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Rogaland Distriktshøgskole

MEASUREMENTS OF GAS AND LIQUIDS

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PHYSICAL PROPERTIES OF GASES AND LIQUIDS

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The importance of accurate metering in connection with oil and gas production is obvious. An error in the determination of the massflow means millions of \$ during a year production. Metering is important in connection with custody transfer, and likewise during well-testing and production monitoring. The following will be a discussion of some physical properties of oil and gas with importance for the accuracy of the measurements.

Although we speak of the production of oil and gas, there is in fact a spectrum of petroleum fluids produced. In the book "The Properties of Petroleum Fluids",¹ McCain distinguishes among the fluids as shown in Table 1.

<u>Type</u>	<u>Typical Yield</u>		<u>Oil Properties</u>	
	<u>Bsto/Brf</u>	<u>GOR</u>	<u>°API</u>	<u>Color</u>
Low-Shrinkage Crude Oil	0.85	0/500	≤ 30	Dark
High-Shrinkage Crude Oil	0.65	500/8000	30-50	Dark
Retrograde Condensate Gas	0.25	8000/70000	50-60	Slight
Wet Gas	Condensate	70000/100000	> 50	White
Dry Gas	No Cond.	100000+	-	-

Table 1. Petroleum Fluid Spectrum.

Shrinkage refers to the fact that gas evolves from the crude oil as its pressure falls below that in the reservoir. As the gas evolves, the remaining liquid volume decreases. Almost all crude oil evolves some gas during processing. This gas is referred to as associated or separator gas. Wet or dry gas is accompanied by little or no liquid production as such and will be referred to as natural gas.

Petroleum Fluid Composition.

The type of chemical compounds present in crude oil and natural gas are shown in Tables 2 and 3.

Hydrocarbons
 Gases - N₂, CO₂
 Sulfur Compounds (H₂S, Mercaptans,
 Alkyl Sulfides)
 Organic Compounds Containing N₂,
 O₂, and Metals
 Water

Table 2. Crude oil components.

Hydrocarbons
 Inert Gases (N₂, He)
 Acid Gases (H₂S, CO₂)
 Sulfur Compounds (Mercaptans, Alkyl
 Sulfides)
 Water

Table 3. Natural Gas Components.

The differences lies in the boiling range and the relative amounts of the components. In petroleum or gas production the hydrocarbons are the desired material, and the other compounds are contaminants or impurities.

In Table 4 the hydrocarbon composition of two typical crude oils are shown. No breakdown of the C₇₊ -fraction is shown, these fractions are complex mixtures of paraffinic, naphthenic and aromatic hydrocarbons.

	<u>Oil A</u>	<u>Oil B</u>
N ₂	0.1	7.1
CO ₂	0.4	0.4
H ₂ S	0.0	0.0
C ₁	17.0	40.5
C ₂	8.7	12.8
C ₃	6.5	9.9
iC ₄	1.4	0.6
nC ₄	3.7	1.6
iC ₅	1.7	0.9
nC ₅	2.0	1.1
C ₆	5.4	1.2
C ₇₊	<u>53.1</u>	<u>23.9</u>
Total	100.0	100.0
GOR (scf/bbl)	290	1060

Table 4. Typical Crude Oil Wellstream Analysis.

The compositions in Table 4 are full wellstream or reservoir fluid analysis. As noted previously, these crude oils will evolve gas (primarily methane and ethane) as the pressure is lowered below the formation pressure. After a series of flash separations to atmospheric pressure, the remaining oil is

referred to as stock-tank oil. The combined gas evolved (expressed in standard cubic feet) divided by the stock-tank oil rate (expressed as 60 F barrels) is termed the gas-oil ratio or GOR. The oils in Table 4 have GORs of 290 (low-shrinkage crude) and 1060 (high-shrinkage crude).

Table 5 shows the composition of some natural and separator gases. The heavier hydrocarbons in these gases are regarded as recoverable liquids. The relative proportion of recoverable liquids is expressed as gallons liquid at 60 F (if totally condensed) per 1000 scf. of the gas. (GPM: gallons condensible hydrocarbon per Mscf. of associated or natural gas.)

	Natural Gas				Separator Gas			
	Full Wellstream		Separated at 950 Psig and 80°F		Separated at 0 Psig and 75°F		Separated at 15 Psig and 75°F	
	Mol %	GPM	Mol %	GPM	Mol%	GPM	Mol %	GPM
C1	89.15		91.91		59.04		63.49	
C2	4.84	1.29	4.81	1.28	10.42	2.78	10.87	2.90
C3	1.80	0.49	1.63	0.45	15.12	4.15	14.48	3.96
iC4	0.55	0.18	0.44	0.14	2.39	0.78	2.00	0.65
nC4	0.84	0.26	0.62	0.19	7.33	2.30	5.78	1.82
iC5	0.27	0.10	0.15	0.05	2.00	0.73	1.32	0.48
nC5	0.38	0.14	0.19	0.07	1.72	0.62	1.02	0.37
C6	0.51	0.21	0.15	0.06	1.18	0.48	0.62	0.25
C7+	1.66	0.89	0.10	0.05	0.80	0.40	0.42	0.21
	100.00		100.00		100.00		100.00	
GPM C2+		3.56		2.29		12.24		10.64
GPM C3+		2.27		1.01		9.46		7.74

Table 5. Typical Gas Analysis.

A gas is termed lean or rich as follows:

<u>Classification</u>	<u>GPM C2+</u>
Lean	< 2.5
Moderately-Rich	2.5 - 5
Very Rich	> 5

$$GPM = \left(\frac{\text{scf}}{\text{Mcf}} \right) / \left(\frac{\text{gal}}{\text{scf}} \right) = \left(\text{Mol \%} \times 10 \right) / \left(\frac{379.49}{\text{gal/lb.mol}} \right)$$

The above classification is based on C2+ because ethane is regarded as a desirable feed for petrochemical processes and is recoverable as a liquid in expander plants.

Natural gas liquids (NGL) include the following products:

<u>Product</u>	<u>Primary Use</u>
Ethane	Petrochemical feedstock
Propane	Fuel
i-Butane	Refinery alkylation plant feedstock
n-Butane	Refinery gasoline blending component
Pentanes and Heavier	Refinery gasoline blending component

Petroleum Fluid Phase Behavior.

The pressure-temperature phase relation of a petroleum fluid is important because it will give the condition of the fluid at various conditions of processing, downhole or at the surface and in transport lines.

In Figure 1 a typical pressure-temperature phase diagram of a petroleum fluid is shown.²

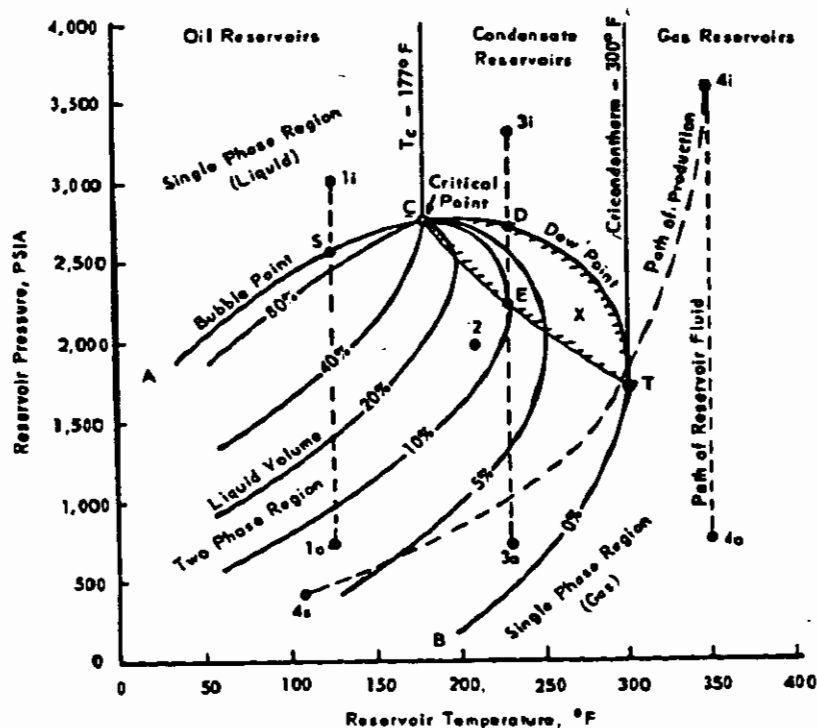


Figure 1. Pressure-temperature phase diagram of a petroleum fluid

P-T diagrams show the effects of pressure and temperature on the physical state of a hydrocarbon system. However, the phase diagram in Figure 1 is for a specific composition. Although a different fluid would have a different P-T diagram, the general configuration is similar. In Figure 1, the area enclosed by the bubblepoint curve, A-S-C, and the dewpoint curve, C-D-T-B, to the lower left, is the region of pressure-temperature combinations in which both gas and liquid phases will exist. The curves within the two-phase region show the gas-liquid percentages for any pressure and temperature. The curve A-S-C-T-B separates the two-phase region from the single-phase regions where all the fluid exist in a single phase. The bubblepoint curve separates the two-phase region from the single-phase liquid region, while the dewpoint curve separates it from the single-phase vapor region. Point C, where the bubblepoint curve and the dewpoint curve meet, is called the critical point, and the corresponding temperature is called the critical temperature.

