekete Inc. also used for better results and understanding. A schematic view of ABD pipeline network by piper is given in Figure 4.4. All calculated and measured pressure values were given in the following tables. Table 4.8 shows the predicted values of pressure by different correlations and measured pressure values for the pipe segment-1.

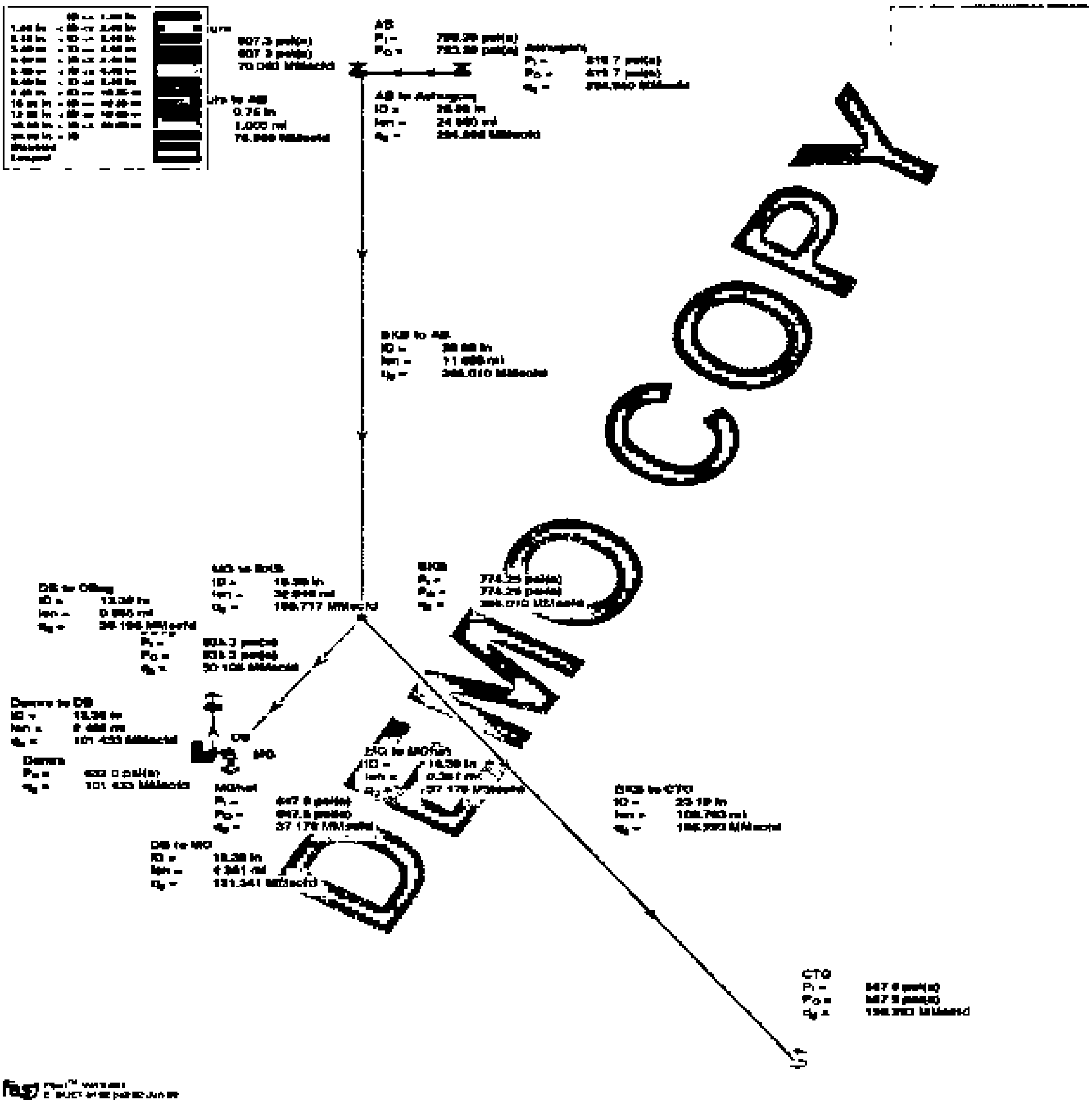

Figure 4.4: Piper Diagram of ABD Pipeline.

Table 4.8: Pressure Calculation for Segment 1 (AGMS-Bangura), diameter 30-inch.

Date	Inlet Pressure (psia)	Outlet Pressures, (Psia)					
		Measured	Panhandle A	Panhandle B	Weymouth	Iteration	Piper
01.02.08	791.7	784.7	717.147	789.473	795.05	-	773.6
03.02.08	833.7	854.7	826.96	833.01	825.64	-	807.3
05.02.08	804.7	799.37	787.65	802.83	825.64	-	790.9
06.02.08	800.7	784.7	780.487	798.471	779.14	341.00	782.5
07.02.08	800.7	789.7	781.18	798.55	758.8	582.78	782.6
08.02.08	788.7	774.7	768.975	789.53	771.11	574.06	771.6
09.02.08	816.7	834.7	801.838	815.087	758.81	583.411	807.2
10.02.08	800.7	789.7	782.055	798.65	794.65	577.96	785.1
11.02.08	796.7	774.7	776.583	794.487	772.55	596.09	779.0
12.02.08	797.7	794.7	779.367	795.69	766.14	578.05	782.6
13.02.08	804.7	789.7	785.182	802.55	770.07	586.03	787.5
14.02.08	812.7	789.7	791.427	808.58	775.10	589.95	793.6
15.02.08	810.7	799.7	794.525	810.708	781.49	589.00	797.0
17.02.08	816.7	824.7	801.235	815.018	835.32	584.76	804.9
18.02.08	819.7	809.7	803.504	817.934	793.68	601.58	807.6
19.02.08	811.7	804.7	794.5253	809.823	795.49	585.60	798.3
20.02.08	817.7	814.7	800.927	815.868	785.92	588.92	805.3
21.02.08	812.7	804.7	787.465	810.836	792.56	580.80	804.8

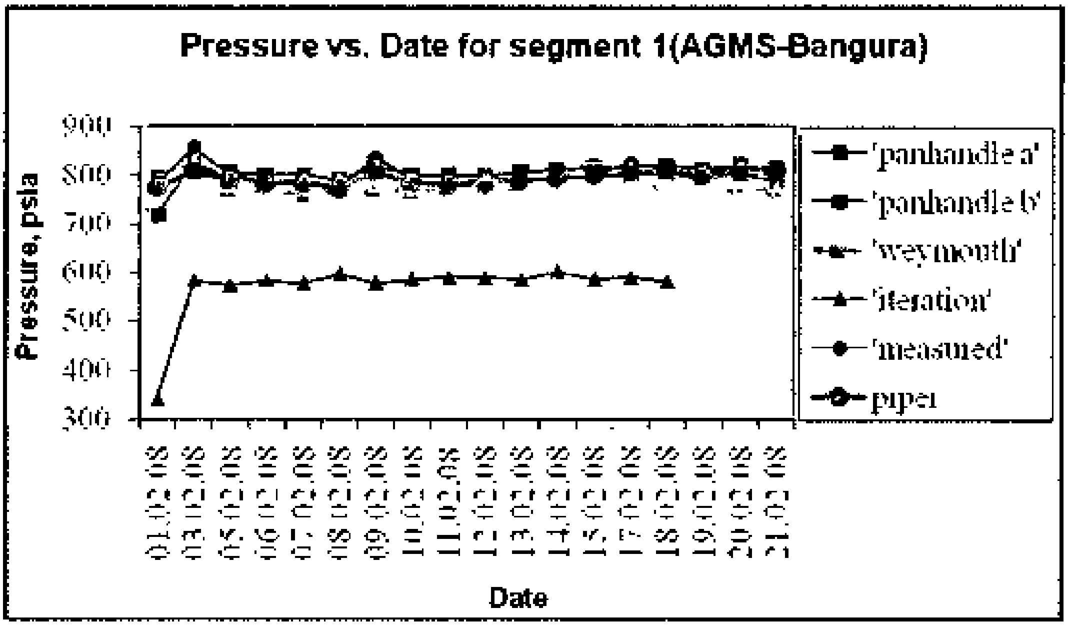

Figure 4.5 (a): Pressure Equations vs. Date for Segment 1 (AGMS-Bangura).

Figure 4.5 (a) shows all the predicted values with the measured values. As most of the predicted and measured values lies into a small range in the graph, Figure 4.5 (b) was drawn by excluding the worst predicting correlation, i.e. iteration method.

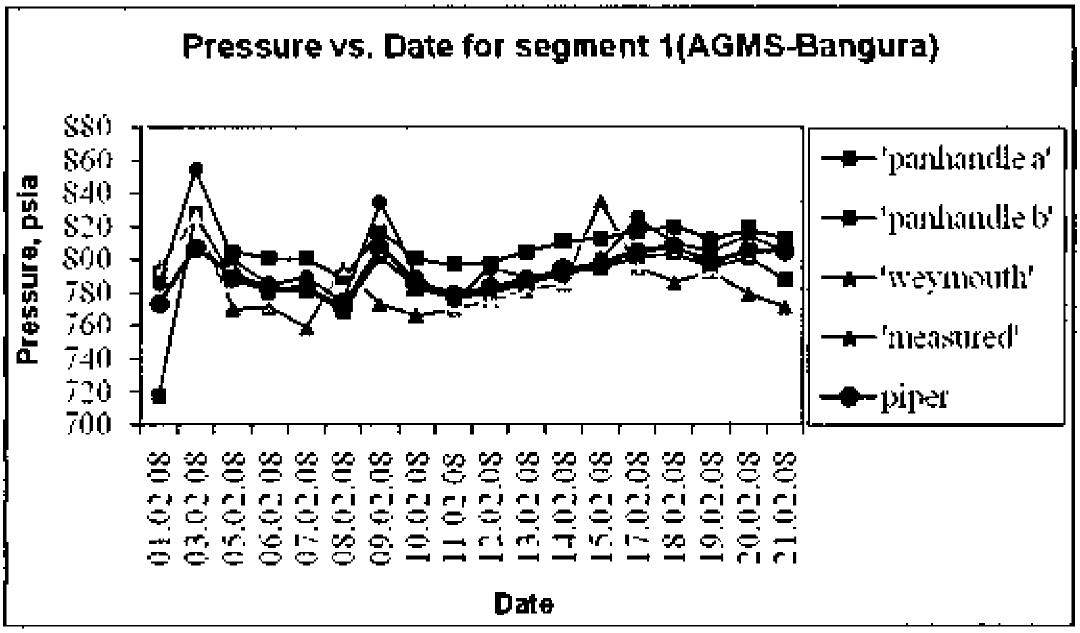

Figure 4.5 (b): Pressure Equations vs. Date for Segment 1 (AGMS-Bangura) except Iteration Method.

From figure 4.5 (b), Panhandle B and Piper gave the best pressure prediction. Next best correlations were Weymouth and Panhandle-A correlations. Panhandle B gave good results as it worked well with diameter greater than NPS (Nominal Pipe Size) 24. Panhandle A also gave good results because it worked well with Reynolds number greater than 300,000 (for example, Re = 10236301.72 for 01.02.08). Iteration method of pressure prediction was the most erroneous and unrealistic. Error analysis of different correlations for this section also supports the above statement. The Error analysis was presented in the Table 4.9.

Table 4.9 Error calculation for Segment 1 (AGMS-Bangura).

Date	Weymouth	Panhandle A	Panhandle B	Iteration	Piper
01.02.08	1.31898	-8.6088	0.60826	-100	-1.4146
03.02.08	-3.4	-3.2456	-2.5377	-100	-5.5458
05.02.08	-3.6779	-1.4662	0.43284	-100	-1.0596
06.02.08	-1.7312	-0.5369	1.75494	-56.5439	-0.2804

Date	Weymouth	Panhandle A	Panhandle B	Iteration	Piper
07.02.08	-3.9129	-1.0789	1.12068	-26.2024	-0.8991
08.02.08	2.57519	-0.739	1.91429	-25.9068	-0.4002
09.02.08	-7.4458	-3.937	-2.3497	-30.1053	-3.2946
10.02.08	-2.9834	-0.9681	1.13334	-26.813	-0.5825
11.02.08	-0.6067	0.24306	2.55415	-23.055	0.55505
12.02.08	-2.4663	-1.9294	0.12458	-27.2619	-1.5226
13.02.08	-1.0384	-0.5721	1.6272	-25.7899	-0.2786
14.02.08	-0.5572	0.21869	2.39078	-25.2935	0.49386
15.02.08	4.45417	-0.6471	1.37652	-26.3474	-0.3376
17.02.08	-3.5407	-2.8453	-1.174	-29.0942	-2.4009
18.02.08	-2.9369	-0.7652	1.01692	-25.7033	-0.2594
19.02.08	-1.5086	-1.2644	0.63663	-27.2275	-0.7953
20.02.08	-4.4016	-1.6906	0.14337	-27.7133	-1.1538
21.02.08	-4.1737	-2.1418	0.76252	-27.824	0.01243
APRE	**-1.76996**	**-1.77636**	**0.640865**	**-28.72543**	**-1.06464**
AAPRE	**2.697554**	**1.827667**	**1.314358**	**28.72543**	**1.182563**
SEE	**4.121626**	**3.353812**	**1.931591**	**33.99654**	**2.343014**

Errors from different correlations were presented in Table 4.9 and it is clear that, Panhandle A and B equation and Piper gave less error i.e - 1.776, 0.641 and -1.064%.

