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A Review on Tubing Performance Relationship by Hagedorn & Brown Model.

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Abstract: Three main components influence the pressure drop inside a vertical pipe when the fluid flow is multiphase are Elevation,Friction and Acceleration.Many authors evaluated the available correlations for multiphase flow in vertical pipes. These studies showed that Hagedorn and Brown correlation is one of the best correlations available for pressure drop prediction over a wide range of parameters such as pipe size, flow rates, etc.Hagedorn& Brown correlation used Griffith's correlation for liquid holdup in bubble flow conditions. Hence this method is selected for calculating the pressure drop in our study.In this study, both Hagedorn& Brown liquid holdup correlation is revised by using various flow rates ranging from 100 bbl/d to 5000 bbl/d with varying tubing sizes(between 1'in to 4' in). In comparison with the calculated pressure drops, Hagedorn& Brown correlation gave consistent results with observed values.

I. INTRODUCTION

Multiphase flow in pipes is defined as the concurrent movement of free gases and liquids in the pipes. Flow may be in any direction. The gas and liquid may exist as a homogeneous mixture, or the liquid may be in slugs with the gas pushing behind it. The liquid and gas may also flow parallel to each other, or other combinations of flow patterns may be present. The gas may be flowing with two liquids (normally oil and water), and the possibility exists that the two liquids may be emulsified.

The multiphase flow problem can be divided into four categories:

A. Vertical multiphase flow

Vertical multiphase flow is found in practically every tubing string used inthe production of oil. It is necessary to be able to predict a vertical multiphase flow pressure traverse in order to correctly select completion strings, predict flow rates, and design artificial lift installations. Most of the progress towards a solution of the problem has been made since the publication of Poettmann and Carpenter's paper in 1952. Most of the approaches use some form of the general energy equation.

The pressure gradient (or rate of change in pressure with respect to unit of flow length) for vertical multiphase flow is the sum of three contributing factors: hydrostatic pressure gradient, friction pressure gradient, and acceleration pressure gradient. The effects of chemical reactions between phases are neglected; however, such factors as viscosity, surface tension, density, etc. are included.

B. Multiphase Flow Correlations

Bril and Mukherjee (1999) placed the empirical multiphase correlations into three categories:

Category A: No slip, no flow pattern considered. Gas and liquid phases are assumed to have the same flow velocity. The calculation procedure is the same for all flow patterns, which requires only one correlation to calculate the friction factor for the "homogeneous" gas-liquid mixture.Category B: Slip considered, no flow pattern considered. Correlations are developed to calculate both liquid holdup and friction factor and used for all flow patterns.Category C: Both slip effect and flow pattern are considered. The flow pattern is predicted at first by using sets of equations of correlations, and then according to which flow pattern was predicted at the first step, corresponding different sets of equations are chosen to estimate liquid holdup and friction factor.

There are numerous correlations that give excellent results depending upon the ranges of flow conditions. Based on comparisons made by Lawson and Brill, the following methods and the order in which they will be discussed are those of Hagedorn and Brown, Duns and Ros, Orkiszewski and Beggs and Brill. All four of these methods represent generalized correlations to take care of all pipe sizes, flid properties and flow rates. Also of significance is the correlation of Govier and Aziz.

C. Poettman and Carpenter (1952) Correlation

Based on field data from 49 flowing and gas-lift wells, Poettman and Carpenter (1952) correlated the friction factor of multiphase flow with the product of the inside diameter of tubing and the mass velocity of the mixture flowing through the pipe. It should be noted that this dimensional product corresponds to the numerator of the Reynolds number.

D. Baxendell and Thomas (1961) Correlation

Due to the unsuccessful extrapolation of the Poettman and Carpenter correlation from low flow rates to high flow rates, Baxendell and Thomas (1961) suggested some modifications to the Poettman and Carpenter correlation to fit smoothly to the high-rate correlation derived from Cia, Shell de Venezuela's La Paz field in Venezuela.

E. Fancher and Brown (1963) Correlation

Fancher and Brown (1963) conducted a series of experiments on a 8000-ft experimental field well for flow rates ranging from 75 to 936 bbl/day at various gas-liquid ratios from 105 to 9433 scf/bbl. They found obvious deviation from the Poettman and Carpenter correlation for a certain range of flow rates and gas-liquid ratios. In order to fit the experimental data, they treated the gas-liquid ratio as an additional parameter and divided the range into three parts. For each range, a correlation was developed between friction factor and the numerator of the Reynolds number.

F. Gray (1978) Correlation

This method takes the effects of liquids (condensate and/or free water) in gas well production into account for pressure gradient calculations and was developed especially for two-phase flow in vertical gas wells (Gray, 1978). It is recommended by the API in their manual for subsurface controlled safety valve sizing computer programs. The Gray correlation uses no-slip holdup and two dimensionless numbers, which are similar to the Velocity 11 Number and Pipe Diameter Number defined by Duns and Ros (1963), to calculate the liquid holdup. A total of 108 selected well test data sets were used in developing this correlation, of which, 88 data sets were obtained from wells reported to produce free liquids. Additionally, another 65 data sets were randomly selected for statistical control purposes. Prediction results were compared to observations from both test and control data sets. The results were superior to the predictions made by conventional dry gas models and for pressure gradient prediction, the average bias is −0.35% and average standard deviation is 5.2% (Gray, 1978). Yet, the accuracy was stated to be questionable when, Mixture velocity, vm> 50 ft/sec; Pipe diameter, d > 3.5 in (nominal); Liquid condensate and gas ratio > 50 bbl/Mmscf; and, Water and gas ratio > 5 bbl/Mmscf.

G. Dukler et al. (1969) Correlation:

In this thesis study, the Eaton, Knowles and Silberbrg (1967) correlation was used for liquid holdup calculations and the Flanigan (1958) correlation for elevational pressure gradient calculation. They were combined with the Dukler et al. (1969) correlation calculation procedure for horizontal flow pressure loss calculation. For the Dukler et al. correlation, a total of approximately 400 horizontal flow experimental data points were utilized to established a graphical relationship between friction factor and Reynolds number, then friction pressure drop in two-phase flow would be calculated through similarity analysis approach (Dukler, Wicks and Cleveland, 1964). Liquid holdup was obtained through a trial and error calculation procedure (Dukler et al., 1969). Data to develop the Eaton, Knowles and Silberbrg (1967) correlation were taken from a horizontal multiphase test unit, consisting of two 1700-ft test lines with diameters of 2 and 12 4 in.

H. Flow condition ranges for the test are as follows (Eaton, Knowles and Silberbrg, 1967)

1) *Liquid rates:* 50 to 2500 bbl/day for the 2-in line; 50 to 5500 bbl/day for the 4-in line.
2) *Gas-liquid ratio*: 0 to 132000 scf/bbl. for the 50 bbl/day liquid rate; a narrower range for the higher liquid rates. The physical properties of test fluids can be summarized as,
3) *Gas:* natural gas with S.G. (specific gravity) of 0.6111 and viscosity of 0.012 cp @ 80F.
4) *Water:* S.G. of 10.01, surface tension of 66.0 dynes/cm, viscosity of 1.01 cp @ 80F.
5) *Crude:* S.G. of 0.865, surface tension of 30.0 dynes/cm, viscosity of 13.50 cp @ 80F.
6) *Distillate:* S.G. of 0.77, surface tension of 26.0 dynes/cm, viscosity of 3.50 cp @ 80F.

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7) Based on studies of small amounts of condensate in gas lines, Flanigan (1958) developed a liquid holdup correlation to account for the hydrostatic pressure difference in upward inclined flow. The Flanigan correlation is utilized in this study to calculate the elevation part of total pressure gradient. As for downhill flow, the elevation gradient is neglected.

I. Duns and Ros (1963) Correlation

Based on extensive laboratory experiments, the Duns and Ros (1963) method is expected to be more general and applicable to the full range of field operating conditions, including tubing and annular flow for a wide range of oil and gas mixtures with varying water cuts. About 4000 two-phase flow tests, comprising some 20,000 data points, were carried out on a vertical string, consisting of an inflow section with a length between 98 and 198 ft, a 32.8 ft long measuring section and a 6.6 ft long outflow section (Ros, 1961). The test configurations and flow conditions can be summarized as below,

1) Pipe diameters ranged from 1.26 to 5.6 in; outer diameter of annulus was 5.6 in and inner diameter ranged from 2.37 to 3.54 in. 13
2) Fluids used in the tests (besides water):
3) Lubricating oil: density of 53.1 to 58.5 lbm/f t3, surface tension of 28.1 to 33.8 dynes/cm and viscosity of 5.62 to 315.8 cp. Gas oil: density of 51.6 to 52.3 lbm/f t3, surface tension of 27 to 28 dynes/cm and viscosity of 3.312 to 4.101 cp.- Mineral spirit: density of 48.7 lbm/f t3, surface tension of 24.5 dynes/cm and viscosity of 0.96 cp.- Liquid superficial velocity: 0 to 328.1 f t/sec;- Gas superficial velocity: 0 to 10.5 f t/sec.