The critical temperature, T_c , is defined as: the temperature above which a gas can not be liquefied by application of pressure alone.

The critical pressure, P_c , is the pressure which a gas exerts when in equilibrium with the liquid phase and at the critical temperature.

The critical volume, v_c , is the specific volume of the fluid at the critical pressure and critical temperature.

If Figure 1 represents a reservoir fluid, the point 1_i , indicates an oil reservoir, called a bubblepoint reservoir. The point 2 indicates an oil reservoir with an initial gas cap. The point 3_i indicates initially a gas reservoir. During production it will show retrograde condensation, i.e. as the pressure decreases liquid will condense. As retrograde condensation occurs, the reservoir fluid composition changes and the P-T envelope shifts, increasing retrograde liquid condensation. Point 4_i indicates a pure gas reservoir. However, after the fluid leaves the reservoir and enters the wellbore, the temperature, as well as the pressure, will decline until surface temperature and pressure conditions are reached. The fluid produced through the wellbore and into

surface separators at point 4_s , though of the same composition as the reservoir fluid, has entered the two-phase region due to the pressure and temperature decline along 4_i-4_s . This accounts for the production of a considerable amount of liquid at the surface, from a gas in the reservoir. If point 4_s lies outside the two-phase envelope, in the single-phase (vapor) region, then only gas will exist at the surface. No liquid will be formed in the reservoir or at the surface, and the gas is called a dry natural gas. The word dry indicates that the fluid does not contain enough of the heavier hydrocarbons to form a liquid at surface conditions. Nevertheless, it may contain liquid fractions which can be removed by lowtemperature separation or by natural gasoline plants.

Equations of State.

Any equation correlating P, V , and T for a fluid is called an equation of state. The most wellknown and simple equation of state is the ideal gas law:

$$P V = n R T \quad (1)$$

This equation describes the behavior of most hydrocarbon gases at pressure and temperature conditions close to atmospheric. However, hydrocarbon gases are real gases, and at moderate pressures the gas tend to compress more than the ideal gas law predicts, particularly at temperatures close to the critical temperature. At very high pressure the gas tend to compress less than the ideal gas law predicts.

To correct for the deviation between the measured volume of an gas and that calculated using the ideal gas law, an empirical factor, Z , called the gas deviation factor, is used. For real gases we can then write the equation:

$$P V = Z n R T \quad (2)$$

or

$$\left(\frac{P}{Z}\right) V = n R T$$

The Z -factor can be interpreted as a term by which the pressure must be corrected to account for the departure from the ideal gas law. The Z -factor is a function of both absolute pressure and temperature.³

Beginning with his thesis in 1873,⁴ J. D. van der Waals proposed his theorem of corresponding states. Before stating the theorem the following terms will be defined:

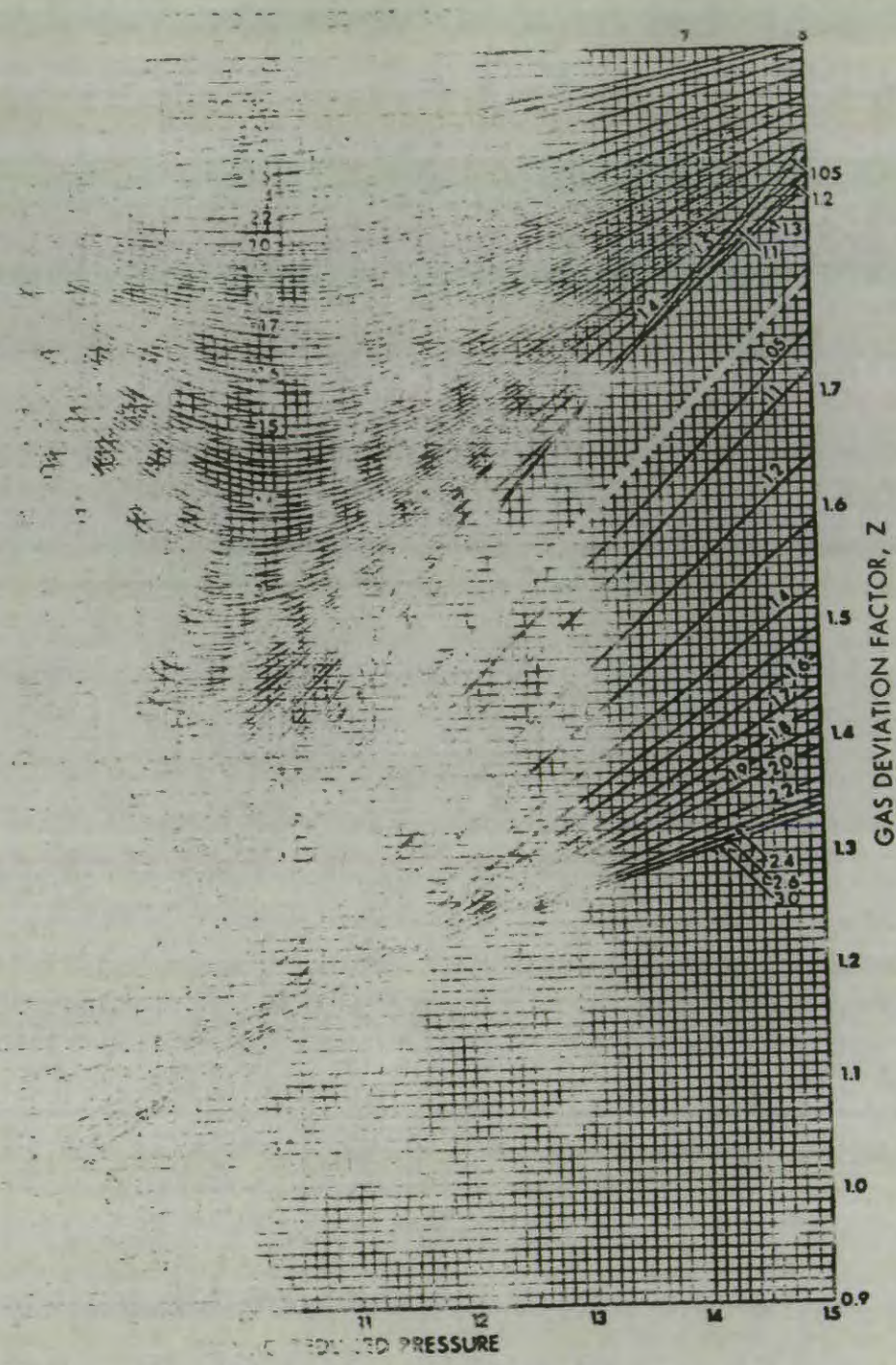
Reduced temperature, reduced pressure, and reduced volume are the ratios of the actual temperature, pressure, and specific volume to the critical temperature, critical pressure, and critical volume, respectively.

$$T_r = \frac{T}{T_c}, \quad p_r = \frac{p}{p_c}, \quad v_r = \frac{v}{v_c} \quad (3)$$

The physical characteristics of a fluid are controlled by the relative nearness of any state point to the critical point. If two different fluids have the same reduced pressure and temperature, then the two fluids are in corresponding states and any other property, like the reduced density will be the same for both fluids. This is the principle of corresponding states. Stated in other words: the deviation of a real gas from the ideal gas law is the same for different gases at the same corresponding conditions of reduced temperature and pressure. Reduced vapor pressure, reduced density, reduced enthalpy, reduced entropy etc. are examples of this theorem's uses for the purposes of generalizing results.

The Z-factor can either be determined experimentally for the actual gas mixture at specified pressure and temperature, or it can be determined from well established correlations. The Z-factor can now easily be calculated using an equation of state.