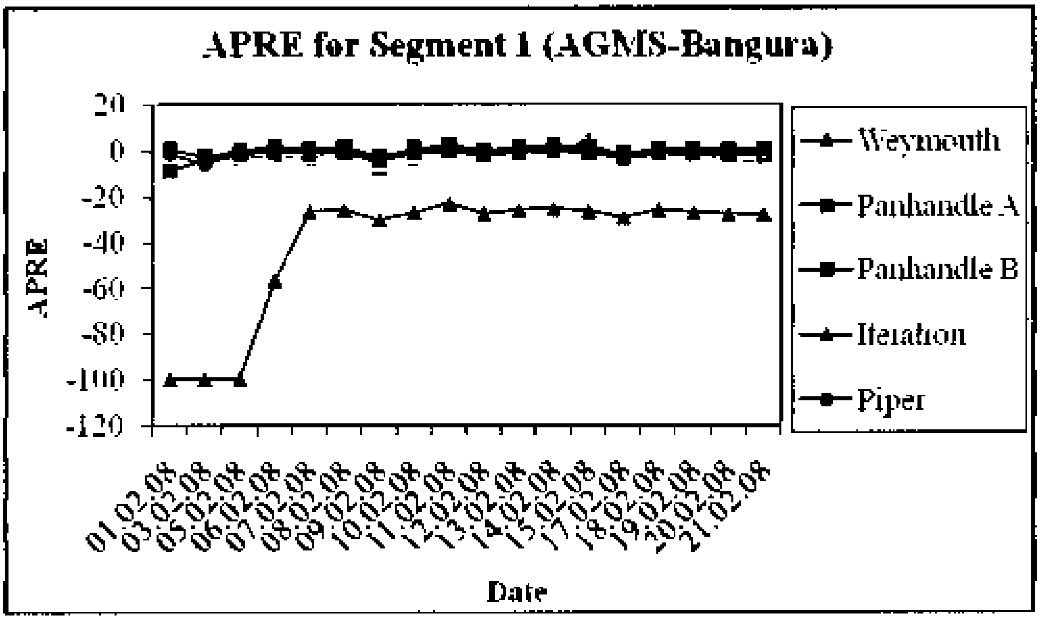

Figure 4.6: APRE for Various Pressure Equations for Segment 1(AGMS-Bangura).

From Figure 4.6, it is clear that, pressure calculated by Piper, Panhandle A, Panhandle B and Weymouth method gave good results but Panhandle B correlations gave the best result with error 0.64%. This is because of Panhandle B correlation developed for long pipeline and Nominal Pipe Size (NPS) over 24 inches. Panhandle A correlation was slightly under predicting and found the second best performing correlation for this segment with AAPRE 1.82% and SEE 3.35. Panhandle A works well where Reynolds number greater than 300,000 and partially turbulent flow occurs. In this section it gave quite satisfactory results due to Reynolds number (for example, Re = 326,266,665.8 for 01.02.08). Other best performing correlations are Piper and Weymouth with AAPRE values 1.18 % and 2.69 %and SEE values 2.34 and 4.12 respectively. Iteration method (Marching Algorithm) for two-phase flow gave unsatisfactory results because of limitation of handling the iteration method by Microsoft-Excel. A better result could be found if a computer program was used.

Similar calculation was done for Segment 2 (Bangura-BKB). Table- 4.10 shows the all predicted pressure values with measured pressure values for segment 2. Predicted pressure values with different correlations and measured values were shown in the Figure 4.7.

Table 4.10 Pressure calculation for Segment 2 (Bangura-BKB), diameter 30 inches.

Date	Inlet Pressure (psia)	Outlet Pressures, (psia)					
		Measured	Panhandle A	Panhandle B	Weymouth	Iteration	Piper
01.02.08	784.70	752.7	784.344	784.663	814.2	0	735.88
03.02.08	854.7	849.7	854.143	854.64	854.7	0	-
05.02.08	799.70	767.7	799.09	854.645	800	0	758.17
06.02.08	784.70	748.7	783.88	784.62	784.7	411.2	745.04
07.02.08	789.70	753.7	788.948	789.62	790	439.74	747.38
08.02.08	774.70	745.7	773.862	774.62	774.7	433.2	734.15
09 02.0	834 70	802.7	833.924	834.62	834.7	435	775.78
10.02.08	789.70	755.7	788.88	789.62	790	435.85	749.19
11.02.08	774.70	740.7	773.866	774.62	774.7	450.576	741.39
12.02.08	794.70	764.7	793.893	794.62	795	435.035	746.96
13.02.08	789.70	754.7	788.883	789.62	790	442.53	750.88
14.02.08	789.70	753.7	788.884	789.62	790	445.34	810.20
15.02.08	799.70	772.7	798.8952	799.622	800	443.448	762.73
17.02.08	824.70	790.7	823.9186	824.62	825	432.43	773.71
18.02.08	809.70	780.7	808.9041	809.623	810	449.93	774.98
19.02.08	804.70	775.7	803.899	804.622	805	439.78	764.38
20.02.08	814.70	785.7	813.91	814.62	815	441.815	772.07
21.02.08	804.70	773.7	803.9	804.622	805	435.08	771.3

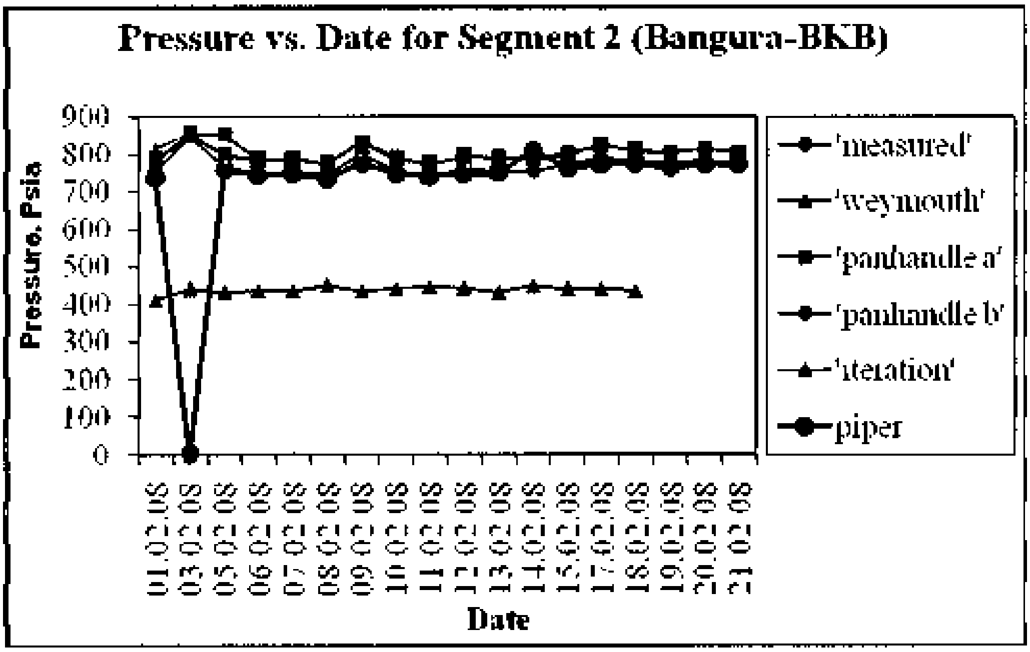

Figure 4.7 (a): Pressure Equations vs. Date for Segment 2 (Bangura-BKB).

Figure 4.7 (b) was derived from Figure 4.7 (a) by excluding the iteration method.

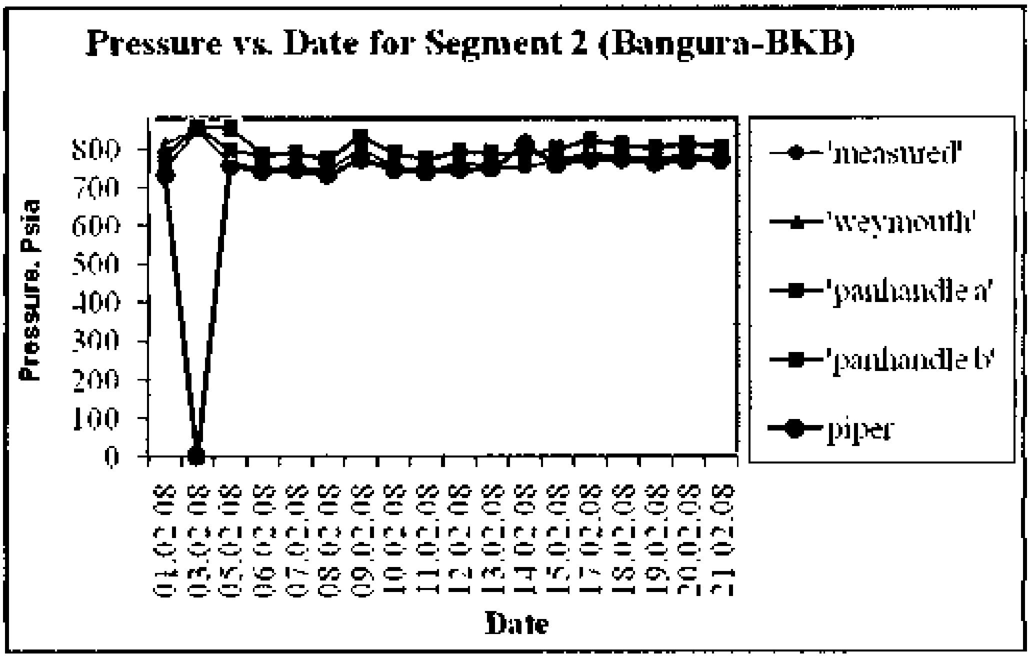

Figure 4.7 (b): Pressure Equations vs. Date for Segment 2 (Bangura-BKB) except Iteration Method.

From figure 4.7 (b) showed that piper correlation closely follows the measured values. Panhandle A, Panhandle B and Weymouth correlation followed the measured values with little over prediction.

Table 4.11 APRE comparison for Segment 2 (Bangura-BKB).

Date	Panhandle A	Panhandle B	Iteration	Weymouth	Piper
01.02.08	4.204065	4.246446	-100	-7.55343	-2.23462
03.02.08	0.5229	0.58138	-100	-0.585	-100
05.02.08	4.088837	11.32539	-100	-4.0375	-1.24137
06.02.08	4.698811	4.797649	-45.0781	-4.58774	-0.48885
07.02.08	4.676662	4.765822	-41.6558	-4.59494	-0.83853
08.02.08	3.776586	11.92437	-41.9069	-3.74338	-1.54888
09.02.08	3.889872	-1.6295	-45.8079	-3.83371	-3.35368
10.02.08	4.390631	2.503639	-42.325	-4.34177	-0.86145
11.02.08	4.477656	7.2796	-39.1689	-4.3888	0.093155
12.02.08	3.817576	3.258794	-43.1104	-3.81132	-2.31986
13.02.08	4.529349	4.627004	-41.3635	-4.46835	-0.50616
14.02.08	4.66817	6.092875	-40.9128	-4.59494	7.496351
15.02.08	3.390087	3.484146	-42.6106	-3.4125	-1.2864
17.02.08	4.201164	4.28987	-45.3105	-4.15758	-2.14873
18.02.08	3.612668	3.704752	-42.3684	-3.61728	-0.73268
19.02.08	3.635297	3.728503	-43.3054	-3.63975	-1.45933
20.02.08	3.590429	3.680794	-43.768	-3.59509	-1.73476
21.02.08	3.903322	3.99664	-43.7663	-3.8882	-0.3102
APRE	**4.122**	**4.8622**	**-42.83**	**-4.28**	**-6.675**
AAPRE	**4.484**	**4.122**	**42.83**	**5.054**	**7.568**
SEE	**5.318**	**4.8**	**42.2**	**5.748**	**31.6**

An error analysis was done for all these correlations and different statistical parameters were calculated. From the error analysis (Table 4.11 and figure 4.8). it is clear that, pressure calculated by Piper , Panhandle A, Panhandle B and Weymouth methods gave good results From the statistical analysis best-performing correlations were Panhandle A and B with AAPRF 4.484 %.and 4.122%. Next best correlations were Weymouth and Piper with AAPRE 5.054% and 7.568%.Error values were presented in Figure 4.8. From evaluating the SEE values, Panhandle B was the best correlation in this segment of the pipe though the difference in SEE value with Panhandle A was very less.

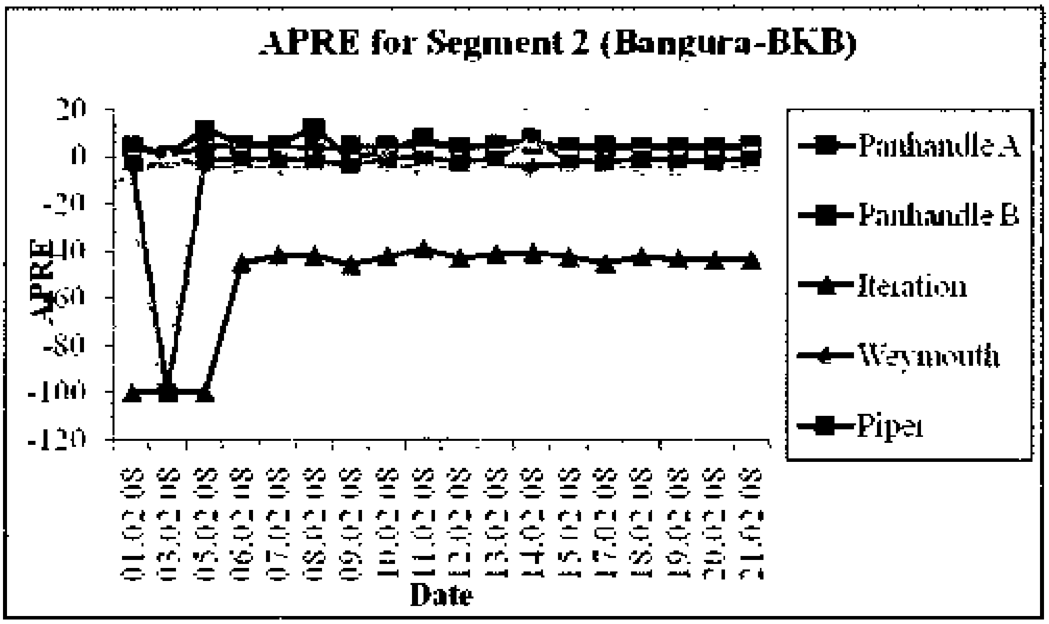

Figure 4.8: APRE for Various Pressure Equations for Segment 2 (Bangura-BKB).