J. Accuracy and Applicability

The validity of the correlation was divided into three regions (Duns and Ros, 1963),

1) Region 1: liquid phase is the continuous phase
a) Bubble flow, plug flow and part of froth-flow regime;
b) Standard deviation of the per cent errors is 3% for dry oil and gas mixture which is equal to the measuring accuracy.
2) Region 2: phases of liquid and gas alternate
a) Slug flow and the rest of froth-flow regime;
b) Standard deviation of 8% for dry oil and gas mixtures.
3) Region 3: gas phase is continuous
a) Mist flow;
b) Standard deviation of 6% after refinements.

The discrepancy can amount to up to 10% in Region 1 and Region 2 with wet mixtures containing less than 10% of water.

K. Orkiszewski (1967) Correlation:

Orkiszewski (1967) tested several existing correlations and found none of them proved accurate over the entire range of conditions of available data. He then presented this correlation for vertical two-phase flow, which is an extension of the Griffith and Wallis (1961) correlation. A new correlation for slug flow regime was developed using Hagedorn and Brown (1965) experimental data, and the Griffith and Wallis method was selected for bubble flow regime and Duns and Ros correlation was chosen to deal with mist flow regime. The pressure drop prediction precision of this method was verified by comparison against 148 measured pressure drops,

1) Standard deviation about 10.0% for two-phase pressure drop prediction in flowing and gas-lift production vertical wells over a wide range of well conditions;
2) Four flow regimes were considered: bubble, slug, annular-slug transition, and annular mist;
3) Unlike most other methods, liquid holdup was derived from observed physical phenomena.

L. Beggs and Brill (1973) Correlation

An experimental apparatus was designed and built to investigate the effect of pipe inclination angle on liquid holdup and pressure loss of gas-liquid flow in inclined pipes (Beggs and Brill, 1973). A total of 584 average liquid holdup and pressure drop measurements, from which this correlation was developed, were taken in transparent acrylic pipes, which could be inclined at any angle from the horizontal. The parameters studied and their range of variation were,

1) Gas (air) flow rate: 0 to 300 Mscf/day;

2) Liquid (water) flow rate: 0 to 1029 bbl/day;
3) Average system pressure: 35 to 95 psia;
4) Pipe diameter: 1.0 to 1.5 in;
5) Liquid holdup: 0 to 0.870;
6) Pressure gradient: 0 to 0.80 psi/f t;
7) Inclination angle: -90'to +90'; and,
8) Different flow patterns were observed.

M. Accuracy and Applicability

Comparing with all the tests results, average percent error for liquid holdup prediction was -0.28% with a standard deviation of 7.89%. And for the pressure gradient prediction, the values were 1.11% and 9.30%:

1) Based on air-water and small diameter pipe;
2) Valid for all inclination angles;
3) Only consider horizontal flow patterns; and,
4) Inclination angle correction made for each flow pattern.

N. Mukherjee and Brill (1985) Correlation

For this experimental work, each leg of the U-shape test section was 56 ft long with 22 ft. entrance lengths followed by 32 ft. long test sections to simulate both uphill and downhill flow at the angle of 0° to 90° from horizontal (Mukherjee and Brill, 1985). The fluids used were Air-Kerosene or air-lube oil. During the test, each liquid flow rate was set at first, and then a series of different gas flow rates were introduced. For each gas and liquid flow rate, flow patterns were observed and holdup recorders and pressure gauges were activated. Approximately 1000 pressure-drop measurements and more than 1500 liquid holdup measurements were taken for a broad range of gas and liquid flow rates (Brill and Mukherjee, 1999). 16 Pressure loss calculation results from this method were compared with the observed horizontal and upward flow data (Air-Kerosene only) with an average percent error of -0.422% and a standard deviation of 17.75% (Mukherjee, 1979). In addition, pressure loss predictions were compared with 14 pipeline data from Prudhoe Bay Field (horizontal flow) and 130 offshore well data from North Sea data (vertical flow) with average percent errors of -9.5% and -3.3%, and standard deviations of 14.67% and 9.7%, respectively.

O. Aziz, Govier and Fogarasi (1972) Correlation:

Based on flow mechanism instead of experiment or field data, Aziz, Govier and Fogarasi (1972) proposed a pressure drop calculation scheme, which can be seen as a precursor of modern mechanistic models. Transition criteria and pressure gradient correlations for bubble, slug, froth (transition), annular mist flow regimes were described. Field data from 48 wells were used to measure the accuracy of the proposed method. The calculation results were compared with other methods and the absolute error was about the same as the Orkiszewski (1967) method but more favorable than the Hagedorn and Brown (1965) and Duns and Ros (1963) methods.

P. Uses of multiphase flow pressure loss calculations in petroleum engineering:

The application of multiphase flow correlations to predict pressure loss in pipes is extremely important to the petroleum industry. Some of the uses are:

1) *Natural flow:* When fluids are produced from the formation into the vertical tubing and through horizontal surface flow lines, energy is dissipated. To prolong the flowing life of this well as long as possible, there is a need to minimize energy loss by proper design of the flow string.
2) *Slim-hole completions:* For ultra-slim-hole completions, minimum casing size is installed to minimize completion costs. The casing size limits the maximum tubing size for a particular well. If the p5roducing mechanism and the water cut are known, the maximum flow rate through tubing of a given size can be calculated with multiphase flow equations.
3) *Dewatering gas wells.:*The selection of a siphon string to unload water from gas wells, or of the production string for a gas well making water, is most important. Pressure calculations must be made to determine the proper size of the vertical tube.
4) *Artificial lift installations.*The flowing pressure loss in vertical tubes is needed for the proper design of most artificial lift installations. Knowledge about such losses is a particular requirement for gas-lift installations where additional gas is injected

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in the tubing. Pressure loss is also important in artificial lift installations where the pump is set far off bottom and the well is producing with a high gas-oil ratio because pressure loss may be significant in that section of the vertical pipe below the pump.

5) *Gathering and separation systems:*In centralized gathering and separation systems it is necessary to transport gas-liquid mixtures for relatively long distances. Correct sizing of the horizontal pipe used in these systems is important to prevent high pressure losses in the systems.
6) *Sizing surface flow lines:*The sizing of surface flow lines for oil production is extremely important in designing for maximum allowable production. The size of the surface flow line from the wellhead to the separator combined with separator pressure establishes the flowing wellhead pressure. The flowing wellhead pressure controls the flowing bottomholepressure which, in turn, controls the productive capacity of the well.
7) *Sizing of transmission lines:* prediction of pressure losses is important in the sizing of large transmission lines containing a liquid phase.
8) *Sizing of gas lines.:*Pressure loss calculations must be made when gas lines where glycol or some other* chemical is being injected to prevent freezing are designed.
9) *Tubing design in deviated wells:* The design of tubing string; for directionally-drilled wells is becoming more and more important as additional offshore wells are drilled.
10) *Surface design for inclined flow.:*The calculation of pressure losses for sizing of surface flow lines and transmission lines for inclined flow over hilly terrain, and for offshore-to-onshore facilities, is a necessity.
11) *Heat exchanger design:*In refineries and chemical plants two-phase mixtures of petroleum fractions sometimes circulate through heat exchangers. The design of the heat exchangers involves two-phase pressure-drop correlations.
12) *Condensate line design.:*Mixtures of partially- condensed vapors flowing through condensate lines in steam and refrigeration plants are in two-phase flow. The design of these lines must take into account the additional pressure loss caused by the existence of the liquid phase.There are other uses for multiphase flow calculations. All of the mentioned applications point out the fact that an economic problem is involved in the optimization of pipe sizes for vertical, horizontal, and inclined flow.There are numerous correlations that give excellent results depending upon the ranges of flow conditions. Based on comparisons made by Lawson and Brill, the following methods and the order in which they will be discussed are those of Hagedorn and Brown, Duns and Ros, Orkiszewski and Beggs and Brill. All four of these methods represent generalized correlations to take care of all pipe sizes, fluid properties and flow rates. Also of significance is the correlation of Govier and Aziz.This section will introduce the procedure for vertical multiphase flow problem , offer a brief historical review, and discuss one of the best correlation (Hagedorn& Brown correlation).

Q. Generalized correlation of Hagedorn and Brown:

An effort was made by Hagedorn and Brown[21] to determine a generalized correlation which would include all practical ranges of flow rates, a wide range of gas- liquid ratios, all ordinarily used tubing sizes and the effects of fluid properties. Data were taken for pipe sizes ranging from 1 in. nominal to 4 in. nominal tubing. The study included all of the prior work done by this team of investigators on the effects of liquid viscosity as discussed under "Limited Correlations”. A kinetic energy term was incorporated in the energy equation because it was considered to be very significant in small diameter pipes in the region near the surface where the fluid has a low density. Two adjustments were found necessary to improve this correlation. The Griffith correlation was used when bubble flow existed and the holdup was checked to make sure that it exceeded the holdup for no slippage and if not, the holdup for no slippage was used.