In 1941, Standing and Katz⁵ presented a Z-factor chart, based on binary mixtures and saturated hydrocarbon vapor data. Figure 2 shows this Z-factor chart. In this chart the Z-factor is given as a function of reduced pressure and reduced temperature. This chart is generally reliable for sweet natural gases and can be corrected for those containing hydrogen sulfide and carbon dioxide. It has become one of the most widely accepted correlations in the petroleum industry. This correlation requires the knowledge of the gas composition, to determine the pseudo critical pressure and temperature, or the apparent molecular weight of the mixture. By use of Kay's



Graph showing the Gas Deviation Factor for natural gases (after ...)

... properties are⁶:

$$\text{Pressure } P_{pc} = \sum_i y_i P_{ci} \quad (4)$$

$$\text{Temperature } T_{pc} = \sum_i y_i T_{ci} \quad (5)$$

$$\text{... } M_g = \sum_i y_i M_i \quad (6)$$

y_i = mole fraction of component i in the vapor
 P_{ci} = critical pressure for component i
 T_{ci} = critical temperature for component i
 M_i = molecular weight for component i

In cases where the gas composition is not available, but the gas gravity is known, the pseudo-critical pressure and temperature can be estimated from Figure 3.

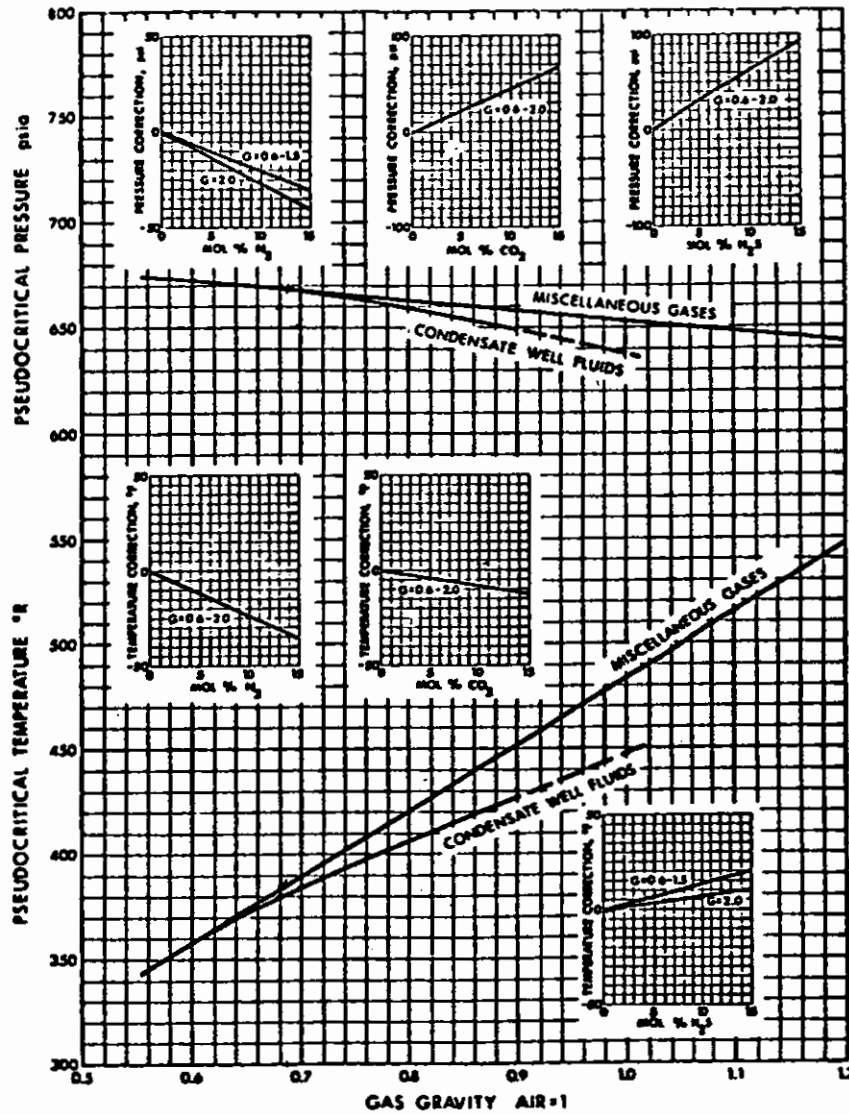


Figure 3. Pseudo critical properties of miscellaneous gases.

Useful correlations derived from Figure 3 are:

$$P_{pc} = 709.604 - 58.718 G \quad (7)$$

$$T_{pc} = 170.491 + 307.344 G \quad (8)$$

In order to try to describe non-ideal, real gas behavior, a large number of equations of state have been developed. Each is empirical in that it correlates a specific set of data using one or more empirical constants. The oldest and most well known two constant equation of state is the one developed by van der Waals:

$$P = \frac{RT}{V-b} - \frac{a}{V^2} \quad (9)$$

$$a = \frac{27 R^2 T_c^2}{64 P_c} \quad , \quad b = \frac{RT_c}{8 P_c}$$

Of other well known equations related to the van der Waals equation can be mentioned:

The Redlich-Kwong equation⁷: The original Redlich-Kwong equation is given in the form:

$$P = \frac{RT}{V-b} - \frac{a}{T^{0.5} V(V+b)} \quad (10)$$

The constants a and b are functions of critical pressure and temperature. Numerous modifications of the Redlich-Kwong equation have been proposed. One of the more recent modifications of R-K is that proposed by Soave in 1972⁸. The S-R-K equation has rapidly gained acceptance by the hydrocarbon processing industry because of its capability for generating reasonably accurate equilibrium ratios in vapor-liquid equilibrium calculations. The S-R-K equation is given in the form:

$$P = \frac{RT}{V-b} - \frac{a(T)}{V(V+b)} \quad (11)$$

The most important change from eqn. 10 to eqn. 11 is that the term $a/T^{0.5}$ is changed to a function $a(T)$.

One of the newest developed equations is the Peng-Robinson equation of state⁹. The equation is given in the form:

$$P = \frac{RT}{V-b} - \frac{a(T)}{V(V+b)+b(V-b)} \quad (12)$$

In the P-R equation the second term has changed from the S-R-K equation.

Any equation of state can be used to generate an expression for the Z-factor. The Peng-Robinson equation of state gives:

$$Z^3 - (1-B)Z^2 + (A - 3B^2 - 2B)Z - (AB - B^2 - B^3) = 0 \quad (13)$$

$$A = \frac{aP}{R^2T^2}, \quad B = \frac{bP}{RT}, \quad Z = \frac{PV}{RT}$$

$$a(T) = a(T_c) \cdot \alpha(T_r, \omega)$$

$$b(T) = b(T_c)$$

Due to modern computers, equations of this kind can easily be used in Z-factor calculations, vapor-liquid equilibrium calculations and to calculate other physical and thermodynamic properties. This kind of equations usually gives good results, but there is a tendency for larger deviations at high pressure and temperatures¹⁰. The Peng-Robinson equation seems to give better predictions for the liquid state than the S-R-K equation.

In vapor-liquid equilibria the equilibrium condition is defined as the situation where the vapor fugacity is equal to the liquid fugacity. The fugacity coefficient can be calculated from the fundamental thermodynamic equation¹¹:

$$\ln \phi_i = \frac{1}{RT} \int_0^P \left(V_i - \frac{RT}{P} \right) dp \quad (14)$$

The fugacity of the component i in the vapor phase is f_i^V , and in the liquid phase f_i^L .

$$f_i^V = \phi_i^V y_i P ; f_i^L = \phi_i^L x_i P$$

At equilibrium between the vapor and liquid phase $f_i^V = f_i^L$. The equilibrium constant $K_i = \frac{y_i}{x_i}$, and thus,

$$K_i = \frac{\phi_i^L}{\phi_i^V}$$

In the form the equations are presented above, they are in principle only valuable for single components. However, real vapor-liquid equilibrium stages deal with multicomponent systems, and the equations have to be adopted to this. The constants in the equations have to be calculated from the single component constants, and the success of the calculation strongly depends on the mixing rules used, and also on reliable PVT data and chemical analysis of the composition of the fluids. Figure 4 shows the results from equilibrium calculations using Peng-Robinson equation, compared with actual PVT laboratory measurements.