Iteration method (Marching Algorithm) for two-phase flow gave unsatisfactory results in this segment because of limitation of handling the iteration method.

Again, pressure values were predicted with the correlations for Segment -3 (BKB-Demra) where gas flow rate is comparatively lower than other segments and diameter also changes to 20 inches. The calculated values by different pressure correlations are shown in the Table 4.12.

Table 4.12: Calculated Pressures for Segment 3 (BKB-Demra) Diameter 20 inch.

Date	Inlet Pressure (psia)	Outlet Pressures, psia					
		Measured	Panhandle A	Panhandle B	Weymouth	Iteration	Piper
01.02.08	752.7	474.7	670.741	740.19	685.44	0	468
03.02.08	849.7	274.7	848.47	849.607	31.87	0	0
05.02.08	767.7	514.7	678.848	758.97	650	0	609
06.02.08	748.7	474.7	627.775	736.785	568.4	334.097	532
07.02.08	753.7	494.7	610.113	739.61	534	332.088	479
08.02.08	745.7	524.7	619.361	733.27	556.2	327.162	499.5
09.02.08	802.7	614.7	717.18	794.17	680	325.435	628.5
10.02.09	755.7	504.7	627.009	743.024	562	328.527	515
11.02.09	740.7	474.7	595.462	726.503	518	341	474.7
12.02.09	764.7	484.7	675.714	755.87	637	328.073	587

Date	Inlet Pressure (psia)	Outlet Pressures, psia					
		Measured	Panhandle A	Panhandle B	Weymouth	Iteration	Piper
14.02.08	753.7	484.7	637.423	742.222	581.23	336.32	553
15.02.08	772.7	574.7	659.256	761.457	605	334.42	560
17.02.08	790.7	574.7	712.0065	782.874	679	323.85	634
18.02.08	780.7	574.7	712.2841	773.92	685	337.85	656.3
19.02.08	775.7	559.7	699.33	768.124	667.75	439	628.4
20.02.08	785.7	574.7	709.0456	778.084	677.2	343.7	637
21.02.08	773.7	584.7	663.67	762.788	611.47	328.247	582

All calculated values and measured pressure values were presented in the Figure 4.9 (a).From the figures, it is clear that, Iteration Method shows the most deviation from measured values. So, another graph, Figure 4.9 (b) was plotted without these two methods.

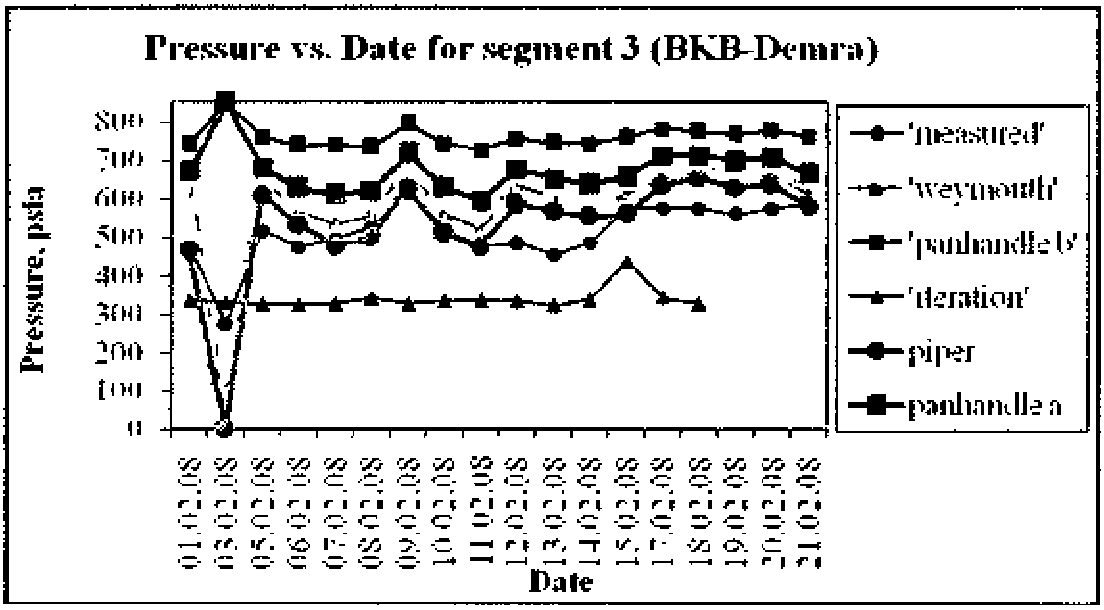

Figure 4.9 (a): Pressure Equations vs. Date for Segment 3 (BKB-Demra).

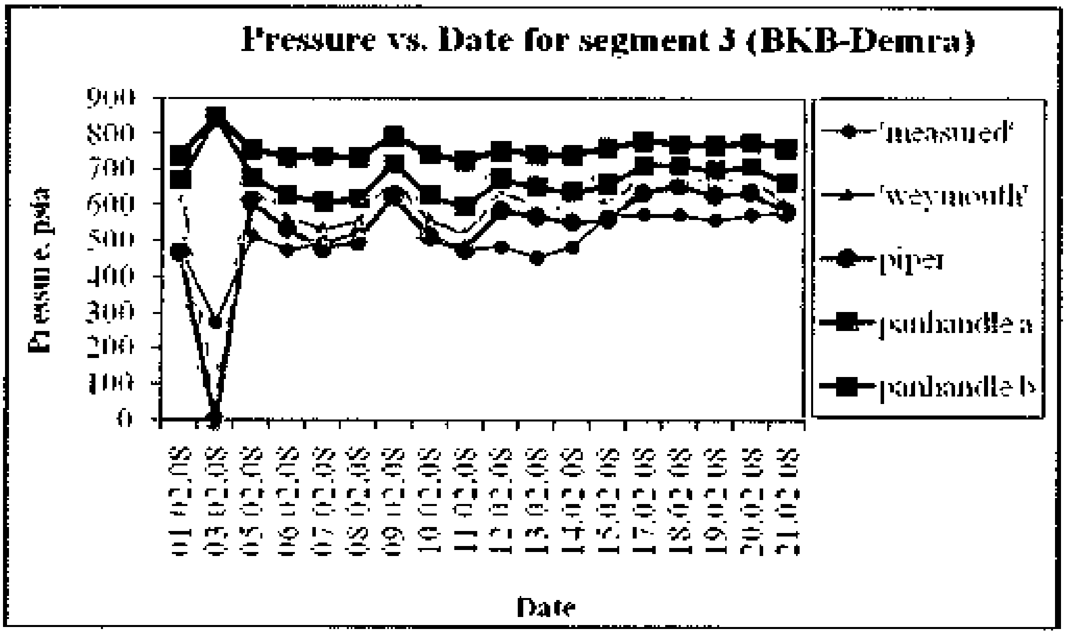

Figure 4.9 (b): Pressure Equations vs. Date for Segment 3 (BKB-Demra) except Iteration Method.

From the figures it is seen that piper correlation flows the measured pressure values very closely. Next better performing correlations are Piper, Weymouth and Panhandle A. Error calculations were performed for this segment and it was represented on Table 4.13 and Figure 4.10.

Table 4.13: APRE comparison for Segment 3 (BKB-Demra).

Date	Panhandle A	Panhandle B	Iteration	Weymouth	Piper
01.02.08	41.2979	55.928	-100	44.3944	-1.4114
03.02.08	208.871	209.285	-100	-88.398	-100
05.02.08	31.892	47.4587	-100	26.2872	18.3214
06.02.08	32.2467	55.2107	-29.619	19.7388	12.0708
07.02.08	23.3299	49.5068	-32.871	7.94421	-3.1736
08.02.08	18.041	39.7503	-37.648	6.00343	-4.8027
09.02.08	16.6715	29.1964	-47.058	10.6231	2.245
10.02.08	24.234	47.2209	-34.906	11.3533	2.04082
11.02.08	25.4396	53.0447	-28.165	9.12155	0
12.02.08	39.4087	55.9459	-32.314	31.4215	21.1058
13.02.08	43.3292	63.7365	-26.377	32.9008	24.9175
14.02.08	31.5088	53.1302	-30.613	19.9154	14.0912
15.02.08	14.7131	32.4964	-41.81	5.27232	-2.5579
17.02.08	23.8919	36.2231	-43.649	18.1486	10.3184

Date	Panhandle A	Panhandle B	Iteration	Weymouth	Piper
18.02.08	23.9402	34.665	-41.213	19.1926	14.1987
19.02.08	24.9473	37.2385	-21.565	19.305	12.2744
20.02.08	23.3766	35.3896	-40.195	17.8354	10.8404
21.02.08	13.5061	30.458	-43.861	4.57842	-0.4618
APRE	**38.861**	**56.816**	**-55.45**	**12.68**	**1.765**
AAPRE	**38.861**	**56.816**	**55.45**	**23.084**	**14.99**
SEE	**30.263**	**41.30118**	**37.8**	**18.3**	**13.65**

From the error analysis, Piper correlation gave the smallest AAPRE (14.99%) and next smaller AAPRE values were given by the Weymouth (23.084%) and Panhandle A (38.861%) correlations. Piper correlation also showed small SEE which was almost 13.65%, where Weymouth showed 18.3% and Panhandle gave 30.263%. It was found that Panhandle A and Iteration predicted pressure better than Panhandle B correlation. Panhandle B gave the worst pressure prediction here. When Panhandle B was used in the piper software it became the best performing correlations where it showed best iteration criterions were met. Figure 4.10 shows the error analysis for segment 3.

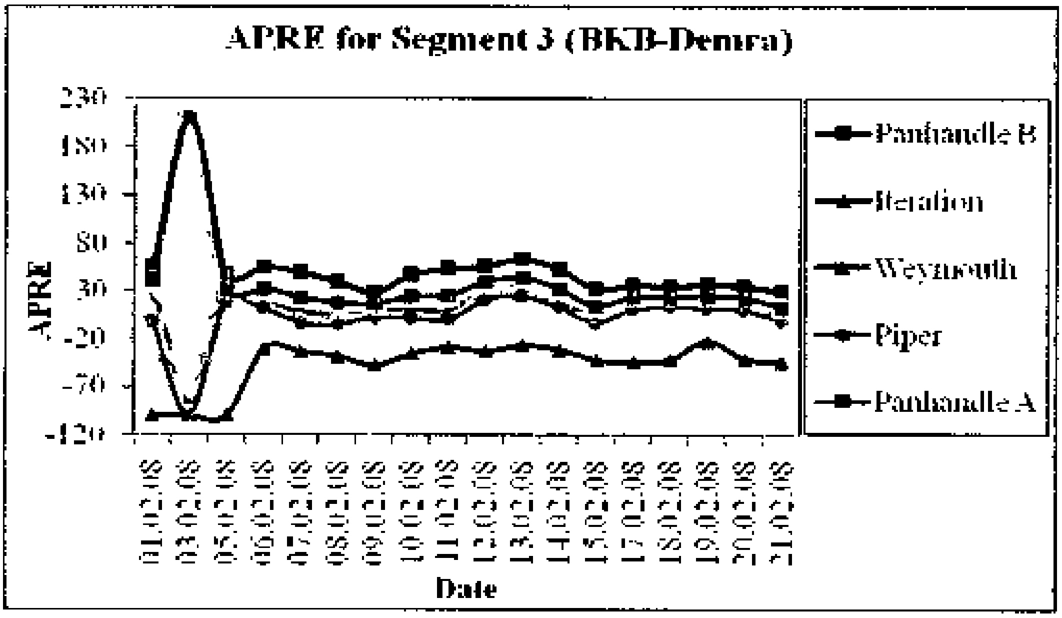

Figure 4.10: APRE for Various Pressure Equations for Segment 3 (BKB-Demra).

From the Figure 4.10, it can be seen that error values by all correlations were high in Segment 3 compared to other sections. This is because of numerous off takes were existed in this segment for different customers and for simplification of the calculations, all of them were ignored. A better result could be found if a pipeline network analysis were done particularly for this segment.

From above analysis it was clear that no single correlation turned out to be the best for all segments which suggested that all correlations have different limiting factors. For instance in literature it was claimed that Panhandle B works well for the long pipe line and large diameter (NPS is over 24) with high Reynolds number. Panhandle A works well for large diameter pipelines with high flow rates. Weymouth equation predicts pressure well for smaller diameter pipelines where diameter is less than 15 inch. According to the literature, Weymouth correlation should not be the best performing correlation for Segment 3 where NPS is greater than 15 inches. This was due to the simplification of the calculations by omitting the off takes. In other sections, literature claims also proved as valid.

In general all correlations gave some errors this is because of the underlying assumptions to formulate the correlations with. The basic assumptions are no mechanical work, steady state flow, isothermal flow, negligible kinetic energy change, constant gas compressibility and horizontal pipeline. In reality none of them can be achieved due to operational and design constraints.

Chapter 5

Conclusion

In this thesis two approaches were used to predict the liquid / condensate volume i.e. Flash method and Heat balance method. All predicted values were compared with measured values. As measured condensate/ liquid values were available only for segment 3, a direct comparison only made here. For instance, gas flow rate at 210 MMSCFD in segment 3, measured condensate value was 800 liters, predicted values were 1362 and 554 liters by flash and heat balance method respectively. This study showed that flash method predicted the condensate volume with a 36% error. This large error was mainly contributed by two things i.e. flow rate change and pipeline undulations.