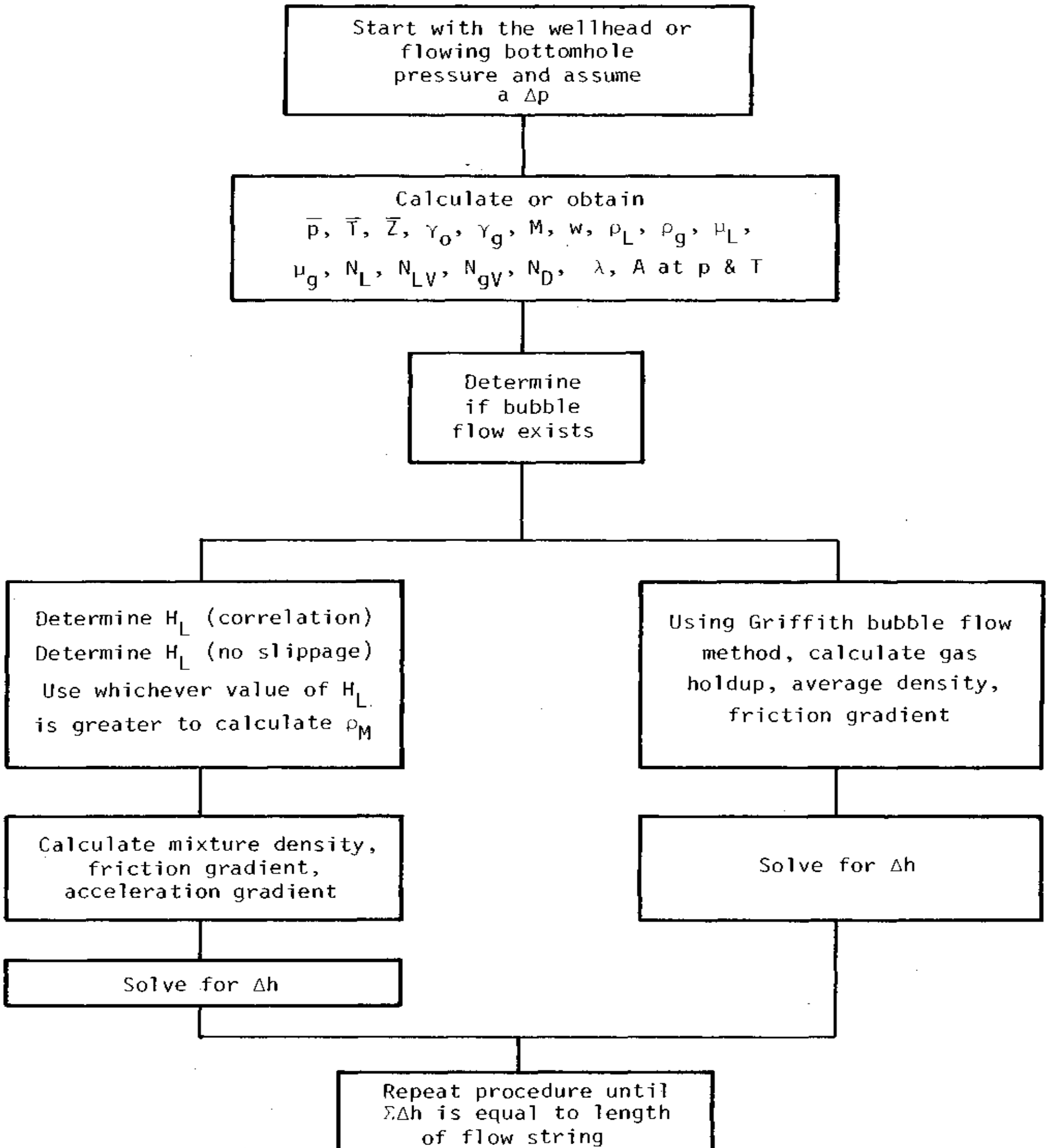

Fig.3.1.1 Flow diagram for generalized method of Hagedorn and Brown.

1) *Reference to Fig. 3.1.1* shows a generalized flow diagram which should help in the following discussions. These investigators also started with the general energy equation to obtain the pressure loss equation:

$$144\frac{\Delta p}{\Delta h} = \overline{\rho_m} + \frac{Fw^2}{2.9652*10^{11}d^5\overline{\rho_m}} + \overline{\rho_m}\,\Delta\frac{\left(\frac{v_m{}^2}{2g_c}\right)}{\Delta h} \qquad (3.1)$$

where

$$\bar{\rho}_m = \bar{\rho}_L H_L + \bar{\rho}_g(1 - H_L) \qquad (3.2)$$

The mixture viscosity was represented in the manner suggested by Arrhenius[28] and a Reynolds number for the two-phase mixture was defined by the equation:

$$(N_{Re})_{TP} = \frac{2.2*10^{-2}w}{(d)\left(\mu_L^{H_L}\right)\left(\mu_g^{(1-H_L)}\right)} \qquad (3.3)$$

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2) Referring to Eq. 3.3, if the limit is taken of the Reynolds number for the mixture as H_L —» 0, q_L —» 0, it reduces to a single-phase gas-flow equation, and if the limit is taken as H_L —» 1, qg —» 0, it reduces to the single-phase liquid- flow equation. The respective Reynolds numbers for all gas and all liquid reduce to

$$(N_{Re})_g = C_1 \frac{V_g \rho_g d}{\mu_g} \qquad (3.4)$$

$$(N_{Re})_L = C_1 \frac{V_g \rho_g d}{\mu_g} \qquad (3.5)$$

Using methods similar to those of Duns and Ros and Ros, Brown and Hagedorn showed that the liquid holdup (H_l) is principally related to four dimensionless parameters:

$$N_{LV} = V_{SL} \left(\frac{\rho_L}{g_\sigma}\right)^{\frac{1}{4}} \qquad (3.6)$$

$$N_{gV} = V_{Sg} \left(\frac{\rho_L}{g_\sigma}\right)^{\frac{1}{4}} \qquad (3.7)$$

$$N_d = d \left(\frac{\rho_{Lg}}{\sigma}\right)^{\frac{1}{2}} \qquad (3.8)$$

$$N_L = \mu_L \left(\frac{g}{\rho_{L\sigma^3}}\right)^{\frac{1}{4}} \qquad (3.9)$$

Converting to common oilfield units, these relations become:

$$N_{LV} = 1.938\, V_{SL} \left(\frac{\rho_L}{g_\sigma}\right)^{\frac{1}{4}}$$

$$N_{gV} = 1.938\, V_{Sg} \left(\frac{\rho_L}{g_\sigma}\right)^{\frac{1}{4}}$$

$$N_d = 120.872 d \left(\frac{\rho_{Lg}}{\sigma}\right)^{\frac{1}{2}}$$

$$N_L = 0.15726\, \mu_L \left(\frac{g}{\rho_{L\sigma^3}}\right)^{\frac{1}{4}}$$

Where;

d = Ft
P lb_m/cu ft
V_SL ft/sec
V_{s6} = ft/sec
PL = Cp
cr dynes/cm.

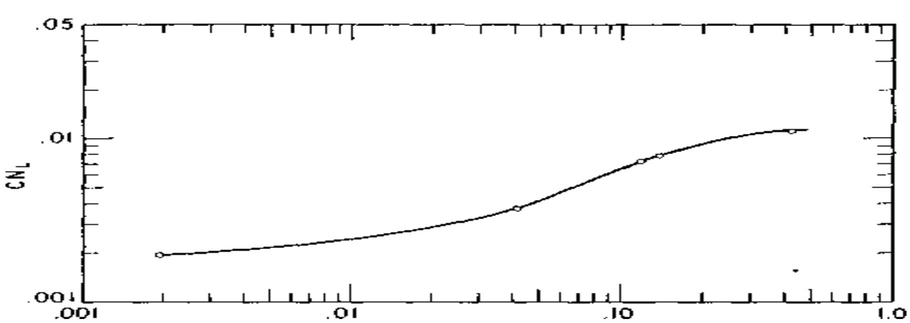

Fig. 3.1.2 Correlation for viscosity number coefficient C(after Hagedorn)

A regression analysis technique was used to relate the four dimensionless groups, as well as a pressure term, and the result is shown as Fig. 3.1.3.

It should be noted at this point that the Hagedorn and Brown holdup correlation is in fact a pseudoholdup correlation. Holdup was not actually measured but back-calculated from knowing the total pressure loss and using a friction factor obtained from a two-phase Reynolds number.