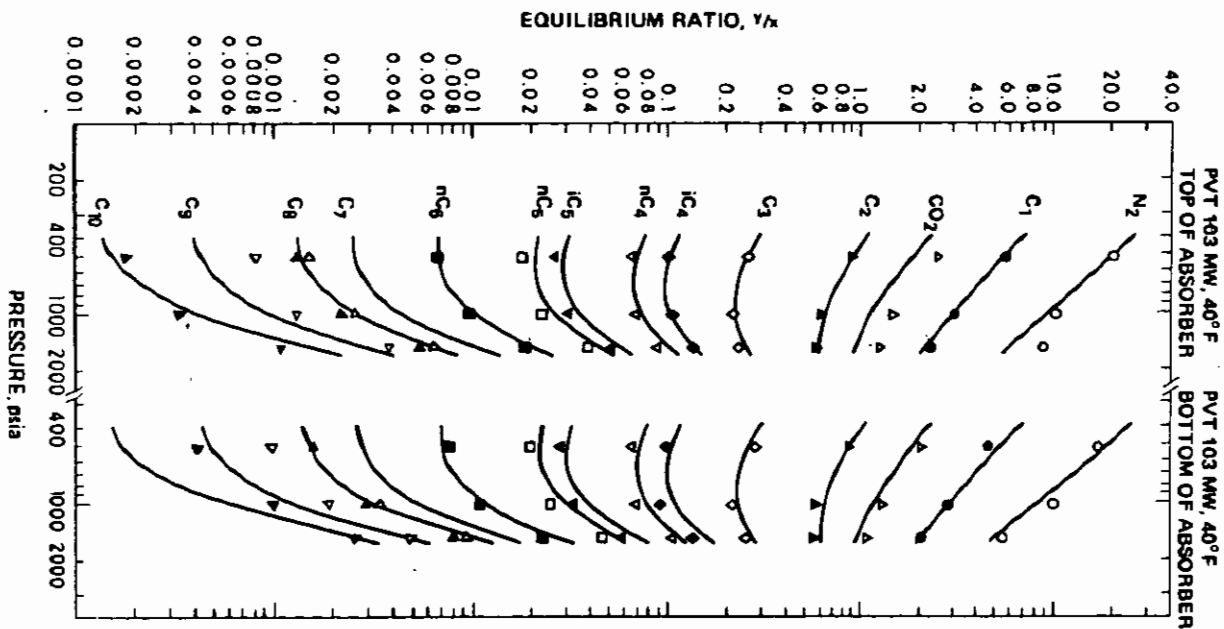
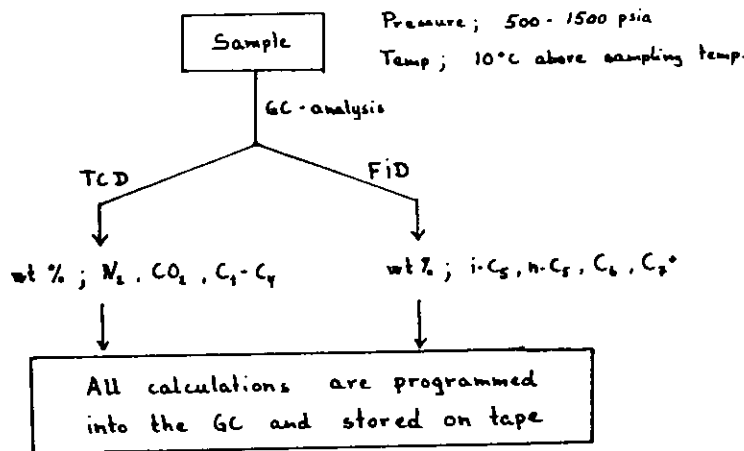


Figure 4. Experimental and calculated equilibrium ratios for 103 Mw absorber oil at 40 °F.

Composition analysis.

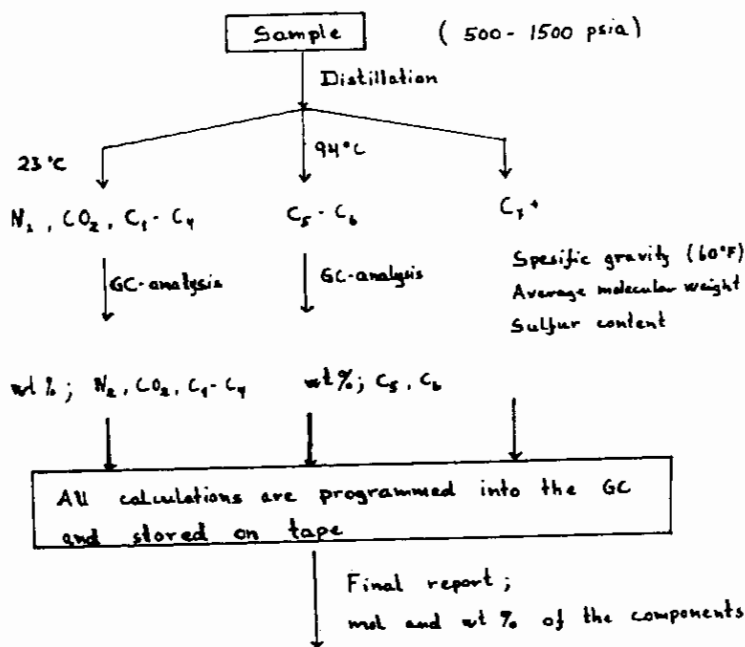
Property predictions and calculation of physical parameters for petroleum fluids, strongly depends on reliable sampling procedures for the fluids and reliable analysis of the samples. At RRI we are presently doing compositional analysis of pressured oil and gas from the North Sea. The oil and gas to be analyzed are sampled on high-pressure sample cylinders. The samples are taken at the test separator (1. stage separator). The gas samples are analyzed directly on a GC-instrument, and the component composition up to heptanes pluss (C7+) are reported, together with the calculated critical temperature, critical pressure, real gas specific gravity, average (apparent) molecular weight for the gas, ideal gas heating value, and the Z-factor. Figure 5 gives a simplified picture of the gas analysis.



	NAMES	WEIGHTS	MOLS
1	NITROGEN	0.55	0.44
2	CARBON DIOXIDE	0.73	3.19
3	METHANE	61.65	90.22
4	ETHANE	10.86	30.93
5	PROPANE	9.17	4.32
6	I-BUTANE	1.35	0.49
7	N-BUTANE	3.31	1.50
8	I-PENTANE	1.00	0.28
9	N-PENTANE	1.74	0.37
10	HEPTANES	1.17	0.28
11	HEPTANES PLUS	1.36	0.26
	CRITICAL TEMPERATURE-DEG.C		-53.9
	CRITICAL PRESSURE-PSIA		676.9
	REAL GAS SPECIFIC GRAVITY		1.7259
	AVERAGE MOLECULAR WEIGHT		20.28
	IDEAL GAS HEATING VALUE-BTU/SCF		19509
	COMPRESSIBILITY FACTOR		0.9907

Figure 5. Analysis diagram for natural gas.

The samples of pressured oil are distilled, giving a gas fraction (boiling range to 23 °C), a C₅-C₆ fraction (boiling range from 23-93 °C), and a distillation bottom, referred to as the C7+ fraction or the heptanes plus fraction. The component composition of the gas fraction and the C₅-C₆ fraction are determined by GC analysis. The C7+ fraction is analyzed for specific gravity, average molecular weight, and sulfur content. By means of a computer programme the composition of the gas fraction, the C₅-C₆ fraction, and the C7+ fraction are recombined, giving the component composition for the pressured oil up to heptanes plus. Figure 6 gives a simplified picture of the pressured oil analysis.



ANALYSE OF PRESSURED OIL
PR: 10457 MAY 26, 1961

	NAME	WEIGHT%	MOLE%
1	NITROGEN	0.01	0.04
2	CARBON DIOXIDE	0.00	0.00
3	ETHANE	0.00	00.04
4	ETHANE	1.75	5.97
5	PROPANE	2.10	5.82
6	ISOBUTANE	0.79	1.49
7	NEOPENTANE	0.11	0.77
8	ISOPENTANE	1.90	2.15
9	N-PENTANE	1.91	2.74
10	HEPTANES	61.16	71.41
11	HEPTANES PLUS	33.00	43.70

Figure 6. Analysis diagram for pressured oil.

Routinely we analyze up to C7+, but on special cases the analysis can be extended to C11+, in that the components in the C₆-C₁₀ range are reported in terms of single carbon numbers. Routinely we also run simulated distillation of crude oil up to C₂₀ reporting the results in terms of single carbon number. Further developments to increase the range of the component analysis is the subject for a research project supported by the NTNf.