A trend analysis was done for all measured and calculated values. Flash method was found the best trend following correlations. In measured trend it showed high amplitudes and flash method gave low amplitudes. This suggested that a terrainian slug occurs. It happens due to condensate accumulation in the lower section of the pipeline and builds a backpressure. As the back pressure builds over a critical value it pushes liquid column to the forward section and thus variation in measured condensate amount occurred.

Here two phase flow parameters were calculated and flow pattern was estimated for horizontal and slightly inclined pipe by Taitel and Duckler's equilibrium stratified flow model. Here Taitel and Duckler flow pattern map and transition criterion were used. All cases flow pattern fell into Stratified flow region. From literature, Taitel and Duckler model works well for small diameter pipe with low pressure conditions was reported. But this system has large diameter with high pressure and found no flow transition criterion were met which suggested that more condensate required to make the transition criterion be met. Taitel and Duckler model used air – water system to develop the flow pattern map but not gas condensate flow pattern maps are available in the literature nowadays.

As flow pattern fell in the stratified region, equilibrium exists in gas condensate interface. As gas flow rate was very high compared to liquid generation/ separation rate, single phase pressure correlations can give the best result. In this thesis, several single phase pressure prediction correlations like Panhandle B, Panhandle A and Weymouth were used and they predicted the pressure. Panhandle B was found the best predicting correlation for Segment 1 and Segment 2 with an AAPRE 1.93% and 4.8%. In Segment 3, it is hard to make a decision which correlation predicts very well because the case was simplified which did not match with real scenario. But from the calculation point of view Panhandle B with piper software and Weymouth found best performing correlation with 13% and 18% error respectively. Two phase flow pressure equation i.e. Marching Algorithm performance was found unsatisfactory due to computational constraints.

CHAPTER 6

FUTURE RECOMMENDATIONS

Taitel and Duckler flow pattern for equilibrium stratified flow was used here with air water system. Taitel- Duckler model is to be modified for gas-condensate system for large diameter and high pressure system.

Single phase pressure correlations have a definite working boundary, such as Panhandle A and B work well for nominal pipe size 24 and above. On the other hand Weymouth equation works well for the smaller diameter i.e. up to 15 inch. For pipe size in between 15 to 24 inch these correlations should be tuned or modified by field data. If possible modify the correlations or different correlations could be developed.

Condensate data were available only for Segment-3. Condensate data for other segments would enhance the chance to analyze the different condensate predicting method's performance more confidently. More over a computational knowledge would greatly help to estimate the condensate and pressure values precisely and help to pin down the error in smaller margin.

References

1. Brill P. James and Mukherjee Hemanta. "Multiphase Flow in wells ". First Printing, Henry L. Doherty Memorial Fund of AIME. Society of Petroleum Engineers Inc. Richardson, Texas. 1999.

2. Felder M. Richard and Rousseau W. Ronald. "Elementary Principles of the Chemical Process". 3rd Edition , 2000.John Wiley & Sons, NY.

3. Shoham Ovadia. "Mechanistic Modeling of Gas-Liquid Two Phase Flow in Pipes"The Society of Petroleum Engineers SPE, 2005, pp. 1-25, 42-54 and 57-87.

4. Kumar S. Contributions in Petroleum Geology and Engineering. Volume 4. "Gas Production Engineering". Gulf Publishing Company. Book Division. Copyright © 1987 by Gulf Publishing Company, Houston, Texas. P. 275-291.

5. John L. Kennedy." Oil and Gas Pipeline Fundamentals". 2nd Edition. Copyright © 1993 by PennWell Publishing Company, 1421 South Sheridan/P.O. Box 1260, Tulsa Oklahoma, 74101.

6. Smith J.M., Van Ness H.C. and Abbott M.M. "Introduction to Chemical Engineering Thermodynamics". 4th Edition. 1987. The McGraw-Hill Companies, InC.

Appendix A

Theory

A.1 Equilibrium Constant:

The Wilson correlation can be used to estimate K values initially. Wilson proposed this simplified thermodynamic expression to estimate K values,

$$K_i = \left(\frac{P_{ci}}{P}\right) \exp\left[5.37\,(1+\omega i)\left\{1-\left(\frac{T_{ci}}{T}\right)\right\}\right] \qquad \text{(A.1.1)}$$

Where P_{ci} = critical pressure of component i, psia; P = system pressure, psia; T_{ci} = critical temperature of component i, °R; T = system temperature, °R; and ω_i = acentric factor of component i.

The following is a step-by-step procedure to calculate equilibrium constants.

1. The input data required for this calculation are the system pressure, p, temperature, T, and the overall system composition. Z_i, for each component.
2. On the basis of Eq. A.1.1, K_i^{A} values for each component are estimated.
3. On the basis of assumed K_i^{A} values from Eq. A.1.1 and the known z_i values, flash calculations are performed by using Eq. A.2.8.
4. With appropriate EOS's, the composition of the liquid and gas phases obtained from flash calculation can be used to determine the fugacity coefficients of each component in each phase, $\Phi^{L}{}_{i}$ and $\Phi^{V}{}_{i}$.
5. Find K_i^{C} from

   $$K_i^{C} = \frac{y_i}{x_i} = \frac{\Phi_i L}{\Phi_i V} \qquad \text{(A.1.2)}$$

6. Compare the equilibrium constants in step 2 with the calculated values in step 5 using the following convergence criteria:

   $$\sum ((K_i^{C}/K_i^{A})-1)^2 \leq 10^{-4}$$

7. If the convergence criterion in step 6 is satisfied for all components, the values of equilibrium constants are used to calculate phase transitions required in determining phase physical properties. Otherwise, these calculated values are used as the new guesses, and steps 3through 6 are repeated until convergence is achieved.

A.2 Flash Calculations:

Some important variables are required in the material balance:

F = number of moles of feed,
L = number of moles of liquid
V =number of moles of vapor,
Z_i = mole fraction of component I in feed,
Xi = mole fraction of component I in liquid phase,
Yi = mole fraction of component I in vapor phase, and
N = total number of components in the feed.

The overall material balance is

$$F = L + V \tag{A.2.1}$$

Individual component balances are

$$Z_i F = xi\, L + y_i\, V \tag{A.2.2}$$

Eq. A.2.1 may be rewritten as

$$\frac{L}{F} = 1 - \frac{V}{F} \tag{A.2.3}$$

Liquid Phase. Solving Eq. A.2.2 for **xi**, and replacing y_i and $\frac{L}{F}$ with $T_{av}=(\frac{T1 + T2}{2})$ and A.2.3, respectively results in

$$Xi = z_i F - y_i \frac{V}{L} = \frac{(zi - yi\frac{V}{F})}{\frac{L}{F}} = \frac{(zi - xiKi\frac{V}{F})}{\{1 - (\frac{V}{F})\}} \tag{A.2.4}$$

On further simplification,

$$X_i = \frac{zi}{\{1 + (\frac{V}{F}(Ki - 1)\}} \tag{A.2.5}$$

Vapor Phase. Similarly, solving Eq. A.2.2 for y_i and using $T_{av}=(\frac{T1 + T2}{2})$ and A.2.3 gives

$$Yi = zi\, F - xi \frac{L}{V} = \frac{\{zi\, F - (\frac{T1}{Ki})\}}{V} \tag{A.2.6}$$

Eq. A.2.6 can be simplified as

$$Yi = \frac{ziKi}{\{1 + \{(\frac{V}{F})(Ki - 1)\}\}} \tag{A.2.7}$$

Solution Procedure. It is important to note that the solutions for x_i and y_i based on eqs. A.2.5 and A.2.7, respectively, require the determination of the $\frac{V}{F}$ value. To solve for $\frac{V}{F}$, Eqs. A.2.5 and A.2.7 are combined by use of this definition of mole fractions.

$\sum xi = \sum yi = 1.0$

Or,

$\sum (yi - xi) = 0$

Therefore, combining Eqs. A.2.5 and A.2.7 give

$$\sum (yi - xi) = \sum \frac{zi\,(Ki - 1)}{\{(Ki - 1)\frac{V}{F}\} + 1} = f(\frac{V}{F}) = 0 \qquad (A.2.8)$$

It is often called the Rachford and Rice equation, which is implicit in V/F. To solve it, K values must be known. As K values are phase-composition dependent and phase compositions are unknown in flash calculations, this imposes a second level of implicitness in Eq. A.2.8. Therefore, solution of Eq. A.2.8 is computer intensive.

The procedure requires that these variables be known:

1. Number of components, n.
2. Mole fraction of each component in the feed mixture, z_i.
3. Equilibrium constant for each component, K_i.
4. A first guess for the mole ratio, $(\frac{V}{F})j$, where the subscript j refers to the iteration step count.

Given at first guess, for instance 0.5 for $(\frac{V}{F})j$, an improved value of $(\frac{V}{F})$ can be estimated from

$$(\frac{V}{F})j+1 = (\frac{V}{F})j - (f(\frac{V}{F})j)/(f'(\frac{V}{F})j) \qquad (A.2.9)$$

Where the derivative, $F'(\frac{V}{F})$, is obtained by differentiating Eq. A.2.8 with respect to $\frac{V}{F}$.

$$F'(\frac{V}{F})j = - \sum \frac{\{zi(Ki - 1)^2\}}{\{(Ki - 1)(\frac{V}{F})j + 1\}^2} \qquad (A.2.10)$$

Convergence is achieved when

$$ABS\{((\frac{V}{F})j+1 - (\frac{V}{F})j)\} < 1.0 \times 10^{-6} \qquad (A.2.11)$$

Once convergence is obtained for $\frac{V}{F}$, the composition of each phase can be determined from Eq. A.2.5 and either $T_{av}=(\frac{T1 + T2}{2})$ or A.2.7.

The previous procedure requires values for K at the pressure, temperature, and composition of each phase.

A.3 Flow Pattern Prediction:

The starting point of the model is equilibrium stratified flow. Assuming stratified flow to occur, the flow variables; including the equilibrium liquid level in the pipe for a given set of flow conditions (for example, pipe diameter, inclination angle and physical properties of the phases) are determined. A stability analysis is then performed to determine whether or not the flow configuration is stable. If the flow is stable- stratified flow occurs. If the flow in unstable, a change to non-stratified flow occurs, and the resulting flow pattern is determined. This requires stability analysis. In this model a simplified Kelvin-Helmholtz stability analysis is applied. Taitel and Dukler extended this analysis to the case of a stationary finite wave on the gas-liquid interface in a pipe flow.

Extensive studies on two-phase flow pattern transitions have been conducted since the early 1950's. Most of the initial work has been focused on horizontal or vertical flow. Inclined flow studies have been initiated in the 1970's, leading to a complete understanding of flow pattern transitions in the entire range of inclination angle, namely, from -90^0 to 90^0.

The most common approach for two-phase flow pattern determination has been visual observation of the flow in a transparent pipe. Usually the data have been mapped on a two dimensional plot and the boundaries between the different flow patterns have been determined. In the initial studies, no physical basis has been suggested for the selection of mapping coordinates. Therefore, these empirical maps are reliable only in the narrow range of conditions under the data have been acquired, and extension for other flow conditions is uncertain. Also, different flow pattern classifications and definitions have been suggested by the various investigators, resulting a poor agreement between their proposed maps.

Since visual observations are often subjective and difficult, especially at high flow rates, efforts have been devoted to developing flow pattern detection techniques, which are objective and can also be used in opaque pipes. Many such devices have been suggested, including hot wire anemometry, x-ray, pressure transducers and conductance probes. All efforts done in this approach have resulted in partial success, since no single technique is able to distinguish between all the flow patterns under different flow conditions confidently.

Beginning at the mid 1970, analytical models for flow pattern prediction, based on the physical phenomena, have been reported. The main advantage of these models is that they can be extrapolated with more confidence to conditions for which no data are available. Also, they provide physical insight and increase the understanding of the flow transition phenomena.

Equilibrium Stratified Flow

Equilibrium stratified flow configuration is shown schematically in Figure A.3.1. The pipeline is inclined at θ inclination angle from the horizontal, and the gas and the liquid average velocities are v_G and v_L, respectively. Also shown is a cross sectional area of the pipe with the geometrical parameters. The area for flow and wetted perimeter of the gas and liquid phases are A_G and S_G, and A_L and S_L, respectively. The interface length is S_I, and the liquid level (under equilibrium conditions) is h_L.

The objective of this part of the model is to determine the equilibrium level in the pipe, h_L, for a given set of flow conditions, namely the gas and liquid flow rates, pipe diameter and inclination angle and the physical properties of the phases. Applying momentum balances on the gas carries this out and the liquid phases in a differential control volume with a axial length of ΔL, as shown in the Figure A.3.1.

Figure A.3.2 is an expansion of the control volume, where the two phases are shown separated from each other, and the forces acting on each of the phases are indicated. For steady-state flow, neglecting the rate of change of momentum across the control volume, the momentum balances are reduced to force balances.