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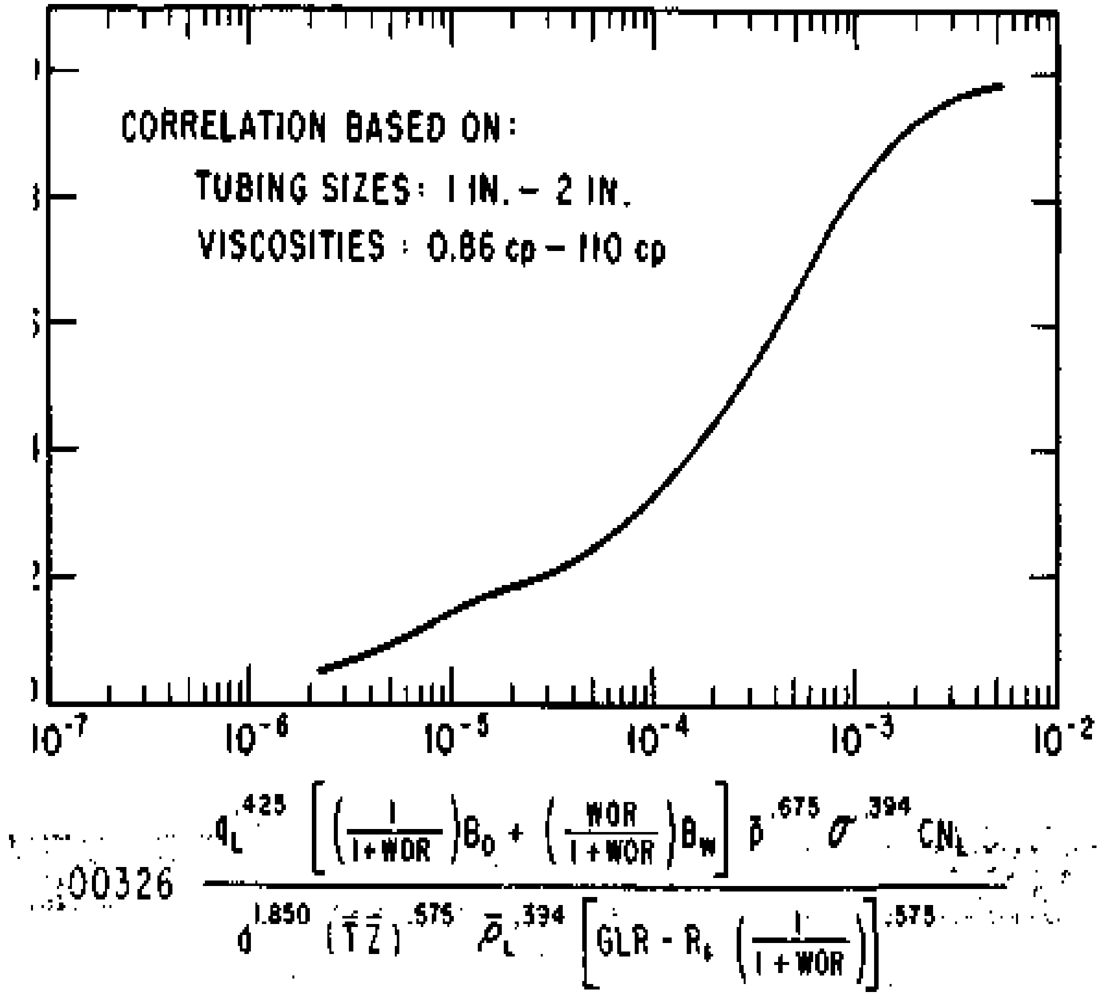


3.1.3 Holdup-factor correlation(after Hagedorn).

To account for the viscosity- of the liquid, the term **CN_L** was included in the numerator of the correlating function used for the abscissa of Fig. 3.1.3. A plot of N_L versus $CN_L$is noted in Fig. 2.56. Water was chosenasthe base curve; C was taken as 1.00 for water. This plot shows that for low values of the liquid viscosity, the viscosity has very little effect.

An additional factor was needed to properly account for the holdup because it was impossible to obtain one curve to account for the deviation at high gas rates in high viscosity crude oils.

This secondary correction factor OP) was plotted against values of the group of terms $N_{gv}N_L°\text{-}^{\mathbf{38}}/N_{\mathbf{d}}{}^{\mathbf{2}}\text{-}^{14}$ ***(Fig. 3.1.4).*** In most cases it will be found that T = 1.00.

The correlating function for liquid holdup can be expressed in field units as follows:

$$\left[\frac{CN_{LV}}{(CN_{gv})^{0.575}}\right]\left(\frac{P}{P_a}\right)^{0.10}\left(\frac{CN_L}{N_d}\right) = 0.00326\,\frac{q_L^{0.425}\left[\left(\frac{1}{1+W_{or}}\right)B_0+\left(\frac{W_{or}}{1+W_{or}}\right)B_w\right]\bar{P}^{0.675}\sigma^{0.394}CN_L}{d^{1.850}(\overline{TZ})^{0.575}\rho_L^{0.394}\left[GLR-R_S\left(\frac{1}{1+W_{or}}\right)\right]^{0.575}}$$

where:

q_L = liquid production rate, stb/d

d = pipe diameter, ft.

Z = average gas compressibility factor, dimensionless

T = average temperature, °R

_P = average pressure, psia

p_L = average liquid density, lb_m/cuft
GLR = gas-liquid ratio, scf/stb _
R_s = gas in solution at T and P, scf/stb
WOR = water-oil ratio, stbw/stbo.

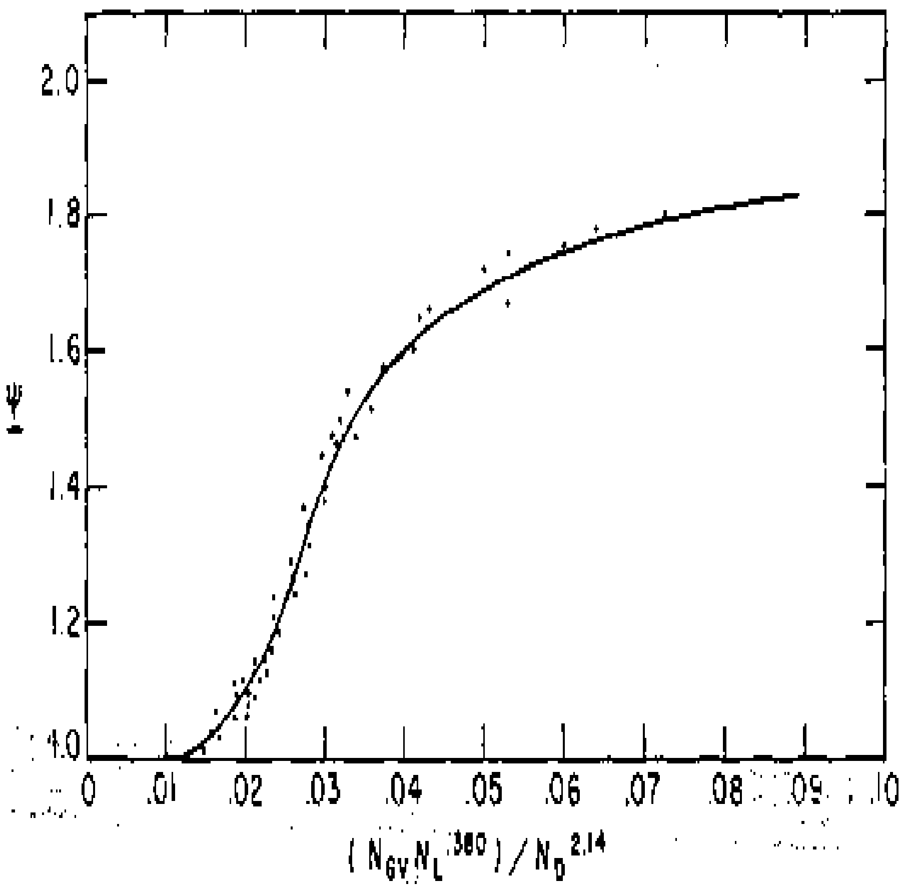


Fig. 3.1.4 Correlation for secondary correction factor(after Hagedorn).

Further work by Brill and Hagedorn has considerably improved this method. This work was performed in connection with better inclusion of the effects of holdup and slippage and the inclusion of the Griffith bubble flow correlation. Figs. 3.1.5 and 3.1.6 give an indication of the discrepancies that needed to be resolved. Fig. 3.1.5 presents a set of pressure traverses as calculated by Hagedorn and Brown for 100 bpd water in 3 in. tubing. There is a wide spread between the 0 and 50 scf/bbl gas-liquid ratio curve. It was thought that this was caused by an error in the original holdup correlation; but later it was found that the predicted holdup for low flow rates and low gas- liquid ratios was less than the holdup would be if there was no slippage. This discrepancy was more pronounced for large pipe sizes. Brill and Hagedornsuggested that a calculation be made for holdup from the Hagedorn and Brown correlation and that these results be compared to the holdup when no slippage is assumed. If the latter was found to be greater than the former, the H_L value excluding slippage was used.