For a component analysis of oil or gas to be reliable, both the sampling procedures and the analytical work in the laboratory are critical. To facilitate the evaluation of the results we have adopted two different methods. Firstly we use a calculation based on a modified version of the Watson characterization factor¹². This factor gives a measure of the relative parafinicity of the oil, and the characterization factor K_{n+} is given by the equation:

$$K_{n+} = 4.5579(M_{n+})^{0.15178} \cdot (G_{n+})^{-0.84573} \quad (15)$$

For the Ekofisk oil the K₇₊ value varies in the range 11.82 - 12.06, with usually very small variations for oil samples from the same well. This gives us a sensitive test for the parameters of the C7+ fraction.

The second method to evaluate the results of the analysis, and the quality of the sampling, has a sound basis in thermodynamic principles. For a representative sample to be taken from the test separator, there has to be equilibrium between the vapor and the liquid in the separator. Further, this equilibrium should not be disturbed during the sampling. The equilibrium constant K is given by the equation:

$$K = \frac{\text{mole \% gas comp.}}{\text{mole \% press. oil comp.}} \quad (16)$$

at test separator conditions. We define a new function, F, given by:

$$F = b \left(\frac{1}{T_b} - \frac{1}{T} \right) \quad (17)$$

F is a free energy function, $b = \frac{L}{R}$, where L is the heat of vaporization for the pure component, and R is the gas constant.¹³ We then have the relation:

$$\ln K_p = b \left(\frac{1}{T_b} - \frac{1}{T} \right) \quad (18)$$

p is the test separator pressure in psia.

The plot of $\ln K_p$ against F shall give a straight line. If not, either the analysis or the sampling is in error. Figure 7 shows this plot for a good analysis, while Figure 8 shows the plot for an analysis which is obviously in error. The C7+ is off the line because of the lack of appropriate F value for C7+. In the plots it has been arbitrarily set equal to zero. Thus this method controls the component analysis up to C7+ or higher if F values and component composition is available.

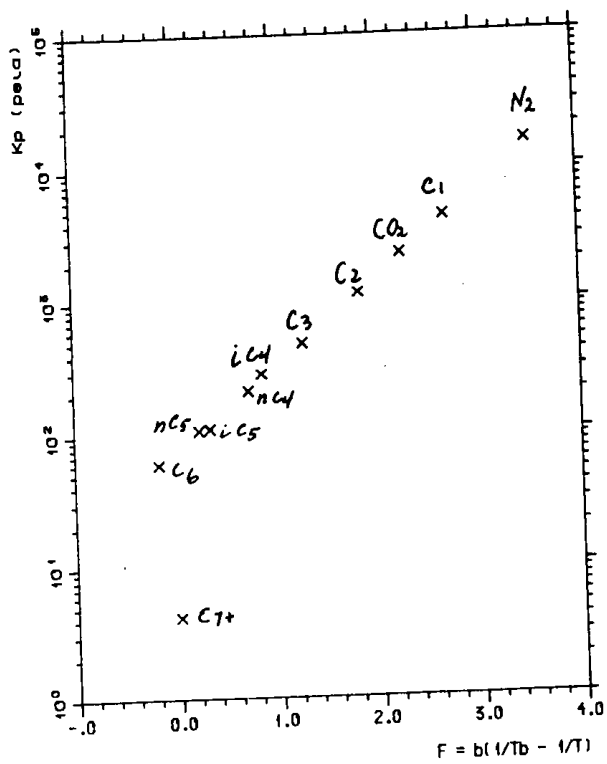


Figure 7. $\log K_p$ plotted as a function of F .

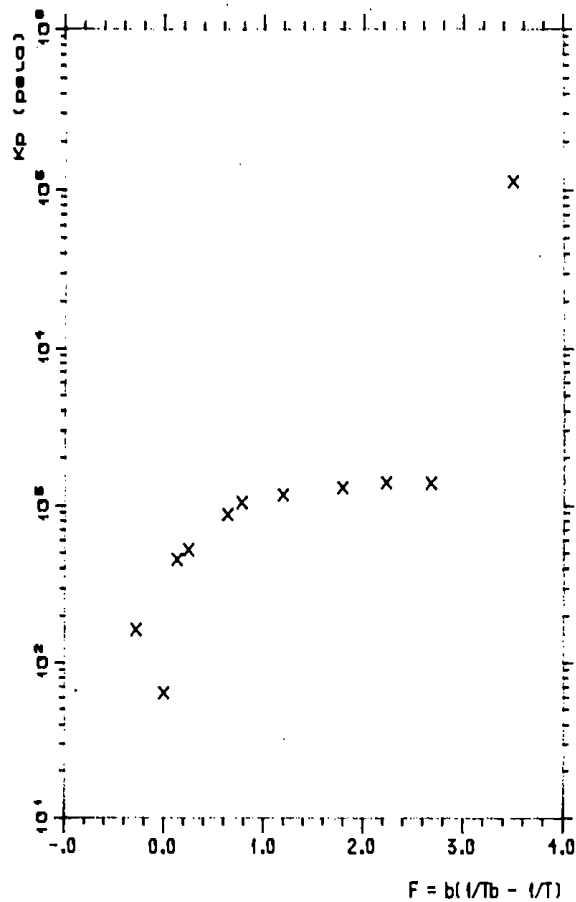


Figure 8. $\log K_p$ plotted as a function of F . The curvature of the plot indicates erroneous results.

Viscosity

For measurements of oil and gas the viscosity of the fluid is of importance. The viscosity coefficient is a measure of the resistance to flow exerted by the fluid. The dynamic or absolute viscosity, μ , of a Newtonian fluid is defined as the ratio of the shear force per unit area to the local velocity gradient. The kinematic viscosity is defined as:

$$\text{kinematic viscosity, } \nu = \frac{\text{dynamic viscosity, } \mu}{\text{density, } \rho}$$

The only way to obtain the accurate viscosity of a fluid is to measure it experimentally. However, experimental determinations are difficult and slow so usually the petroleum engineer must rely on viscosity correlations.

The viscosity of a pure gas depends on the temperature and pressure, but for gas mixtures it is also a function of the composition of the mixture. The following equation, initially proposed by Hering and Zipperer¹⁴, may be used to calculate the viscosity of a mixture of gases when the composition of the gas mixture is known, and the viscosities of the components are known at the pressure and temperature of interest.

$$\mu_g = \frac{\sum_i (\mu_{gi} y_i \sqrt{M_i})}{\sum_i (y_i \sqrt{M_i})} \quad (19)$$

The viscosity of gases is strongly depending on pressure only in certain regions of pressure and temperature. Usually, pressure variations are not significant at very high reduced temperatures or low reduced pressures.

The effect of pressure on viscosity is perhaps best seen from Figure 9. Here the viscosity has been reduced by dividing by the value at the critical point¹⁵.

At low reduced pressure we can see that except near the saturated vapor state there is little effect of pressure. The lower limit of the P_r curves indicates a dilute gas state. In such a state the viscosity increases with temperature. At high reduced pressures we see that there is a wide range of temperatures where the viscosity decreases with increasing temperature. In this region the viscosity behavior more closely simulates a liquid state, and an increase in temperature results in a decrease in viscosity. Finally, at very high reduced temperatures, there is but little effect of pressure on gas viscosity.