The momentum (force) balances for the liquid and gas phases are given, respectively. by

$$-A_L\,\left[\frac{dp}{dL}\right]_L - \tau_{WL}\,S_L + \tau_I\,S_I - \rho_L\,A_L\,g\,\sin\theta = 0 \qquad (A.3.1)$$

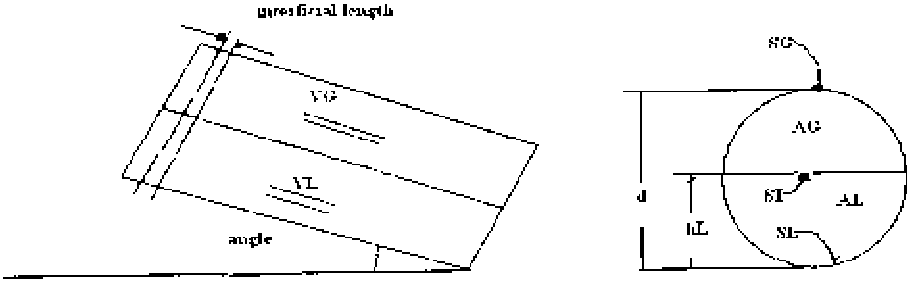

Figure A.3.1: Equilibrium Stratified flow.

$$-A_G \left[\frac{dp}{dL}\right]_G - \tau_{WG} S_G - \tau_I S_I - \rho_G A_G g \sin\theta = 0 \qquad (A.3.2)$$

Eliminating the pressure gradient from Eq. A.3.1 and Eq. A.3.2, the combined momentum equation for the two phases is obtained, as follows,

$$\tau_{WG}\left(\frac{SG}{AG}\right) - \tau_{WL}\left(\frac{SL}{AL}\right) + \tau_I S_I \left\{\left(\frac{1}{AL}\right)+\left(\frac{1}{AG}\right)\right\} - (\rho_L - \rho_G) g \sin 0 = 0 \qquad (A.3.3)$$

The combined momentum equation is an implicit equation for h_L, the liquid level in the pipe. It combines all the forces that act on the liquid and gas phases, which in turns determine the location of the liquid level in the pipe. In order to solve the equation for h_L, it is necessary to determine the different geometrical and force variables in the equation. The calculation of the forces in the equation is carried out utilizing single-phase flow method based on the hydraulic diameter concept. The respective hydraulic diameters of the liquid and gas phases are given by

$$d_L = \left(\frac{4\,AL}{SL}\right) \text{ and } \quad d_G = \left(\frac{4AG}{SG + SI}\right) \qquad (A.3.4)$$

The Reynolds numbers and the friction factors (for a smooth pipe) of each of the phases are,

$$f_L = c_L \left(\frac{dL\, vL\, \rho L}{\mu L}\right)^{-n} \quad \text{and} \quad f_G = c_G \left(\frac{dG\, vG\, \rho G}{\mu G}\right)^{-m} \qquad (A.3.5)$$

Where $c_I = c_G = 16$ and $m = m = 1$ for laminar flow, $c_L = c_G = 0.046$ and $m = n = 0.2$ for turbulent flow.

The wall shear stresses corresponding to each phase are

$$\tau_{WL} = f_L\left(\frac{\rho L\, vL^2}{2}\right) \quad \text{and} \quad \tau_{WG} = f_G\left(\frac{\rho G\, vG^2}{2}\right) \qquad (A.3.6)$$

The interfacial shear stress is given, by definition, as

$$\tau_I = f_I \left\{\frac{\rho G\,(vG - vI)^2}{2}\right\} \qquad (A.3.7)$$

In this model it is assumed that $f_I = f_G$, namely, a smooth interface exits. In addition the interface velocity is neglected, i.e., $v_G >> v_I$. With these approximations the interfacial shear stress is equal to the gas-phase wall shear stress.

Substitution of Eq. A.3.4 through Eq. A.3.7 into Eq. A.3.3 enables the determination of the liquid level in the pipe. However, the final solution for h_L is presented in a dimensionless form. All the variables can be written in non-dimensional form by

area and v_{SL} and v_{SG} for the liquid and gas velocities, respectively. The dimensionless variables are designated by a tilde(~), as follows,

$$\tilde{S}_L = \frac{SL}{d},\ \tilde{h}_L = \frac{hL}{d},\ \tilde{A}_L = \frac{AL}{d^2},\ \tilde{v}_L = \frac{vL}{vSL},\ v_G = \frac{vG}{vSG} \qquad (A.3.8)$$

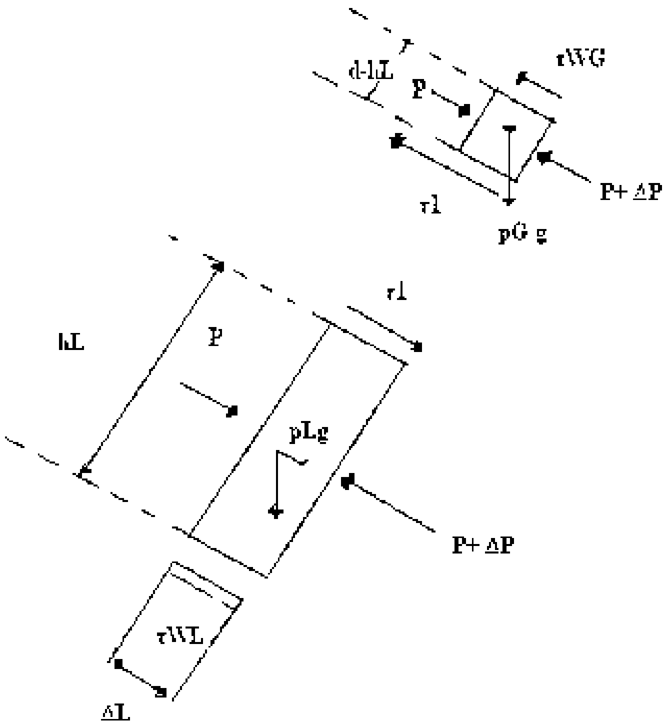

Figure A.3.2: Equilibrium Stratified Flow Gas and Liquid Momentum Balances.

Rearranging Eq. A.3.3 yields

$$\left(\frac{\tau WL\ SL}{\tau WG\ AL}\right) - \left\{\left(\frac{SG}{AG}\right) + \left(\frac{SI}{AL}\right) + \left(\frac{SI}{AG}\right)\right\} + \left\{\frac{(\rho L-\rho G)\, g \sin\theta}{\tau WG}\right\} = 0 \qquad (A.3.3)^{/}$$

Substituting the dimensionless parameters into Eq. (A.3.3)$^{/}$ results in the dimensionless form of the combined momentum equation, namely,

$$X^2[(\tilde{v}_L\, \tilde{d}_L)^{-n}\, \tilde{v}_L^{\,2} (\tfrac{\tilde{S}L}{\tilde{A}L})] - [(\tilde{v}_G\, \tilde{d}_G)^{-m}\, \tilde{v}_G^{\,2}\{(\tfrac{\tilde{S}G}{\tilde{A}G}) + (\tfrac{\tilde{S}I}{\tilde{S}L}) + (\tfrac{\tilde{S}I}{\tilde{S}G})\}] + 4y = 0 \qquad (A.3.9)$$

Two dimensionless groups emerged from the analysis, namely, X, the Lockhart and Martinrlli parameter and Y, an inclination angle parameter, given by

$$X^2 = \frac{\left(\frac{4cL}{d}\right)\left(\frac{\rho L\, vSL\, d}{\mu L}\right)^{-n}\left(\frac{\rho L\, vSL^2}{2}\right)}{\left(\frac{4cG}{d}\right)\left(\frac{\rho G\, vSG\, d}{\mu G}\right)^{-m}\left(\frac{\rho G\, vSG^2}{2}\right)}$$

$$= \frac{(-dp/dL)SL}{(-dp/dL)SG} \qquad (A.3.10)$$

$$Y = \frac{\{(\rho L - \rho G)\, g \sin\theta\}}{\left(\frac{4CG}{d}\right)\left\{\left(\frac{\rho G\, vSG\, d}{\mu G}\right)^{-n}\right\}\left(\frac{\rho G\, vSG^2}{2}\right)}$$

$$= \frac{\{(\rho L - \rho G)\, g \sin\theta\}}{(-dp/dL)SG} \qquad \text{(A.3.11)}$$

All the dimensionless variables in Eq. A.3.9 are unique functions of the dimensionless liquid level,

$\tilde{h}_L = \frac{hL}{d}$. Thus it is proved that

$$\tilde{h}_L = f(X, Y) \qquad \text{(A.3.12)}$$

The functional relationships between the tilde dimensionless variables and $\tilde{h}_L$ are given below in Eq. A.3.13 (refer to Figure A.3.3), as follows,

$$\tilde{A}_L = 0.25[\Pi - \cos^{-1}(2\tilde{h}_L - 1) + (2\tilde{h}_L - 1)\sqrt{\{1 - (2\tilde{h}_L - 1)^2\}}]$$

$$\tilde{A}_G = 0.25[\cos^{-1}(2\tilde{h}_L - 1) - (2\tilde{h}_L - 1)\sqrt{\{1 - (2\tilde{h}_L - 1)^2\}}]$$

$$\tilde{S}_L = \Pi - \cos^{-1}(2\tilde{h}_L - 1)$$

$$\tilde{S}_G = \cos^{-1}(2h_L{}^{\sim} - 1)$$

$$\tilde{S}_I = \sqrt{\{1 - (2\tilde{h}_L - 1)^2}$$

$$\tilde{v}_L = (\tilde{A}_P / \tilde{A}_L)$$

$$\tilde{v}_G = (\tilde{A}_P / \tilde{A}_G) \qquad \text{(A.3.13)}$$

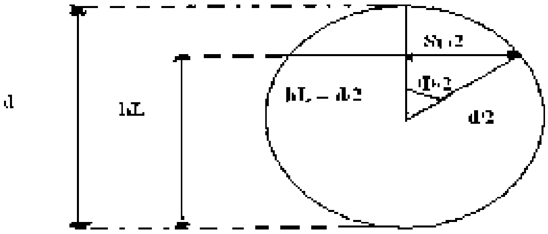

Figure A.3.3: Geometrical parameters for stratified flow.

A plot of $\tilde{h}_L$ as a function of X and Y is given in Figure 4. This is an example of similarity analysis through basic equations. The dimensionless groups controlling the phenomena have been obtained based on the proposed model. The generalized chart should be applicable to all flow conditions, for horizontal and slightly inclined flow, subject to model assumptions. There are four possible solutions, depending upon the gas and liquid phases being in laminar and' or turbulent flow. X and Y can be determined from the inlet flow conditions. Then depending upon the flow conditions of the gas and liquid phases the liquid level can be determined from the chart. The solution can be also obtained by solving either Eq. A.3.3 or Eq. A.3.9.

A.4 Pressure Calculation:

Many pipeline equations have been developed from the basic mechanical energy balance (Eq.A.3.13):

$$dp + \left(\frac{\rho}{2g_c}\right) dv^2 + \left(\frac{g}{g_c}\right)\rho dz + \rho dl_w = 0 \tag{A.4.1}$$

Assuming horizontal, steady-state, adiabatic, isothermal flow of gas, with negligible kinetic-energy change, Eq. A.4.1 becomes:

$$dp + \rho dl_w = 0$$

Frictional losses for a length dL of pipe are given by (Eq. A.4.1):

$$\rho dl_w = \left(\frac{f \rho v^2}{2 g_c d}\right) dL$$

Substituting for frictional losses:

$$dp + \left(\frac{f \rho v^2}{2 g_c d}\right) dL = 0$$

Substituting for gas density $\rho = \frac{pM}{ZRT}$

and gas velocity $v = q_{sc} \left\{\left(\frac{ZTp_{sc}}{pT_{sc}}\right)\left(\frac{4}{\pi d^2}\right)\right\}$

it is obtained that

$$- dp = \left(\frac{f}{2 g_c d}\right)\left(\frac{pM}{ZRT}\right)\left(\frac{16 q_{sc}^2 Z^2 T^2 p_{sc}^2}{T_{sc}^2 p^2 \pi^2 d^4}\right)$$

or,

$$-\int\left(\frac{p}{Z}\right)dp = \frac{8fMT\, p_{sc}^2 q_{sc}^2}{R \pi^2 d^5 g_c T_{sc}^2} \int dL \tag{A.4.2}$$

Here T is constant (or, independent of length) since isothermal flow is assumed. Otherwise, an average temperature, T_{av}, is commonly used instead of T in the previous relationship. The two types of averages used are the arithmetic average,

$$T_{av} = \left(\frac{T1 + T2}{2}\right) \tag{A.4.3}$$

And the log-mean average.

$$T_{av} = \frac{(T1 - T2)}{\ln(T1/T2)} \tag{A.4.4}$$

In practice, both these averages are quite close since temperature T_1 and T_2 are used as absolute temperatures. Using an average temperature is practically expedient, because an analytical description of the variation of temperature along the pipeline length is rather difficult and introduces some complexity. The gas compressibility factor, Z, is made independent of temperature and pressure by using an average compressibility factor, Z_{av}, for simplicity. Integrating over the pipe length from 0 to L and pressure p_1 (at L = 0, at the upstream) to p_2 (at L = L, at the downstream end), Eq. A.4.2 becomes:

$$-\left\{\frac{(p1^2 - p2^2)}{2}\right\} = \left[\left\{\frac{(8)(28.97)}{(R\Pi^2 g_c)}\frac{Tsc^2}{psc^2}\right\}\left(\frac{qsc^2 \gamma g Zav T f L}{d^5}\right)\right.$$

$$\text{or,} \qquad q_{sc}^2 = \left(\frac{Rgc}{46.9644}\right)\left(\frac{Tsc^2}{psc^2}\right)\left\{\frac{(p1^2 - p2^2)d^5}{(\gamma g Zav T f L)}\right\} \tag{A.4.5}$$

any consistent set of units can be used in Eq. A.4.1. In common units, with q_{sc} in MCSFD, p in psia, T in °R, d in inch, L in ft, and with R = 10.732 psia ft^3/ lb mole °R and g_c = 32.17 lbm ft/lbf-sec^2, Eq. A.4.5 becomes:

$$\left(\frac{10^3}{3600 \times 24}\right)^2 q_{sc}^2 = \frac{\{(10.732 \times 144)(32.17)(\frac{1}{12})^5\}}{[(46.9644)\left(\frac{Tsc^2}{psc^2}\right)\left\{\frac{(p1^2 - p2^2)d^5}{\gamma g Zav T f L}\right\}]}$$

$$\text{or,} \qquad q_{sc} = 5.6353821\left(\frac{Tsc}{psc}\right)\left[\frac{\{(p1^2 - p2^2)d^5\}}{(\gamma g Zav T f L)}\right]^{0.5} \tag{A.4.6}$$

where q_{sc} = gas flow rate measured at standard conditions, MSCFD

p_{sc} = pressure at standard conditions, psia

T_{sc} = temperature at standard conditions, °R

p_1 = upstream pressure, psia

p_2 = downstream pressure, psia

d = diameter of the pipe, ft

γ_g = gas gravity (air = 1 basis)

T = flowing temperature, °R

Z_{av} = average gas compressibility factor

f= Moody friction factor

L = Length of the pipe, ft

Eq. A.4.6, Attributed to Weymouth, is the general equation for steady state isothermal flow of gas through a horizontal pipe.