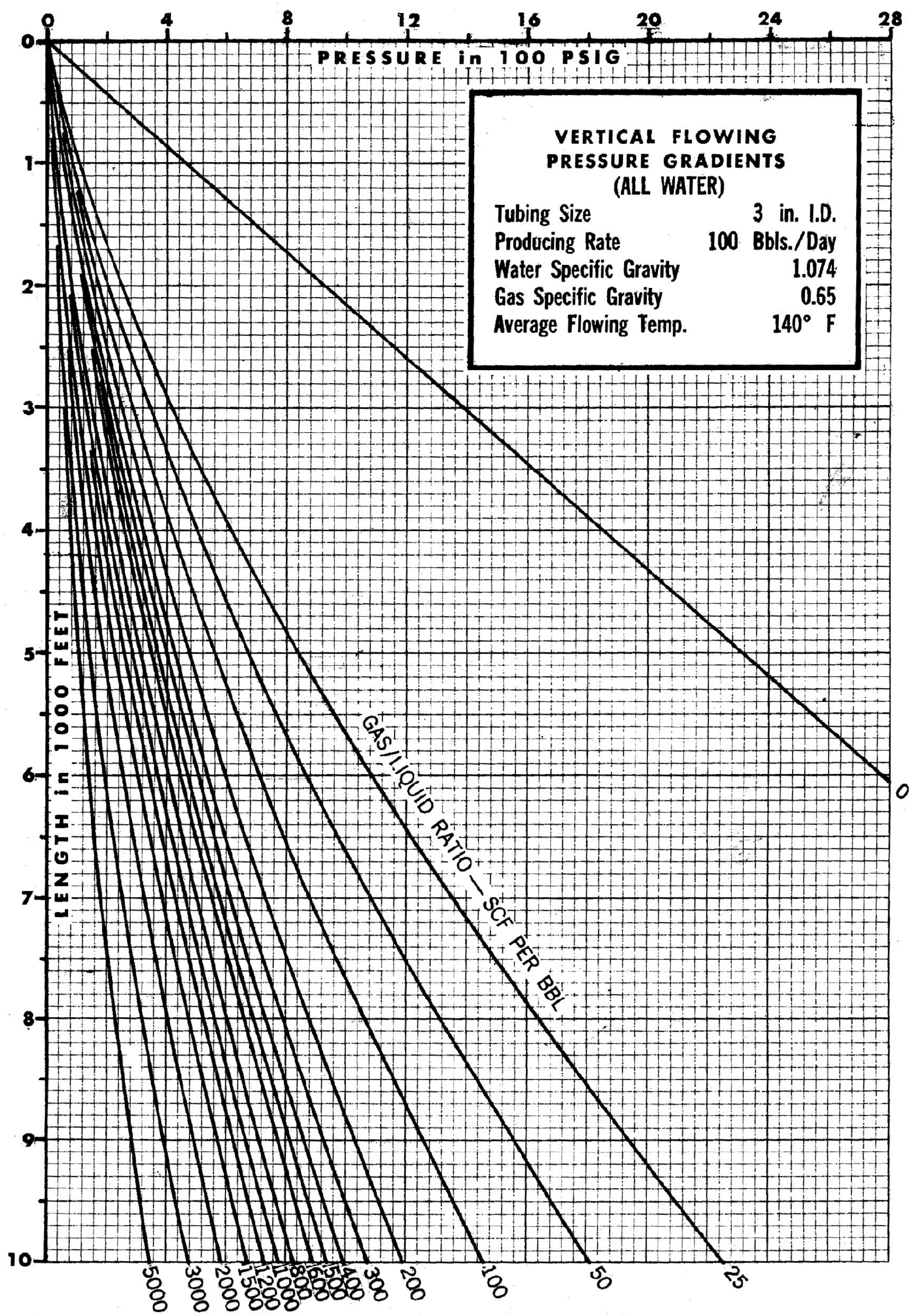


Fig 3.1.5 Pressure traverses by Hagedorn original method.

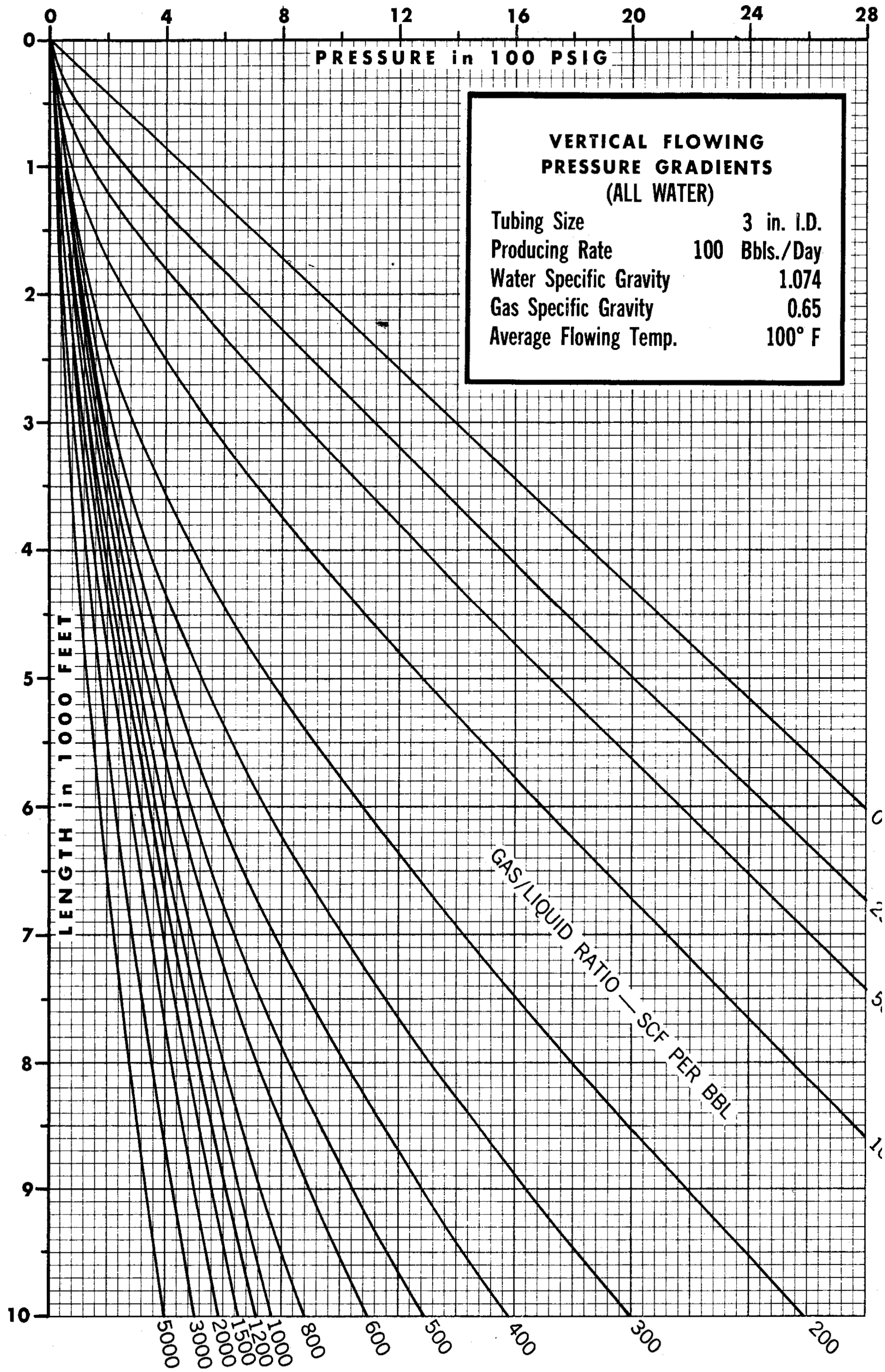


Fig 3.1.6 Pressure traverses by modified Hagedorn method.

(Fig. 3.1.5 which requires using the H_L value for no slippage shows the same pressure traverse as found in Fig. 3.1.6.)

The curves recently presented by Brownwere prepared in this manner. This discrepancy occurs only in larger tubing sizes at low flow rates and low gas-liquid ratios; therefore, except for these cases, this adjustment is not necessary.
At the suggestion of Brill and Hagedorn, the Griffith correlation was added to the Hagedorn
and Brown correlation. The Griffith correlation is used when flow is in the bubble flow regime. For a complete description of this correlation and its incorporation into the multiphase vertical flow problem, reference should be made to Orkiszewski**.**The Griffith method is explained by Orkiszewski and an example problem is worked by him. The modification incorporates the Griffith correlation after the manner suggested by Orkiszewski but only for bubble flow.
In summary, the following two adjustments are made:
The density of the mixture is calculated using the Hagedorn and Brown holdup correlation and this value is compared to the density of the mixture assuming that there is no slippage. The largest of these two values is then used.
The flow regime is determined, and if it is bubble flow, the pressure traverse is calculated by the Griffith correlation.
It is recommended that these two adjustments always be used in the generalized Hagedorn and Brown method.
It is recommended that Eq. 3.2 be solved for Ah and that a pressure traverse be calculated from the result. Rearranging Eq. 3.2, the equation becomes:

$$\Delta h = \frac{144\,\Delta P - \overline{\rho_m}\,\Delta\left(\frac{V_m^2}{2g_c}\right)}{\overline{\rho_m} + \frac{Fw^2}{2.9652*10^{11} d^5 \overline{\rho_m}}}$$

This form eliminates the necessity for a trial-and-error solution and permits solving for the distance between two pressure points. Fig.3.1.5 shows a detailed flow diagram of the Hagedorn and Brown method which shows how to calculate a pressure traverse by this method.

R. Detailed procedure for calculating a vertical pressure traverse by the method of Hagedornand Brown

Equation:

$$144\frac{\Delta p}{\Delta h} = \overline{\rho_m} + \frac{Fw^2}{2.9652*10^{11} d^5 \overline{P_m}} + \overline{\rho_m}\,\Delta\,\frac{\left(\frac{V_m{}^2}{2g_c}\right)}{\Delta h}$$

Solving for the depth increment;

$$\Delta h = \frac{144\,\Delta P - \overline{\rho_m}\,\Delta\left(\frac{V_m^2}{2g_c}\right)}{\overline{\rho_m} + \frac{Fw^2}{2.9652*10^{11} d^5 \overline{\rho_m}}}$$

Start with the known pressure p_1, assume a value for p_2 and calculate the depth increment.

1) Calculate the average pressure between the two pressure points;

$$\bar{P} = \frac{P_1 + P_2}{2} + 14.7$$

2) Calculate the specific gravity of oil;

$$\gamma 0 = \frac{141.5}{131.5 + \text{API}}$$

3) Find total mass associated with one bbl of stock tank liquid;

$$M = \gamma_{0(350)}\left(\frac{1}{1+\mathrm{W_{or}}}\right) + \gamma_{\mathrm{w}(350)}\left(\frac{\mathrm{W_{or}}}{1+\mathrm{W_{or}}}\right) + (0.0764)(\mathrm{GLR})\gamma_{\mathrm{g}}$$

4) Calculate the mass flow rate;

W = q * M

5) Obtain R_s at $\bar{P}$ and $\bar{T}$ from gthe given graph .