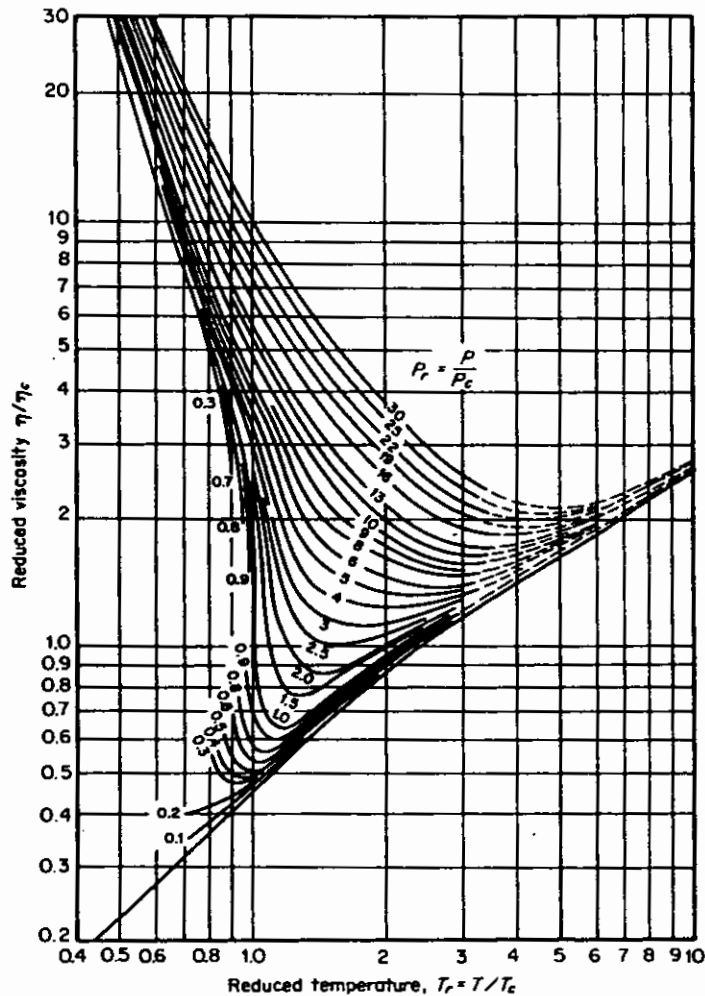


Figure 9. Generalized reduced gas viscosities.

Childs and Hanley¹⁶ have deduced criteria which indicate whether or not the pressure effect is significant. Their results are summarized in Figure 10. For any given reduced temperature and pressure, one can determine whether the gas is "dilute" or dense (so that pressure correction should be applied). The dividing line is located so that

the necessary dense gas correction is 1 percent or less.

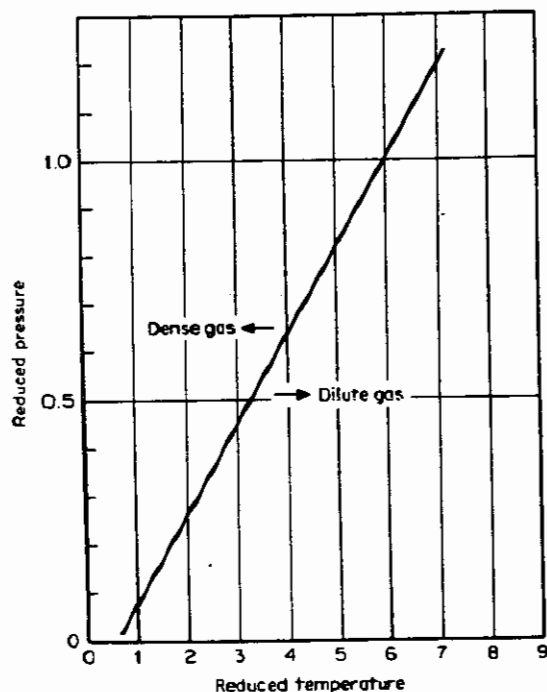


Figure 10. Ranges of reduced pressure and temperature for separating dilute and dense gases.

For natural gases, Carr, Kobayashi and Burrows¹⁷ have correlated gas viscosity to molecular weight, temperature, and pressure:

$$\mu_0 = f(M, T)$$

$$\frac{\mu}{\mu_0} = f(P_R, T_R)$$

Figure 11 gives the Carr et al. correlation for viscosity of paraffin hydrocarbon gases at 1 atm. pressure. The inserts are corrections for the presence of N₂, CO₂, and H₂S. The effect of each of the non-hydrocarbon gases is to increase the viscosity of the mixture.

In most cases, the viscosities must approach pressures far removed from 1 atm. The theorem of corresponding states has been used to develop the correlations given in Figures 12 and 13. Figure 12 gives the viscosity ratio as a function of reduced temperature and Figure 13 gives the viscosity ratio as a function of reduced pressure.

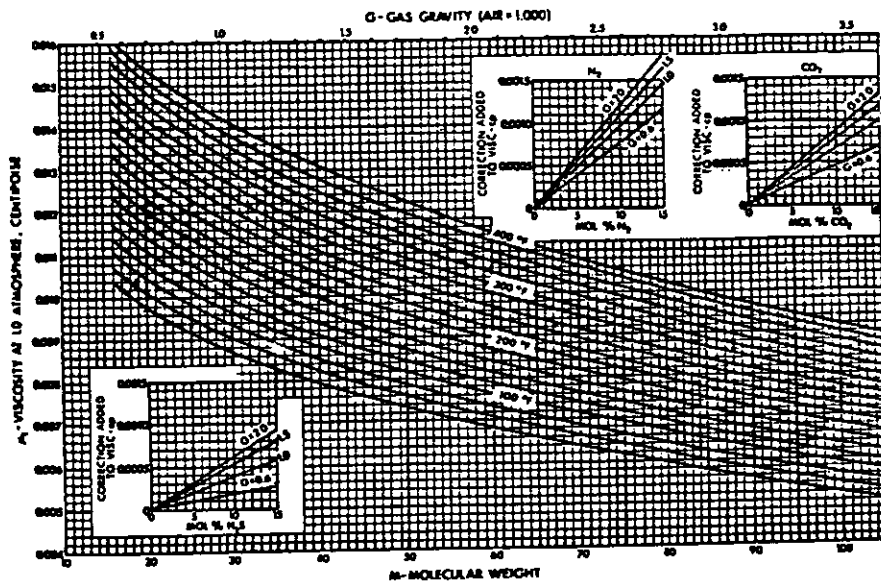


Figure 11. Viscosity of paraffin hydrocarbon gases at 1 atm.