Appendix B

TABLES

Table B.1: (a) Physical Properties for segment 2 (Bangura-BKB).

Date	Oil viscosity, cp	Gas Density, Kg/m^3	Gas viscosity, cp
01.02.08	2.132	92.966	0.01369
03.02.08	2.1745	88.2726	0.01346
05.02.08	2.1592	93.5017	0.0137
06.02.08	2.2142	93.3	0.01366
07.02.08	2.1838	94.102	0.0129
08.02.08	2.2161	92.818	0.0137
09.02.08	2.0729	93.482	0.0137
10.02.08	2.07366	92.609	0.0137
11.02.08	2.14263	96.1305	0.01383
12.02.08	2.1643	93.173	0.01308
13.02.08	2.16975	94.3813	0.013746
14.02.08	2.1664	95.2327	0.013777
15.02.08	2.1364	94.86934	0.0137798
17.02.08	2.135128	94.1328	0.013747
18.02.08	2.10604	96.8229	0.01388
19.02.08	2.1558	94.4832	0.01375
20.02.08	2.10787	94.6766	0.01379
21.02.08	2.13615	93.43	0.013718

Table B.1: (b) Physical Properties for segment 3 (BKB-Demra).

Date	Oil viscosity, cp	Gas Density, Kg/m^3	Gas viscosity, cp
01.02.08	2.52	71.288	0.01254
03.02.08	2.527	62.7	0.01228
05.02.08	2.51423	70.4389	0.012564
06.02.08	2.6076	71.884	0.0125711
07.02.08	2.6385	71.8148	0.012553
08.02.08	2.74	71.07	0.0124
09.02.08	2.373	69.9446	0.012623
10.02.08	2.64	71.09	0.0125
11.02.08	2.608	73.554	0.01263
12.02.08	2.5446	70.6466	0.012556
13.02.08	2.6113	72.18956	0.01258
14.02.08	2.6445	72.3164	0.01257
15.02.08	2.56088	72.14438	0.0126044
17.02.08	2.52	70.10837	0.01255
18.02.08	2.43777	72.7427	0.01269

Date	Oil viscosity, cp	Gas Density, Kg/m^3	Gas viscosity, cp
19.02.08	2.25826	95.1453	0.01371
20.02.08	2.1852	94.6761	0.01374
21.02.08	2.532356	70.759	0.01256

Table B.2: (a) Ashuganj Condensate calculation (L/F =.007)

Date	Gas Flow Rate (MSC FD)	Gas lb Mole /day	Liquid lb Mole	Liquid Mass	Liq. Volume (ft^3)	Liquid Volume (liter)
01.02.08	321300	992907	6950.35	343903.546	5.48	155.2
03.02.08	170000	501736.9	3512.15	173781.591	2.77	78.4
05.02.08	295800	874988.4	6124.92	303061.004	4.83	136.7
06.02.08	323000	956524.3	6695.67	331301.77	5.28	149.5
07.02.08	317050	938904.1	6572.33	325198.843	5.18	146.7
08.02.08	316200	938503.1	6569.52	325059.939	5.18	146.7
09.02.08	277100	817831.1	5724.81	283263.993	4.51	127.8
10.02.08	309400	916249.6	6413.74	317352.222	5.05	143.2
11.02.08	321300	951490	6660.43	329558.077	5.25	148.7
12.02.08	306000	906180.9	6343.26	313864.835	5	141.6
13.02.08	317900	940361.1	6582.52	325703.493	5.2	146.9
14.02.08	317050	936791.8	6557.54	324467.237	5.17	146.4
15.02.08	307700	909165.3	6364.15	314898.498	5.018	142.1
17.02.08	283050	844885	5914.19	292634.376	4.66	132
18.02.08	290700	867719.7	6074.038	300543.413	4.79	135.6
19.02.08	298350	881538.7	6170.77	305329.759	4.86	137.8
20.02.08	295800	873022.1	6111.15	302379.968	4.82	136.4
21.02.08	297500	880017.1	6160.12	304802.734	4.85	137.5

Table B.2: (b) Ashuganj Condensate calculation (Z factor calculation).
S.G. Gas = 0.57595, P_c = 680.1564, T_c = 345.9474

Date	p psig	P psia	T (°F)	P_r	T_r	Z
01.02.08	777	791.7	513.6	1.163997	1.484619	0.85
03.02.08	819	833.7	513.6	1.225748	1.484619	0.89
05.02.08	790	804.7	513.6	1.18311	1.484619	0.888
06.02.08	786	800.7	513.6	1.177229	1.484619	0.887
07.02.08	786	800.7	513.6	1.177229	1.484619	0.887
08.02.08	774	788.7	513.6	1.159586	1.484619	0.885
09.02.08	802	816.7	513.6	1.200753	1.484619	0.89
10.02.08	786	800.7	513.6	1.177229	1.484619	0.887
11.02.08	782	796.7	513.6	1.171348	1.484619	0.887
12.02.08	783	797.7	513.6	1.172819	1.484619	0.887

Date	P psig	P psia	T (°F)	P_r	T_r	Z
13.02.08	790	804.7	513.6	1.18311	1.484619	0.888
14.02.08	796	810.7	513.6	1.191932	1.484619	0.889
15.02.08	798	812.7	513.6	1.194872	1.484619	0.889
17.02.08	802	816.7	513.6	1.200753	1.484619	0.88
18.02.08	805	819.7	513.6	1.205164	1.484619	0.88
19.02.08	797	811.7	513.6	1.193402	1.484619	0.889
20.02.08	803	817.7	513.6	1.202224	1.484619	0.89
21.02.08	790	804.7	513.6	1.18311	1.484619	0.888

Table B.2: (c) Condensate calculation using C_p (Date: 01/02/2008).

Place	Composition	Mole%	MW, (lb/lb-mol)	MW of Mix (n*MW)	Heat of condensation (Btu/Lb)	∑nH (Btu/lb)
AGMS	i-butane	0.10567	58	6.12886	-157.67	-16.66
	n-butane	0.08757	58	5.07906	-165.79	-14.52
	i-pentane	0.03639	72	2.62008	-153.72	-5.59
	n-pentane	0.0189	86	1.6254	-144.08	-2.72
			100	0	-136.12	0
				15.4534		**-39.5**
Meghna-ghat	i-butane	0.13	58	7.54	-157.68	-20.5
	n-butane	0.05	58	2.9	-165.79	-8.29
	i-pentane	0.04	72	2.88	-147.26	-5.89
	n-pentane	0.01	72	0.72	-153.72	-1.54
	hexane	0.02	86	1.72	-144.08	-2.88
	heptane+	0.04	100	4	-136.12	-5.44
				19.76		**-44.5413**
Demra	i-butane	0.157	58	9.106	-157.6735545	-24.7547
	n-butane	0.082	58	4.756	-165.7896114	-13.5947
Demra	i-pentane	0.044	72	3.168	-147.2595261	-6.47941
	n-pentane	0.028	72	2.016	-153.723109	-4.30424
	hexane	0.047	86	4.042	-144.0793744	-6.77173
	heptane+	0.047	100	4.7	-136.1268436	-6.39796
				27.788		**-62.3028**

Table B.2: (d) Ashuganj Condensate calculation using C_p (Date: 01/02/2008) contd. Basis: 100 mol.

ΔT, °C	CPΔT	n, lbmole/ lbmole	n, lb mole	mass (lb)	density, lb/ft^3	volume, Litre	total condensate, Litre
-0.5285	1505.32	-0.0262	-2.624	40.5467	8.473	135.5	
0.4410	1508.496	-0.0295	-2.9527	58.345	6.42	257.344	
-0.625	1244.995	-0.05	-5.00426	139.058	7.1455	551.073	**943.925**

Table B.2: (e) Condensate calculation by V/F method:

Date	AGMS, Litre	BKB, Litre	Demra, Litre	Total (calculated)	Total (measured)
01.02.08	155.2	206.35	105.973	**467.5247**	**800**

Table B.3: (a) Flash calculation for Demra (K calculation).

Name	T_c, R	T_B,R	P_c, atm	a=(log P_c-1.167)/(T_c - T_b)	ω_i = (3/7*T_b*a)$^{-1}$	exp calc.	z_i	K_i=K_1 A
methane	343.68	201.534	45.38	0.00344	-0.702	0.58	96.537	5.29
ethane	550.14	332.826	48.08	0.00236	-0.662	-0.04	2.193	3.02
propane	666.24	416.544	41.92	0.001824	-0.674	-0.41	0.453	1.8
i-butane	735.18	471.012	37.46	0.00154	-0.69	-0.61	0.157	1.33
n-butane	765.78	491.388	36	0.001419	-0.701	-0.68	0.082	1.19
pentane	846.06	542.382	33.25	0.001168	-0.728	-0.83	0.072	0.94
hexane	972.96	557.178	27.04	0.00064	-0.847	-0.66	0.047	0.91
heptane	1024.2	616.002	24.57	0.000547	-0.855	-0.70	0.047	0.79

Table B.3: (b) Flash calculation for Demra (K calculation) contd.

	let $(v/f)_i$=0.99				
name	f(v/f)	f' ' (v/f)	$(v/f)_{i+1}$	calc. x_i	calc. y_i
methane	78.92717	-64.5296	**2.218675**	18.39911	97.32627
ethane	1.476565	-0.99418		0.731201	2.207766
propane	0.203334	-0.09127		0.2517	0.455033
i-butane	0.039023	-0.0097		0.118367	0.15739
n-butane	0.013345	-0.00217		0.068789	0.082133
pentane	-0.00429	-0.00026		0.076248	0.071957
hexane	-0.00444	-0.00042		0.051397	0.046956
heptane	-0.01192	-0.00303	**diff**	0.058806	0.046881
			1.228675	**$\sum x_i$=19.75**	**$\sum y_i$=100.4**

Table B.3: (c) Flash calculation for Demra (K calculation, contd.)

name	Z	Fugacity f	Φ_i v	Φ_iL	new k_i=k_iC	$(k_iC/K_iA)-1)^2$
methane	0.286	6.975	8.05E-05	0.000426	5.2897285	4.93038E-32
ethane	0.279	6.650	0.003385	0.01022	3.01936843	4.93038E-32
propane	0.276	6.516	0.016092	0.029091	1.80784285	4.93038E-32
i-butane	0.274	6.428	0.045895	0.061025	1.32967666	0
n-butane	0.282	6.787	0.092858	0.110872	1.19399559	0
pentane	0.27	6.256	0.097695	0.092197	0.94371918	0
hexane	0.266	6.088	0.1457	0.13311	0.91358795	1.2326E-32
heptane	0.261	5.885	0.141061	0.112456	0.79721645	1.2326E-32
						$\sum$=1.7256E-31

Table B.4: (a) Pressure calculation using Weymouth Equation. (Date: 02/02/2008)
S.G. =0.57595, Z_{av} = 0.889, P_{sc} = 14.7 psia, T_{sc}= 520 R, smooth pipe is considered.

Segment	Qg, mmscfd	P_1, psig	Dia., in	T, °C	L, km	Re	f(from curve)	P_2,psia (calculated)	P_2, psia (measured)	APRE
1	294.95	805	30	14.903	40	9396816.6	0.0128	795.0	814.7	2.47
2	70.06	800	30	16.472	18.5	2234579	0.01	814.2	787.7	3.254
3	168.71	773	20	16.472	64	8071898.5	0.012	685.4	539.7	21.26

Table B.4: (b) Pressure calculation using Panhandle A Equation. (Date: 01/02/2008)
S.G. =0.57595, Z_{av} = 0.889, P_{sc} = 14.7 psia, T_{sc}= 520 R, Gas viscosity = 0.0110125 cp.

Segment	Qsc, mmscfd	P_1, psig	Dia, inch	T,°C	L, km	P_2,psia (calculated)	P_2, psia (from daily report)	APRE
1	321.3	777	30	14.9039	40	1007.44	784.7	22.1096
2	70.08	770	30	16.4723	18.5	746.673	752.7	5.09284
3	210.031	738	20	16.4723	64	2699.81	474.7	70.935

Table B.4: (c) Pressure calculation using Panhandle B Equation. (Date: 01/02/2008)
S.G. =0.57595, Z_{av} =0 .889, P_{sc} = 14.7 psia, T_{sc}= 520 R, Gas viscosity = 0.0110125 cp
Smooth pipe is considered for Re calculation.

Segment	Q_{sc}, mmscfd	P_1, psig	Dia. inch	T, °C	L, km	Re	P_2,psia (calculated)	P_2, psia (measured)	APRE
1	321.3	777	30	14.9	40	10236301.7	791.7	784.7	0.884
2	70.08	770	30	16.4	18.5	2232679.81	784.7	752.7	1.12E-10
3	210.031	738	20	16.4	64	10037071.3	752.7	474.7	4.25

Note: This equation gave close result to Daily Production Report.