6) Calculate the density of liquid phase;

$$\rho_L = \left[\frac{\gamma_0(62.4) + R_s\gamma_g\left(\frac{0.0764}{5.614}\right)}{B_0}\right]\left(\frac{1}{1+\mathrm{W_{or}}}\right) + \left[\gamma_{\mathrm{w}}(62.4)\left(\frac{\mathrm{W_{or}}}{1+\mathrm{W_{or}}}\right)\right]$$

7) Assuming $\overline{T}$ = constant , find a values of $\overline{Z}$ for a constant $\overline{T}$, $\overline{P}$ and Υ_g .
8) Calculate the average density of the gas phase;

$$\overline{\rho_g} = \gamma_g (0.0764) \left(\frac{\overline{P}}{14.7}\right)\left(\frac{520}{\overline{T}}\right)\frac{1}{\overline{Z}}$$

9) Calculate the average viscosity of the oil from appropriate correlations. As noted , a knowledge of the fluid properties of the oil , $\overline{P}$ and / or $\overline{T}$ is required.
10) Determine the average water viscosity from the given graph.
11) Calculate the liquid mixture viscosity;

$$\mu_L = \mu_0 \left(\frac{1}{1 + W_{or}}\right) + \mu_w \left(\frac{W_{or}}{1 + W_{or}}\right)$$

(this can only be an approximate since the viscosity of the two immiscible liquid is quite complex)
12) Assuming constant surface tension at each pressure point , calculate the liquid mixture surface tension ;

$$\boldsymbol{\sigma}_L = \boldsymbol{\sigma}_0 \left(\frac{1}{1 + W_{or}}\right) + \boldsymbol{\sigma}_w \left(\frac{W_{or}}{1 + W_{or}}\right)$$

(again this represents only an approximation of the surface tension of the liquid phase)
13) Calculate the viscosity number ;

$$N_L = 0.15726\, \mu_L \left(\frac{1}{\rho_L \boldsymbol{\sigma}_L^3}\right)^{\frac{1}{4}}$$

14) From fig. 3.1.2 determine CN_L .
15) Calculate area of tubing ;

$$A_P = \frac{\pi d^2}{4}$$

16) From fig. 3.1.3 obtain B_o at $\overline{P}$, $\overline{T}$.
17) Assuming B_w= 1.0 , calculate the superficial liquid velocity ;

$$V_{sl} = \frac{5.61\, q_L}{86400\, A_p}\left\{B_0 \left(\frac{1}{1 + W_{or}}\right) + B_w \left(\frac{W_{or}}{1 + W_{or}}\right)\right\}$$

18) Calculate the liquid velocity number ;

$$N_{LV} = 1.938\, V_{sl} \left(\frac{\rho_L}{\boldsymbol{\sigma}_L}\right)^{\frac{1}{4}}$$

19) Calculate superficial gas velocity ;

$$V_{sg} = q_l \left\{\frac{GLR - R_s\left(\frac{1}{1+W_{or}}\right)}{86400\, A_P}\right\}\left(\frac{14.7}{\overline{P}}\right)\left(\frac{\overline{T}}{520}\right)\left(\frac{\overline{Z}}{1}\right)$$

20) Determine the gas velocity number ;

$$N_{GV} = 1.938\, V_{sg} \left(\frac{\rho_L}{\boldsymbol{\sigma}_L}\right)^{\frac{1}{4}}$$

21) Check the bubble floe regime to determine whether to continue with Hagedorn&Brown correlation (or) to proceed with the Griffith's correlation for bubble flow.

A number 'A' must be calculated by the following ;

$$A = 1.071 - \left[0.2218\, \frac{(V_{sl} + V_{sg})^2}{d}\right]$$

If $A \geq 0.13$, then use the calculated value and if $A \leq 0.13$, then use $A = 0.13$
Another number 'B' is calculated by the following ;

$$B = \frac{V_{sg}}{V_{sl} + V_{sg}}$$

(if (B-A) is positive proceed with Hagedorn& Brown correlation , if (B-A) is negative proceed with Griffith's correlation)

22) Find the pipe diameter number ;

$$N_d = 120.872d\sqrt{\frac{\rho_L}{\sigma_L}}$$

23) Calculate the holdup correlation function ;

$$\phi = \left(\frac{N_{lv}}{N_{gv}^{0.575}}\right)\left(\frac{\bar{P}}{14.7}\right)^{0.10}\left(\frac{C_{NL}}{N_d}\right)$$

24) Obtain $\left(\frac{H_L}{\psi}\right)$from the given graph.

25) Determine the secondary correction factor correlating parameter ;

$$\phi = \left(\frac{N_{gv}N_L^{0.380}}{N_d^{2.4}}\right)$$

26) Obtain φ from fig 3.1.3

27) Calculate the value for H_L ;

$$H_L = \left(\frac{H_L}{\psi}\right)\boldsymbol{\psi}\ \text{(for low viscosity there willbe no correlation ,}\varphi\text{=1)}$$

28) In order to obtain friction factor , determine a value for the two phase Reynold's number

$$(N_{Re})_{TP} = \frac{2.2 * 10^{-2} w}{(d)(\mu_L^{H_L})\left(\mu_g^{(1-H_L)}\right)}$$

29) Determine the value for ϵ/d . If the value for ϵ is not known , a good value to use is 0.00015 ft which is an average value given for commercial steel.

30) Obtain the friction factor from the given input.

31) Calculate the average two-phase density of the mixture by two methods;

a) Using the value of H_L from step(27) , calculate the $\bar{\rho}_m$ as follows;

$$\bar{\rho}_m = \bar{\rho}_L H_L + \bar{\rho}_g(1 - H_L)$$

b) Calculate the value of $\bar{\rho}_m$ assuming no slippage .

ompare the two values of $\bar{\rho}_m$ from a) & b) & use the greater value.

32) Repeat steps (5),(7),(16),(17) and (19) for both P_1& P_2 .

33) Calculate the two-phase mixture velocity at both P_1& P_2 ;

$$V_{m1} = V_{sl1} + V_{sg1}$$
$$V_{m2} = V_{sl2} + V_{sg2}$$

34) Determine a value for $\Delta(V_m^2)$;

$$\Delta(V_m^2) = [V_{m1}^2 - V_{m2}^2]$$

35) CalculateΔh corresponding to ΔP ;

$$\Delta h = \frac{144\,\Delta P - \overline{\Delta_m}\,\Delta\left(\frac{V_m^2}{2g_c}\right)}{\overline{\rho_m} + \frac{Fw^2}{2.9652*10^{11}d^5\overline{\rho_m}}}$$

36) Starting with P_2 and the known depth at P_2 , assume another pressure point and repeat the procedure until reaching total depth (or) until reaching the surface depending upon whether you are starting from the bottom (or) top of the tube.

II. RESULTS &OBSERVATION

By using the above equations following the procedure with the help of given data and assumed data pressure drop for various tubing sizes (ranging from 0.13' in to 3.5' in) were calculated and respective graphs were obtained, which gives us the clear view of pressure vsflowrate curve (TPR curve).Sensitivities were made for varying tubing sizes and flowrate for which respective pressures were obtained at constant depth (5000 ft.).

The following flowratevs pressure graph were obtained from the calculated sensitivities;

The assumed data to develop this correlation was acquired under the following conditions:

A. 5000-ft experimental vertical well;

B. 1.9, 2.28, 2.8,3, 4.2 in nominal diameter tubes;

C. Liquids with different viscosities: 0.63 cp water, 17 cp oil;

D. Liquid flow rates: 100-5000bbl/day; and,

E. Gas-liquid ratios: 0-1000scf/bbl.

Fig.1. was obtained by using Hagedorn& Brown correlation calculation method for predicting pressure drop with the help of given and assumed data. This graph is constructed with 'flowrate in the X-Axis' and 'pressure in the Y-Axis'. It was calculated with the tubing size of 2 3/8' inches diameter and respective pressure were obtained for the flowrate ranging from 100 to 5000 bpd at the depth of 5000ft.Fig.2was obtained by using Hagedorn& Brown correlation calculation method for predicting pressure drop with the help of given and assumed data. This graph is constructed with 'flowrate in the X-Axis' and 'pressure in the Y-Axis'. It was calculated with the tubing size of 2 5/8' inches diameter and respective pressure were obtained for the flowrate ranging from 100 to 5000 bpd at the depth of 5000ft.

Fig.3. was obtained by using Hagedorn& Brown correlation calculation method for predicting pressure drop with the help of given and assumed data. This graph is constructed with 'flowrate in the X-Axis' and 'pressure in the Y-Axis'. It was calculated with the tubing size of 2 7/8' inches diameter and respective pressure were obtained for the flowrate ranging from 100 to 5000 bpd at the depth of 5000ft.