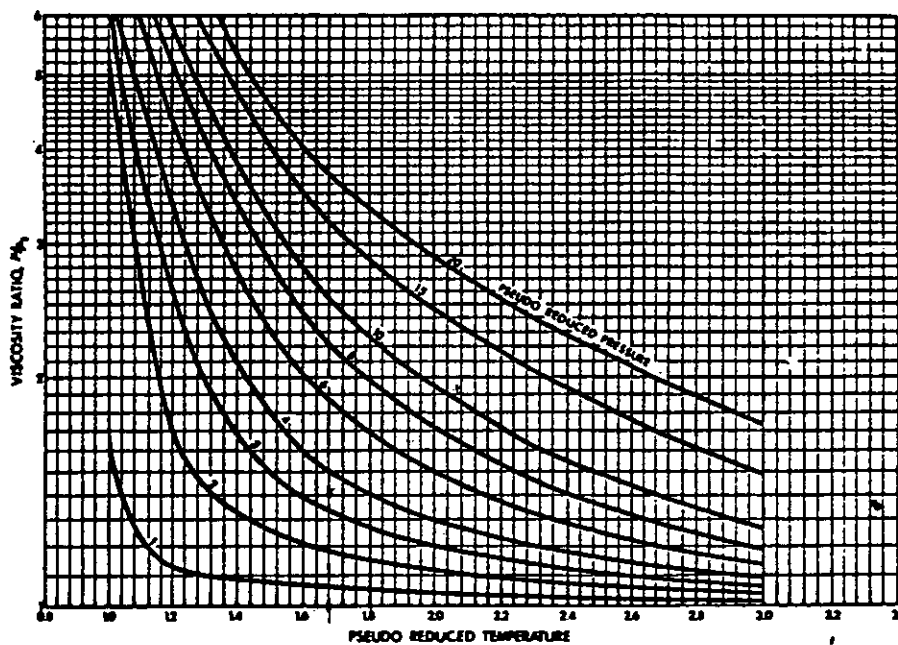


Figure 12. Viscosity ratio vs. pseudoreduced temperature

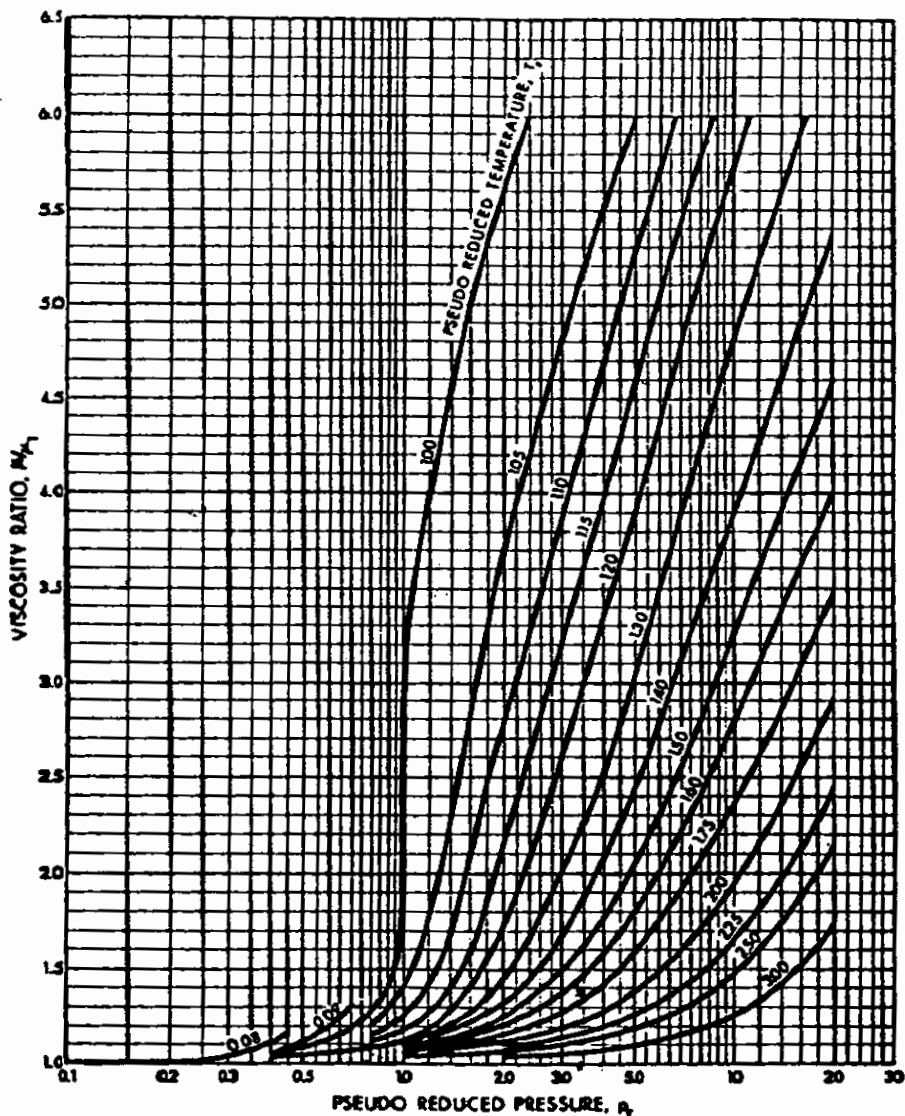


Figure 13. Viscosity ratio vs. pseudoreduced pressure.

Liquid viscosity.

As a rule, the liquid viscosity decreases with increasing temperature, and increase with increasing pressure. The prediction of viscosity is somewhat analogous to the prediction of density. One can use equations of state and combination rules or more empirical correlations based on physical parameters like molecular weight, EMR etc. The use of equations of state are strongly dependant on the access to computers.

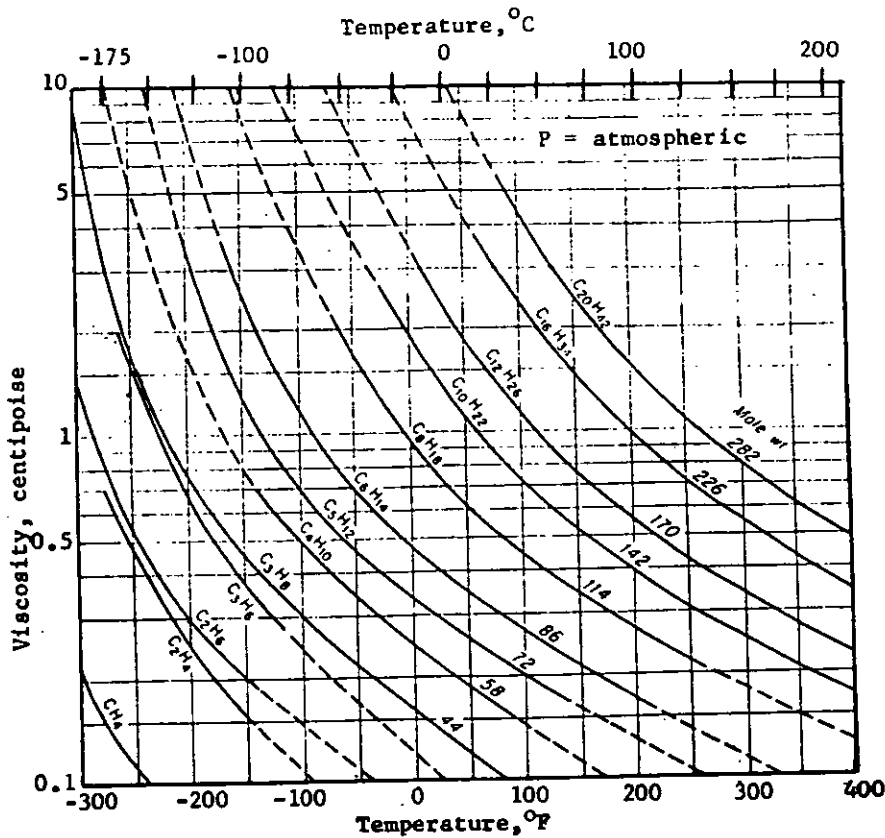


Figure 14. Viscosity of pure hydrocarbon liquids.

Figure 14 shows the viscosity, at atmospheric pressure, for common light hydrocarbons, as a function of temperature¹⁸. The viscosity of liquid mixtures can be calculated using the formula:

$$\mu_m = \left[\sum_i x_i (\mu_i)^{1/3} \right]^3 \quad (20)$$

where μ_m = viscosity of mixture in cp

μ_i = component viscosity in cp

x_i = mole fraction of each component in the mixture

The viscosity of a crude oil possessing an API gravity greater than 30°API (less than 0.88 specific gravity) can be estimated by the equation:

$$\log \mu = a - 0.035(^{\circ}\text{API}) \quad (21)$$

$$\mu \text{ is in cp, } ^{\circ}\text{API} = \frac{141.5}{\text{sp. grav.}} - 131.5$$

$^{\circ}F$	100	130	160	190	220
a	2.05	1.83	1.55	1.30	1.08

Eykman Molecular Refraction (EMR)¹⁹ is related to the refraction index of a liquid and can readily be measured. EMR is given by the equation:

$$EMR = \left[\frac{(n^2 - 1)}{(n + 0.4)} \right] \cdot (Mw/p) \quad (22)$$

n = refraction index

For normal hydrocarbons it has been found that:

$$EMR = 2.4079 + 0.7293M + 0.00003268M^2 \quad (23)$$

Figure 15 shows the relationship between liquid viscosity and EMR²⁰.

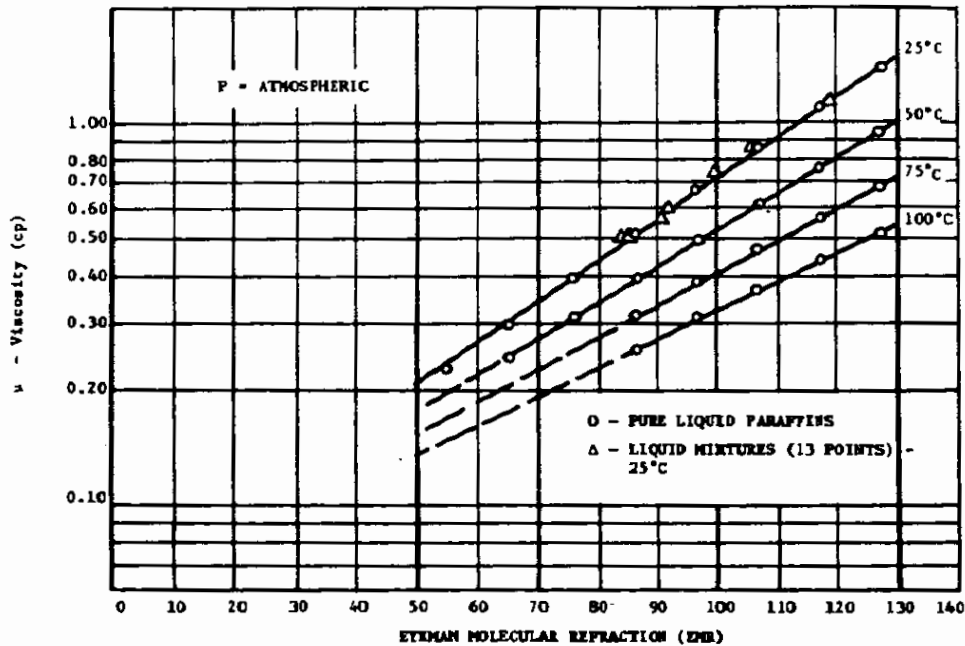


Figure 15. Viscosity vs. Eykman Molecular Refraction (EMR).