Table B.4: (d) Pressure calculation using iteration. (Date: 01/02/2008)
V_{sl} and V_{sg} calculation for segment 1 (Diameter = 28.874 inch)
Q_l = 1.05303 m^3/s, w_l insitu = 835.054957 kg/s, Gas Density = 127.8384833 kg/m^3,
Liquid Density = 793 kg/m^3, Pin = 777 psig, Pipe Length, L = 131233.6 ft,
ΔL=L/n= 131233.6/1 ft = 131233.6 ft, Flow = 321.3 MMSCFD

No	$P_{i+1(G)}$ psia	P_{avg}, psia	Gas flow, insitu (m3/s)	W_g, insitu (kg/s)	liq. Flow, insitu (m3/s)	V_{sl},m/s	q_G,m3/s	V_{sg}, m/s
1	529.23	660.46	2.32	296.55	0.0232	0.04027	2.29	3.987
2			0.000125	0.0160063	1.25E-06	2.17E-06	0.00012	0.00021
3	589.06	690.38	2.212	283.705	0.022192	0.03852	2.197	3.814
4	2293795	11469373.9	0.0001336	0.01707	1.34E-06	2.32E-06	0.000132	0.000229
5	575.503	683.6015	2.24125	286.518	0.022412	0.0389	2.21884	3.852
6	2327112	11635957.9	0.0001317	0.01684	1.32E-06	2.28E-06	0.00013	0.000226
7	578.6	685.15	2.2362	285.87	0.022362	0.03882	2.214	3.8433
8	2319429	11597542.4	0.000132	0.0169	1.32E-06	2.29E-06	0.00013	0.000227
9	577.9	684.8	2.2374	286.016	0.022373	0.03884	2.214954	3.845
10	2321162	11606206.4	0.000132	0.01688	1.32E-06	2.29E-06	0.00013	0.000226

Table B.4: (e) Pressure calculation using iteration. (Date: 01/02/2008) contd.
$(dp/dl)_{sl}$, $(dp/dl)_{sg}$ and dp/dl_f calculation for segment 1
Q_L = 1.05303 m^3/s, w_l insitu = 835.054957 kg/s, Gas Density = 127.8384833 kg/m^3
Liquid Density = 793 kg/m^3

Iteration	Re_{SL}	Re_{SG}	F_l	$(dp/dl)_{sl}$	f_g	$(dp/dl)_{sg}$	X	H_L	Φ_l^2	dp/dl
				psia/ft		psia/ft				psia/f
1	12.496	23900	1.28	4.49	0.00612	33.94	0.3637	0.13	41.55	186.5
2	0.0006	1.29	23721	0.00024	12.4028	0.0002	1.10027	0.2418	6.37	0.001
3	11.955	22865	1.338	4.29606	0.00617	31.339	0.37024	0.1314	40.7	174.8
4	0.0007	1.376	22233	0.00025	11.625	0.0002	1.10021	0.2418	6.37	0.001
5	12.073	23091	1.325	4.33866	0.00616	31.901	0.36878	0.131	40.9	177.4
6	0.0007	1.356	22556	0.00025	11.794	0.0002	1.10021	0.2418	6.37	0.001
7	12.046	23040	1.328	4.32885	0.0061	31.77	0.36911	0.131	40.84	176.8
8	0.0007	1.361	22482	0.00025	11.755	0.0002	1.10021	0.2418	6.37	0.001
9	12.052	23051	1.327	4.33106	0.0061	31.8	0.36904	0.131	40.85	176.9
10	0.0007	1.36	22499	0.00025	11.763	0.0002	1.10021	0.2418	6.37	0.001
11	12.496	23900	1.28	4.49063	0.0061	33.94	0.36374	0.13	41.54	186.5

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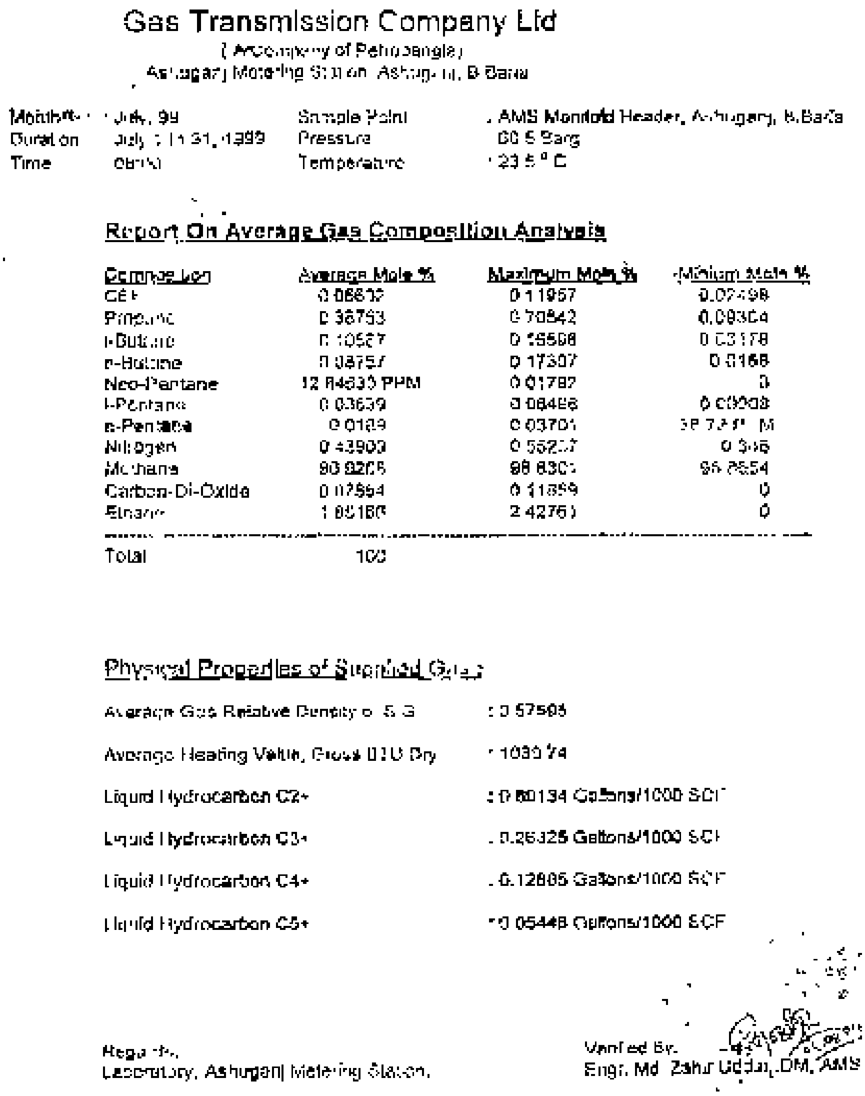
Appendix C
Various Reports



Gas Transmission Company Ltd
(A Company of Petrobangla)
Ashuganj Metering Station, Ashuganj, B.Baria

Month	July, 99	Sample Point	AMS Manifold Header, Ashuganj, B.Baria
Duration	July 1 to 31, 1999	Pressure	60.5 Barg
Time	[illegible]	Temperature	23.5 °C

Report On Average Gas Composition Analysis

Components	Average Mole %	Maximum Mole %	Minimum Mole %
C6+	0.06632	0.11957	0.07498
Propane	0.36753	0.70542	0.09304
i-Butane	0.10527	0.15508	0.03178
n-Butane	0.08757	0.17307	0.0168
Neo-Pentane	12.84630 PPM	0.01787	0
i-Pentane	0.03639	0.08486	0.00908
n-Pentane	0.0189	0.03701	[illegible] PPM
Nitrogen	0.43900	0.55207	0.306
Methane	96.9205	98.8301	95.7554
Carbon-Di-Oxide	0.07564	0.11859	0
Ethane	1.85180	2.42761	0
Total	100		

Physical Properties of Supplied Gas

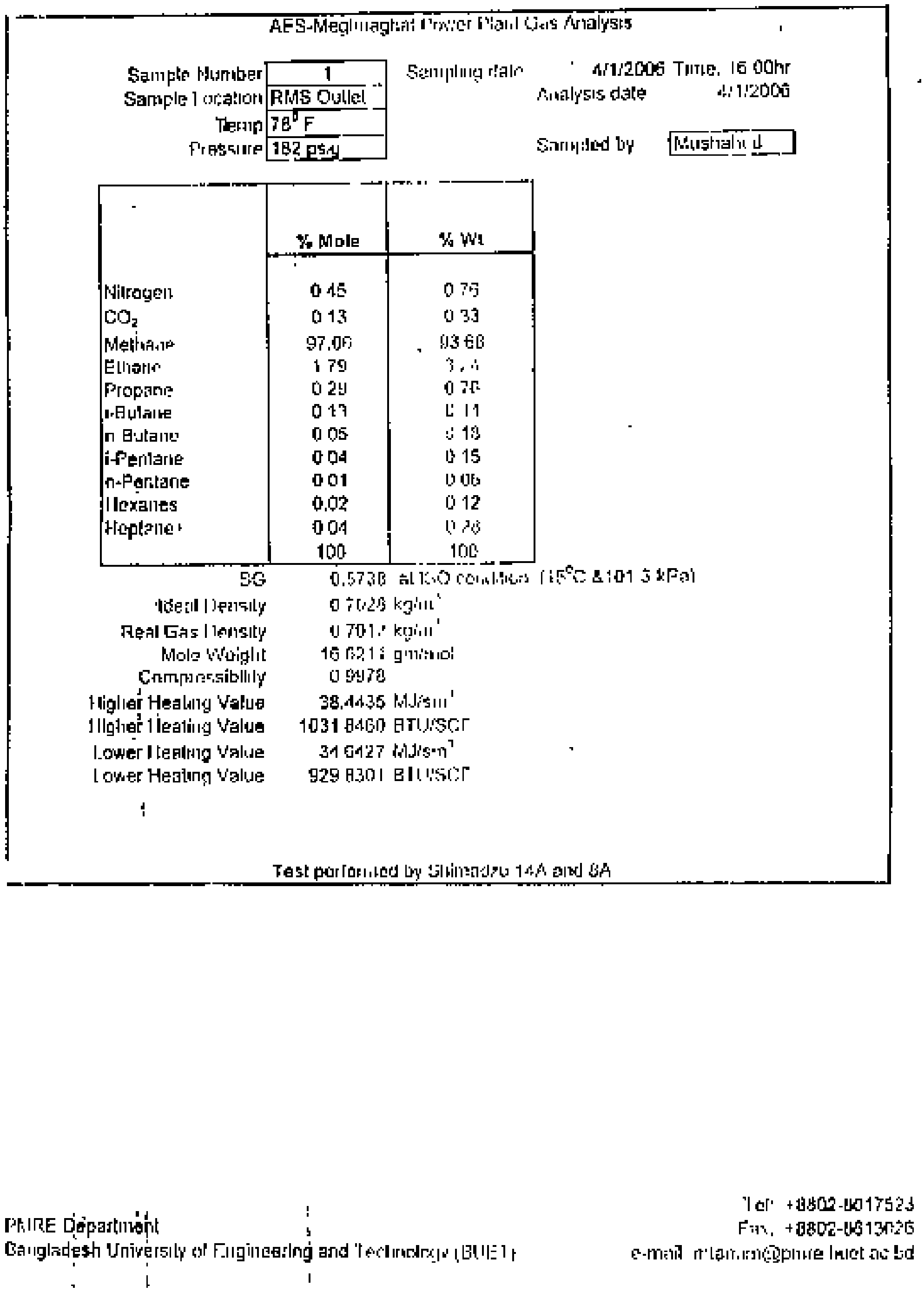
Average Gas Relative Density or S.G	: 0.57505
Average Heating Value, Gross BTU Dry	: 1030.74
Liquid Hydrocarbon C2+	: 0.60134 Gallons/1000 SCF
Liquid Hydrocarbon C3+	: 0.26325 Gallons/1000 SCF
Liquid Hydrocarbon C4+	: 0.12805 Gallons/1000 SCF
Liquid Hydrocarbon C5+	: 0.05448 Gallons/1000 SCF

Regards,
Laboratory, Ashuganj Metering Station.

Verified By:
Engr. Md. Zahir Uddin, DM, AMS

Figure C.1: Chromatographic Gas Analysis of AGMS.