Fig.4 was obtained by using Hagedorn& Brown correlation calculation method for predicting pressure drop with the help of given and assumed data. This graph is constructed with 'flowrate in the X-Axis' and 'pressure in the Y-Axis'. It was calculated with the tubing size of 3 1/2' inches diameter and respective pressure were obtained for the flowrate ranging from 100 to 5000 bpd at the depth of 5000ft.Fig.5. was obtained by using Hagedorn& Brown correlation calculation method for predicting pressure drop with the help of given and assumed data. This graph is constructed with 'flowrate in the X-Axis' and 'pressure in the Y-Axis'. It was calculated with the tubing size of 4 1/2' inches diameter and respective pressure were obtained for the flowrate ranging from 100 to 5000 bpd at the depth of 5000ft.

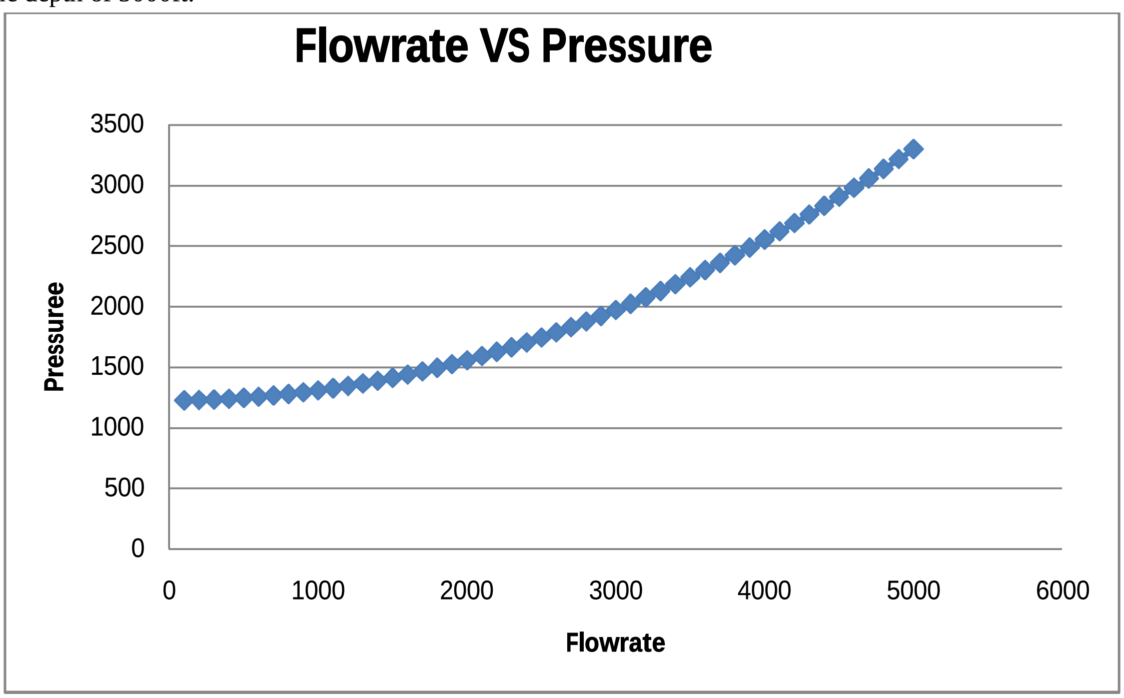


Fig. 1.Flowratevs pressure graph.

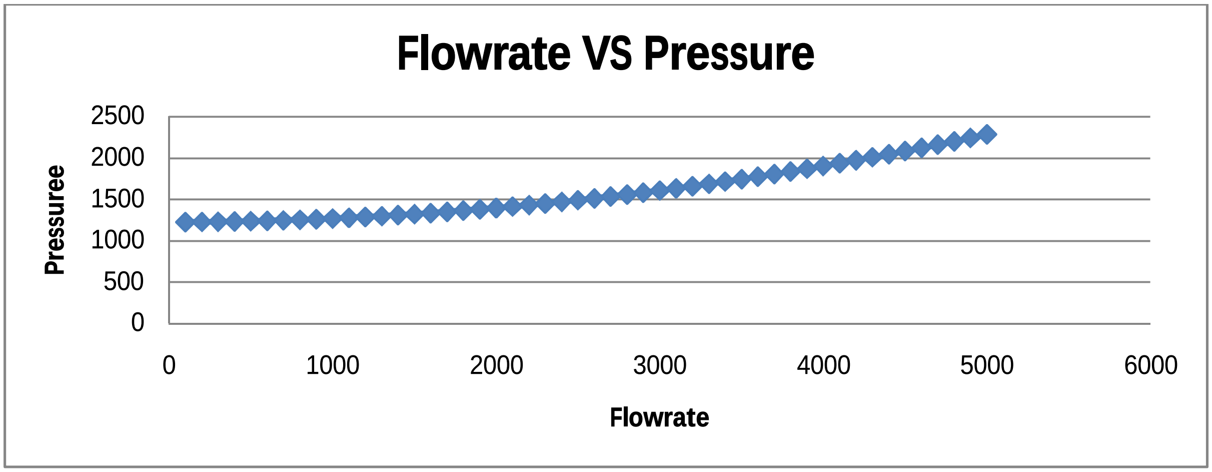


Fig. 2.Flowrtevs pressure graph.

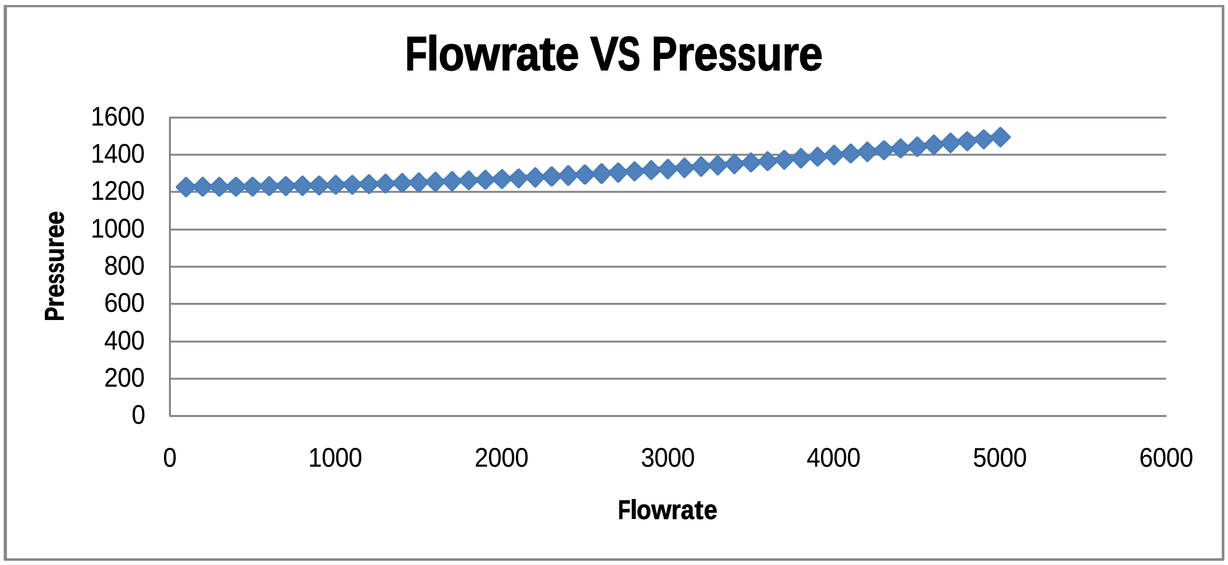


Fig. 3.Flowratevs pressure graph.

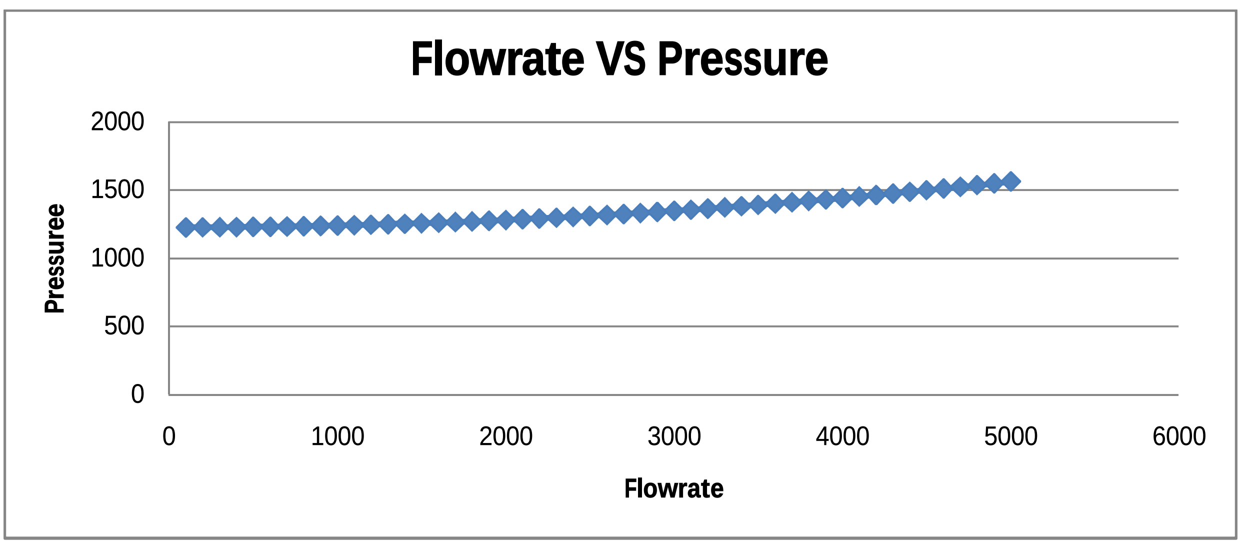


Fig. 4.Flowratevs pressure graph.