As for gases, liquid viscosity often has to approach high pressures. Figure 16 enables us to calculate liquid viscosity at elevated pressure and temperature²¹. P_r , T_r , and apparent molecular weight are determined as given by equations 3 - 6.

Measurement and physical properties.

By far the most common type of differential meter used in the fuel gas industry is the orifice meter. For the calculation of the quantity of gas AGA committee Report no. 3 recommends the formula:

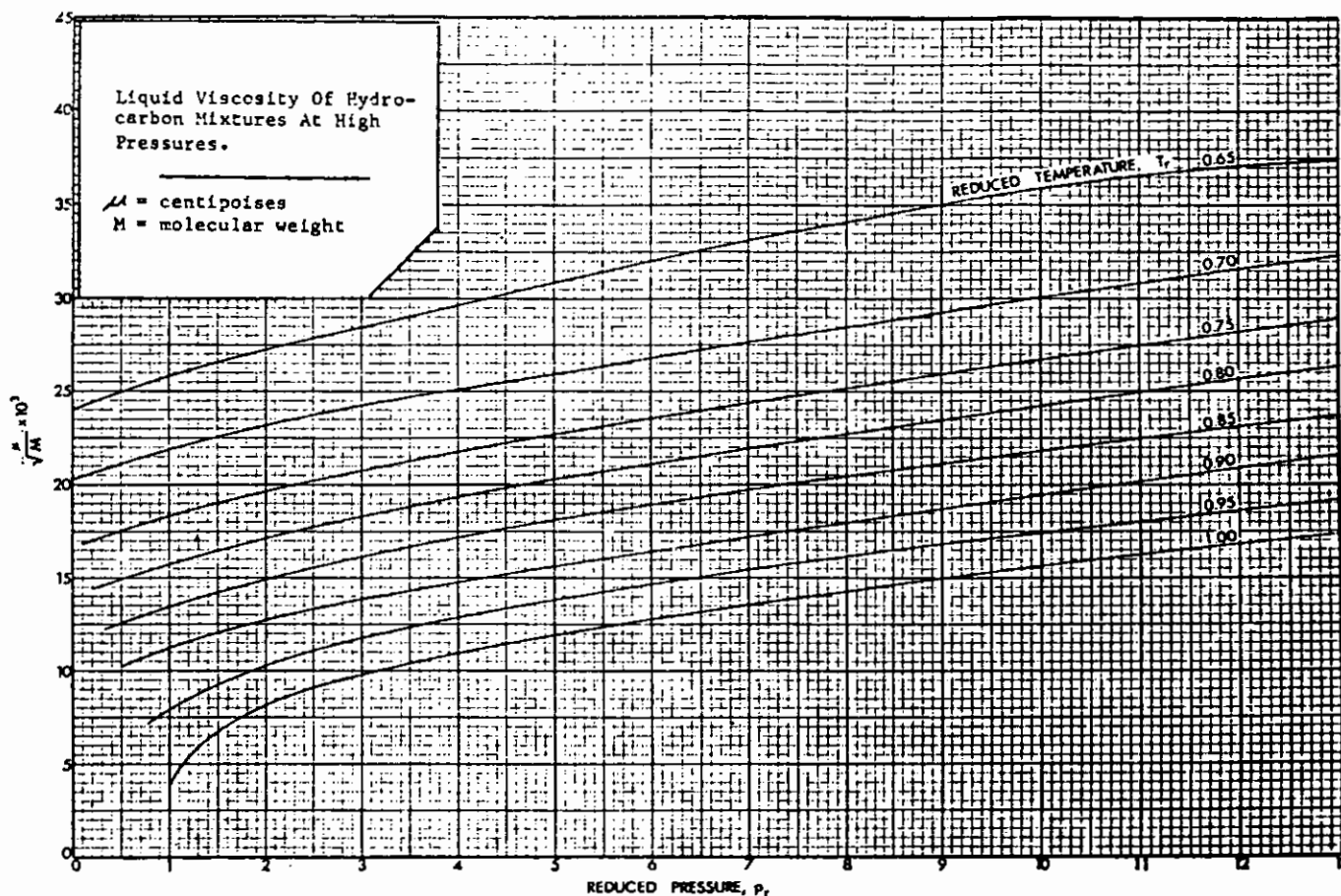


Figure 16. Liquid viscosity of hydrocarbon mixtures at high pressures.

$$Q_h = C' \sqrt{h_w p_f} \quad (24)$$

here Q_h = quantity rate of flow at base conditions, ft^3/hr

C' = orifice flow constant

h_w = differential pressure in inches of water at 60°F

p_f = absolute static pressure, psia

$\sqrt{h_w p_f}$ = pressure extension

Because the general orifice meter equation appears to be so simple, one may wonder where all these physical laws become involved in the measurements calculations. The orifice flow constant C' may be defined as the rate of flow in ft^3/hr , at base conditions, when the pressure extension equals unity. C' is obtained by multiplying a basic orifice factor F_b , by various correcting factors that are determined by the operating conditions, contract requirements, and physical nature of the installation. This is expressed in the following equation:

$$C' = F_b \cdot F_r \cdot Y \cdot F_{pb} \cdot F_{tb} \cdot F_{tf} \cdot F_g \cdot F_{pv} \cdot F_m \cdot F_l \cdot F_a \quad (25)$$

- F_b = basic orifice factor, ft^3/hr
- F_r = Reynold's number factor (viscosity)
- Y = expansion factor
- F_{pb} = pressure base factor (contract)
- F_{tb} = temperature base factor (contract)
- F_{tf} = flowing temperature factor
- F_g = specific gravity factor
- F_{pv} = supercompressibility factor
- F_m = manometer factor for mercury meter
- F_l = gauge location factor
- F_a = orifice thermal expansion factor

The derivation of some of these factors is very complex. Actually, several factors can be determined only by very extensive tests and experimentation, from which data have been accumulated so that a value may be obtained.²²

The factor F_b depends on design and is a constant for a specific installation.

The Reynolds number factor F_r depends on the pipe diameter, and the viscosity, density, and velocity of the gas

$$F_r = 1 + \frac{b}{\sqrt{h_w \rho_f}} \cdot 3.443 \mu_{cp} \sqrt{\frac{T_f}{G}}$$

The expansion factor Y : unlike liquids, when a gas flows through an orifice, the change in velocity and pressure is accompanied by a change in the density. The expansion of the gas is essentially adiabatic. Under these conditions, the density of the stream changes because of the pressure drop and the adiabatic temperature change. The expansion factor Y , computed for the adiabatic and reversible case is included to correct for this variation in density.

The specific gravity factor F_g is used to correct for changes in the specific gravity and should be based on actual flowing specific gravity of the gas. To make the basic orifice factor usable for any gas, the proper correction for the specific gravity of the gas must be applied. F_g is related to G by the equation:

$$F_g = \frac{1}{\sqrt{G}}$$

Supercompressibility factor F_{pv} : This factor corrects for the fact that gases do not follow the ideal gas law. It varies with temperature, pressure, and specific gravity. The development of the general hydraulic flow equation involves the actual density of the fluid at the point of measurement. In the measurement of gas this depend on the flowing pressure and temperature. To convert the calculated volume at the flowing pressure and temperature to the base pressure and temperature one has to rely on the ideal gas law. As discussed before, all gases deviate from this law to a greater or lesser extent. A factor to account for this deviation is necessary in the measurement of gases. The factor is particularly important at high line pressure. The following equation is used to determine the supercompressibility factor:

$$F_{pv} = \frac{Z_b}{Z}$$

Where Z_b = gas deviation factor at base conditions
 Z = gas deviation factor at line conditions.

Accurate knowledge of the physical properties of petroleum fluids are fundamental for proper production, processing, and transportation of the fluids. It is also seen that accurate knowledge of composition, Z-factor, viscosity, and density, among many other factors, play an important role to improve measurement reliability.

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[1] Paper presented at the North Sea Flow Measurement Workshop, a workshop arranged by NFOGM & TUV-NEL

Note that this reference was not part of the original paper, but has been added subsequently to make the paper searchable in Google Scholar.