AES-Meghnaghat Power Plant Gas Analysis



AES-Meghnaghat Power Plant Gas Analysis

Sample Number	1
Sample Location	RMS Outlet
Temp	78^0 F
Pressure	182 psig

Sampling date 4/1/2006 Time. 16 00hr

Analysis date 4/1/2006

Sampled by Mushahid

	% Mole	% Wt
Nitrogen	0.45	0.76
CO_2	0.13	0.33
Methane	97.06	93.68
Ethane	1.79	[illegible]
Propane	0.29	0.76
i-Butane	0.13	0.11
n Butane	0.05	0.18
i-Pentane	0.04	0.15
n-Pentane	0.01	0.06
Hexanes	0.02	0.12
Heptane+	0.04	0.26
	100	100

SG	0.5738	at ISO condition (15°C & 101.3 kPa)
Ideal Density	0.7028	kg/m^3
Real Gas Density	0.7017	kg/m^3
Mole Weight	16.6211	gm/mol
Compressibility	0.9978	
Higher Heating Value	38.4435	MJ/sm^3
Higher Heating Value	1031.8460	BTU/SCF
Lower Heating Value	34.6427	MJ/sm^3
Lower Heating Value	929.8301	BTU/SCF

Test performed by Shimadzu 14A and 8A

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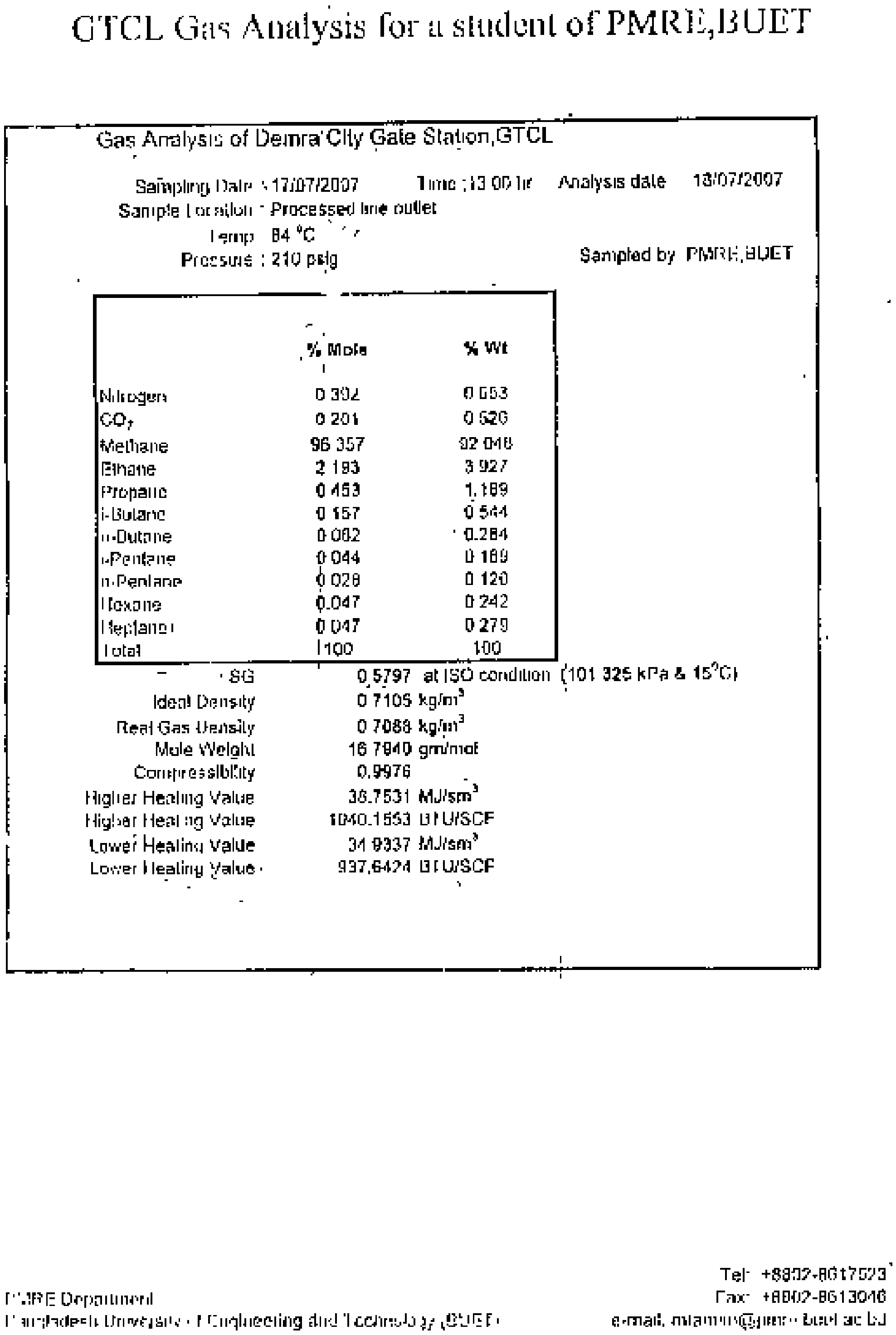
Figure C.2: Chromatographic Gas Analysis of AES-Meghnaghat Power Plant. (It is taken as BKB chromatographic gas analysis)



GTCL Gas Analysis for a student of PMRE,BUET

Gas Analysis of Demra City Gate Station,GTCL

Sampling Date : 17/07/2007 Time : 13 00 hr Analysis date 18/07/2007
Sample Location : Processed line outlet
Temp 84 °C
Pressure : 210 psig
Sampled by PMRE,BUET

	% Mole	% Wt
Nitrogen	0.302	0.653
CO_2	0.201	0.520
Methane	96.357	92.048
Ethane	2.193	3.927
Propane	0.453	1.189
i-Butane	0.157	0.544
n-Butane	0.082	0.284
i-Pentane	0.044	0.189
n-Pentane	0.028	0.120
Hexane	0.047	0.242
Heptane+	0.047	0.279
Total	100	100

SG	0.5797 at ISO condition (101.325 kPa & 15°C)
Ideal Density	0.7105 kg/m^3
Real Gas Density	0.7088 kg/m^3
Mole Weight	16.7940 gm/mol
Compressibility	0.9976
Higher Heating Value	38.7531 MJ/sm^3
Higher Heating Value	1040.1553 BTU/SCF
Lower Heating Value	34.9337 MJ/sm^3
Lower Heating Value	937.6424 BTU/SCF

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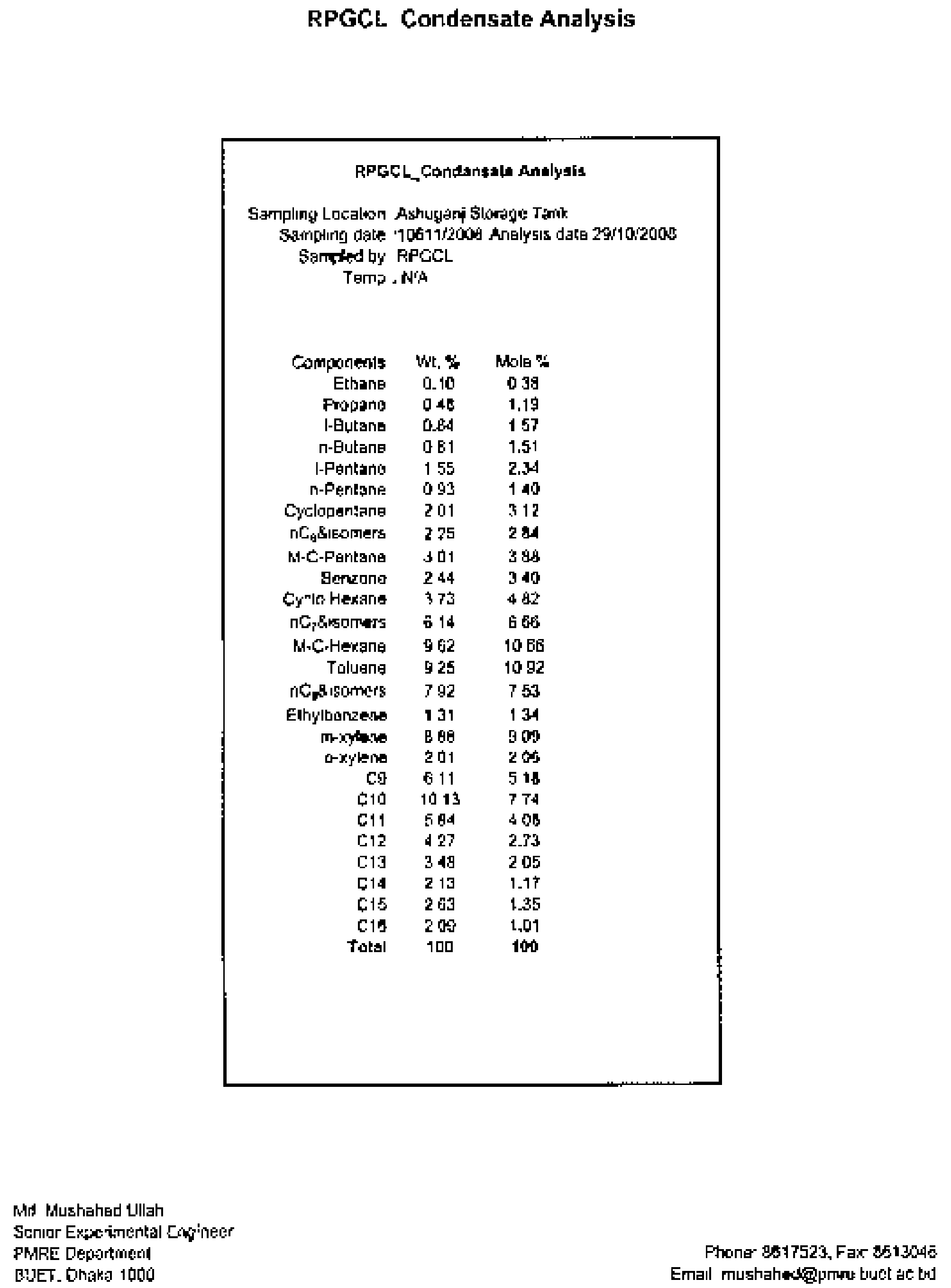
Figure C.3: Chromatographic Gas Analysis of Demra CGS.



RPGCL Condensate Analysis

RPGCL_Condensate Analysis

Sampling Location Ashuganj Storage Tank
Sampling date '10611/2008 Analysis date 29/10/2008
Sampled by RPGCL
Temp . N/A

Components	Wt. %	Mole %
Ethane	0.10	0 38
Propane	0 46	1.19
I-Butane	0.84	1 57
n-Butane	0 81	1.51
I-Pentane	1 55	2.34
n-Pentane	0 93	1 40
Cyclopentane	2 01	3 12
nC_6&isomers	2 75	2 84
M-C-Pentane	3 01	3 84
Benzene	2 44	3 40
Cyclo Hexane	3 73	4 82
nC_7&isomers	6 14	6 66
M-C-Hexane	9 62	10 66
Toluene	9 25	10 92
nC_8&isomers	7 92	7 53
Ethylbenzene	1 31	1 34
m-xylene	8 88	9 09
o-xylene	2 01	2 06
C9	6 11	5 18
C10	10 13	7 74
C11	5 84	4 06
C12	4 27	2.73
C13	3 48	2 05
C14	2 13	1.17
C15	2 63	1.35
C16	2 09	1.01
Total	100	100

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Figure C.4: Chromatographic Condensate Analysis of RPGCL.

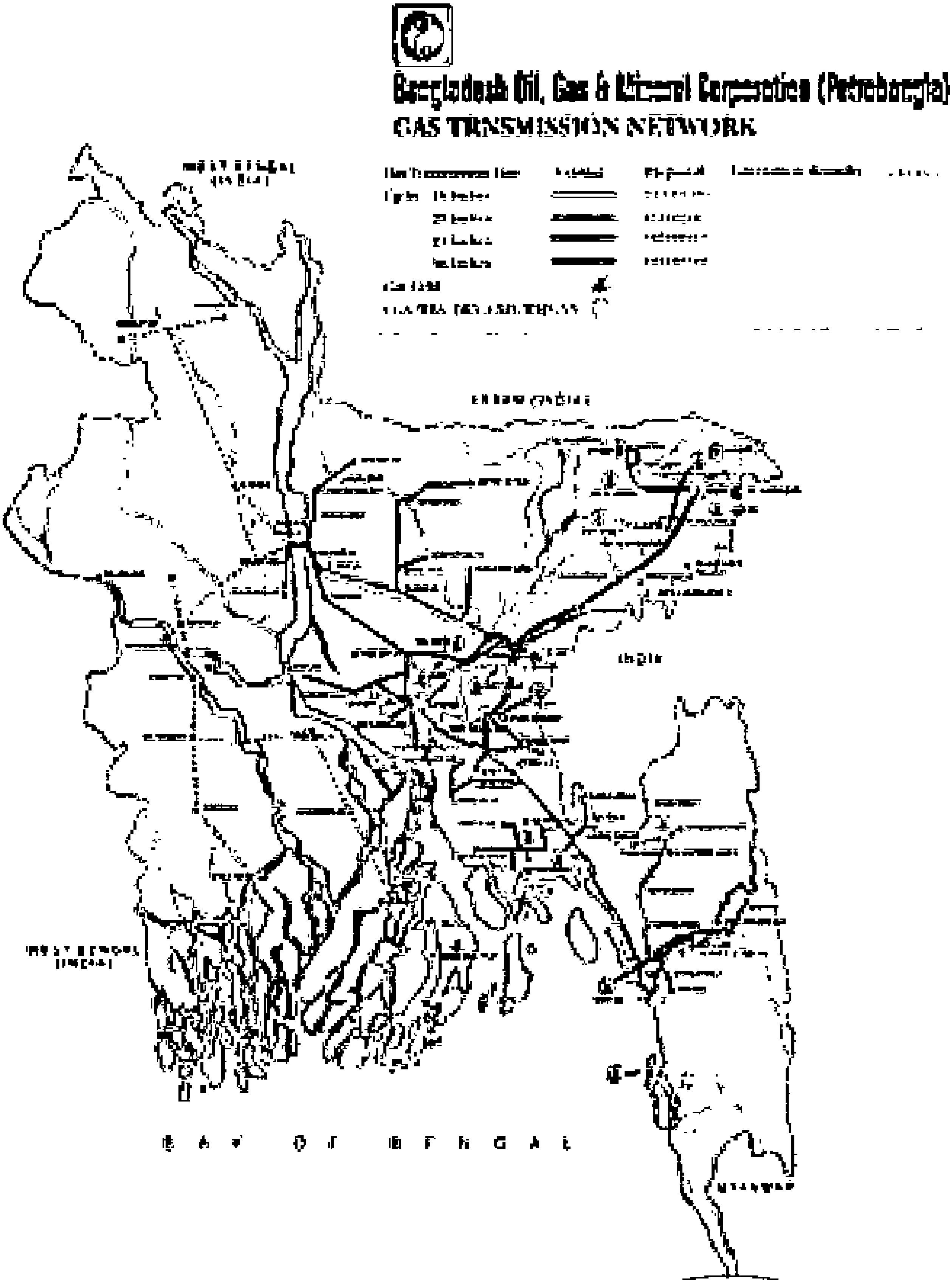

Figure C.5: Gas Transmission Network of Bangladesh.