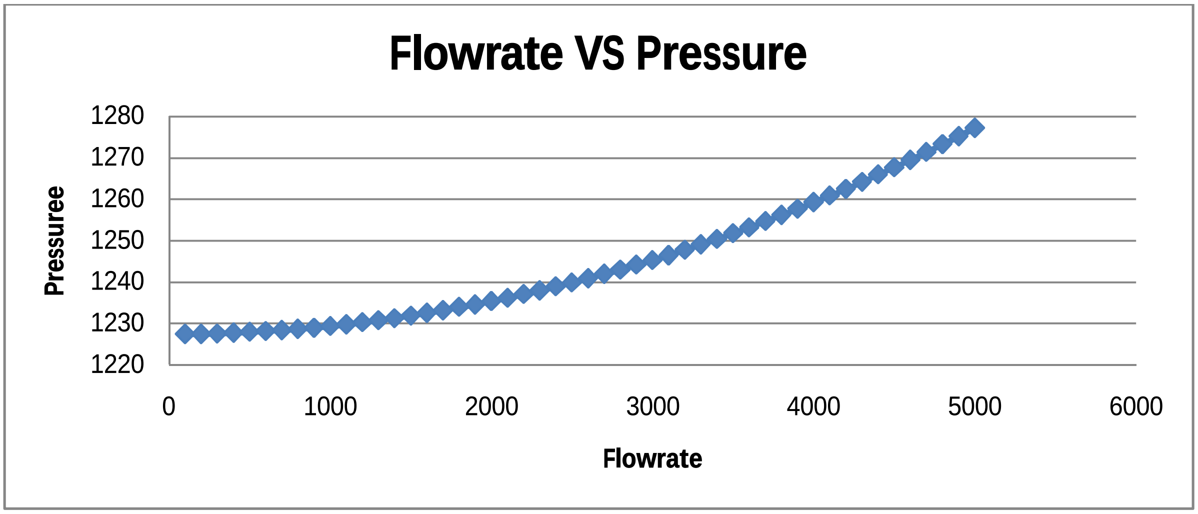
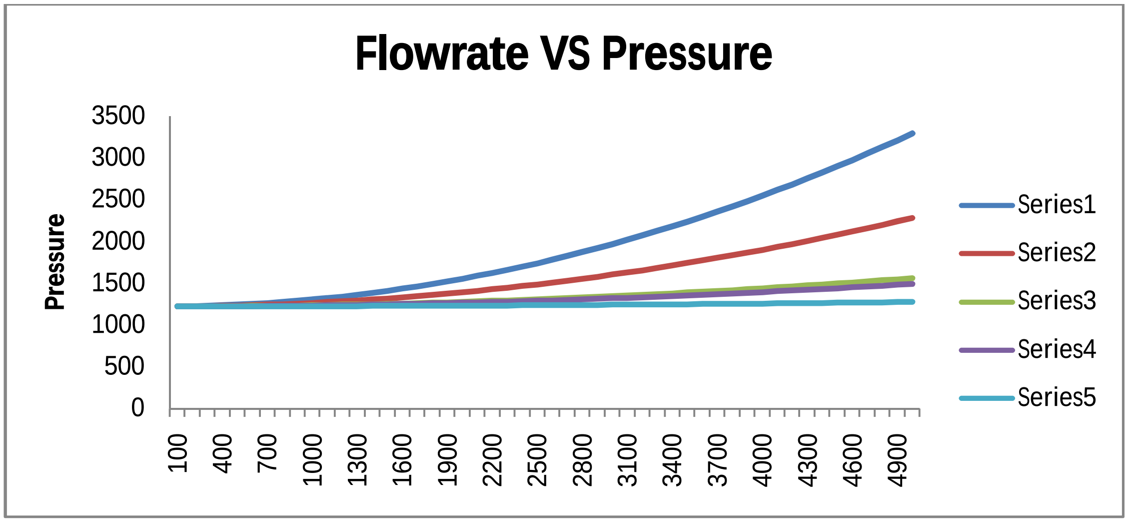


Fig. 5.Flowratevs pressure graph.



Where;

Series1 – from fig.1
Series2 – from fig.2
Series3 - from fig.3
Series4 – from fig.4
Series5 – from fig.5

By combining these graphs together (fig.1 to fig.5), taking flowrate in X-Axis and pressure in Y-Axis, creating flowratevs pressure graph gives us a clear result of tubing performance relationship curves for respective tubing sizes from the given and assumed data. Observation tells us that series2 – from fig.2 is the best tubing size for the given data.

III. CONCLUSION

The observed Haagedorn& Brown correlation by running sensitivities for various inputs gave us the appropriate tubing size to be used for the given data.Also other parameters like flow rate, pressure drop, liquid holdup can be obtained through this correlation method.Similarly, we can obtain different graphs for different inputs for various vertical multi-phase flow, which helps us to predict the pressure drop and to choose the reliable tubing sizes from the particular data.It is necessary to be able to predict a vertical multiphase flow pressure traverse in order to correctly select completion strings, predict flow rates, and design artificial lift installations. Most of the progress towards a solution of the problem has been made since the publication of Poettmann and Carpenter's paper in 1952. Most of the approaches use some form of the general energy equation. This work can be extended in future by running sensitivities on changing temperatures for various types of flow pattern.

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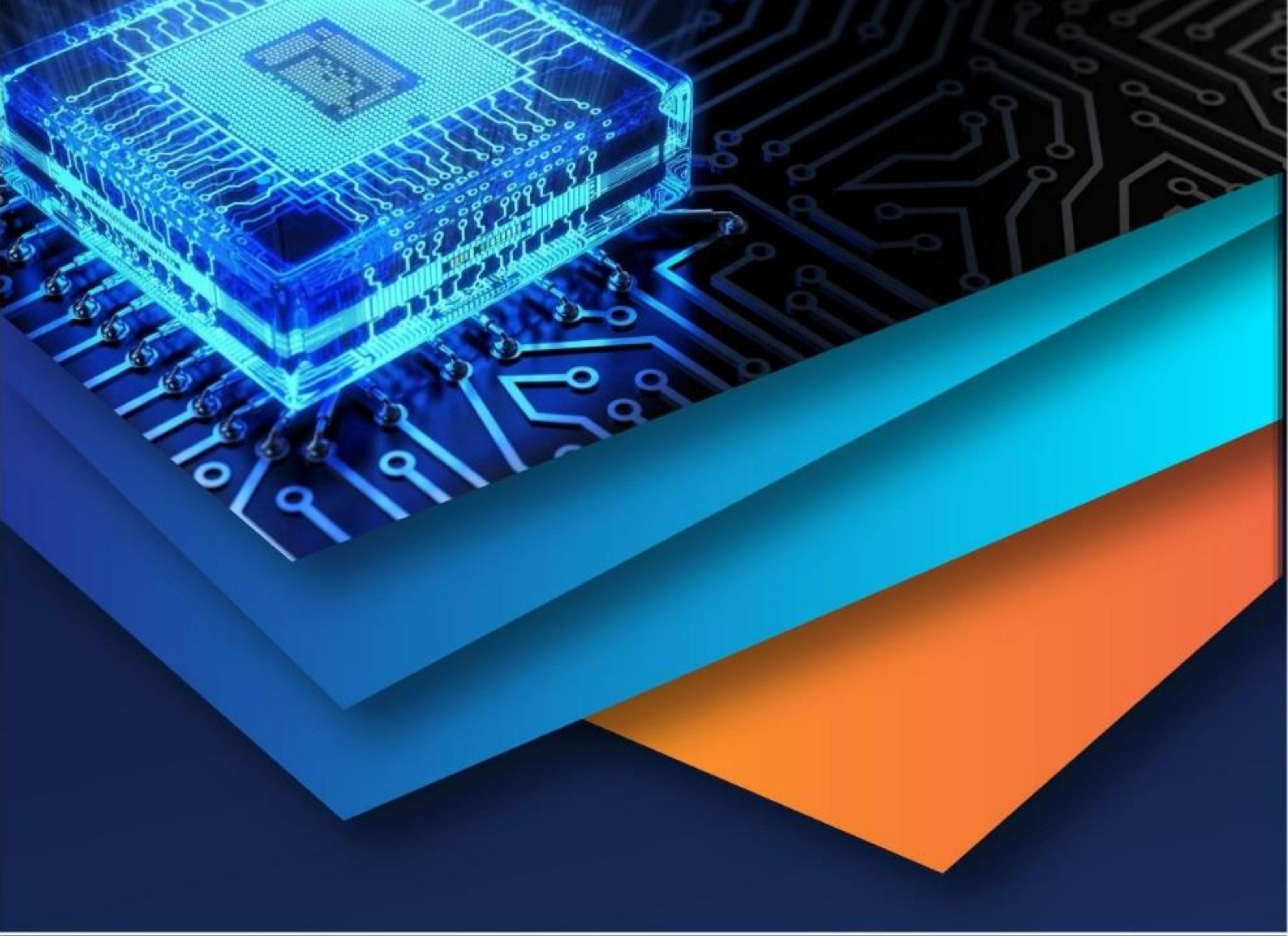

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