

Poster Paper

The Impact of the Use of Generic Crude Oil Expansion Coefficients to Correct Fluid Densities to Other Conditions

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1 INTRODUCTION

In the North Sea and elsewhere, many oil companies make use of shared pipeline systems to transport oil from the wells to shore. Each operator measures the volumetric flow rate of oil from their well to the shared pipeline. To enable correct allocation of oil to each operator once it reaches the refinery, it is necessary to know the effects of temperature and pressure on the volume of oil, i.e. the variation of density as a function of temperature and pressure. Current practice is based on measuring the density of the oil at process conditions, using a densitometer, and then using generic crude oil expansion coefficients to correct fluid densities to other conditions. However, there are concerns about the calibration procedures for densitometers and the applicability of current crude oil volume correction procedures.

TUV NEL is currently running a Joint Industry Project aimed at addressing issues associated with the characterisation and traceable calibration of densitometers used in the North Sea oil industry. However, even with perfect densitometer operation, errors will be introduced by the use of inappropriate coefficients to correct the density at densitometer conditions to that at base conditions and then to that at meter conditions, as expansion coefficients for North Sea crude oils have never been adequately defined. The validity of using generic expansion coefficients on crude oils containing substantial amounts of water or natural gas liquids is highly questionable since the tables and procedures were developed for dry, stock tank crude oils.

Whilst the issues surrounding the use of generic expansion coefficients have been known for many years [1], they are only becoming of significant concern now that large oil companies are selling off fields to smaller operators and the issues of shared pipeline systems are becoming more apparent. However, these issues are not confined to the North Sea nor to pipelines with densitometer-based systems; on-shore pipelines and tanker transportation of hydrocarbons also currently rely on generic expansion coefficient correction methods.

It is even more important now to define the consequences of using the standard tables in North Sea accounting, even if the tables continue to be the basis for trade. The increasing use of uncertainty analyses in allocation contracts leaves the uncertainty in the use of the tables as an assumed value, one which increases as metering conditions change. From a comparison between the results predicted by a more physically correct liquid mixture density model and current volume correction procedures, it was found that temperature and pressure have a significant effect on the applicability of current volume correction procedures. It has become clear that more detailed and physically realistic methods must be used for volume correction to prevent misallocation errors, as this paper demonstrates.

2 VOLUME CORRECTION FACTORS

The density, and therefore the volume, of hydrocarbons is a function of temperature and pressure. Within the oil industry it is common practice to use Volume Correction Factors (VCFs) to correct observed volumes to equivalent volumes at a standard temperature and pressure. Measurements corrected to standard, or base, conditions ensure that the volumetric measures are equitable in commerce.

VCFs, in their basic form, are the output of sets of equations derived from empirical data relating to the volumetric change of hydrocarbons over a range of temperatures and pressures. Correction factors to account for the thermal expansion of liquid hydrocarbons were first formally developed in 1916 by the United States National Bureau of Standards (see "Introduction and History" in [2]). Traditionally the factors have been listed in tabular form for various commodity types (crude oil, refined products, lubricating oils and special applications).

2.1 The Petroleum Measurement Tables

In 1952 the American Society for Testing and Materials (ASTM) and The Institute of Petroleum (IP) jointly published a series of tables, The Petroleum Measurement Tables, commonly referred to as the 1952 Tables or "Blue Book Tables". These contained many sets of correction and conversion factors tables used in the measurement of hydrocarbon liquids but dealt only with the effects of temperature. Pressure corrections were made separately, using an American Petroleum Institute (API) standard (API Standard 1101, Appendix B, Table II).

In 1965 the API adopted the 1952 Tables and in 1974 initiated a programme of measurement and modelling to revise the Tables, leading to the 1980 Tables. Although an extensive measurement programme was carried out, on 349 fluid samples, just over one third (124) were crude oils, covering a density range from 770 to 990 kgm⁻³ at temperatures between 4 and 56°C. As with the 1952 Tables, the 1980 Tables only dealt with the effects of temperature, the API 1101 Standard still being used for the effects of pressure.

The API 1101 Standard was based on limited data, mostly on pure compounds and lubricating oils. In 1981 a revised version of the standard was issued (Chapter 11.2.1 and 11.2.1M, covering customary and metric units respectively). However, despite an extensive literature survey, only three sources of compressibility data were found [2].

Changes in operational procedures and requirements in the petroleum industry between 1980 and the mid-1990s led to the requirement for further revision to the Tables, culminating in the 2004 edition (API MPMS Chapter 11.1-2004/Adjunct to IP 200/04/Adjunct to ASTM D 1250-04 (ADJD1250CD) [2]). Although numerous changes have been made to the standard, these are mostly to do with the calculation procedures. In particular, including the pressure correction in the standard represents an important change from the "temperature only" correction factors given in the 1980 Petroleum Measurement Tables. Furthermore, the inputs to all the procedures are now in terms of conventional units (°F and psig).

The combination of density and volume correction factors for both temperature and pressure is collectively referred to in the standard/adjunct(s) as a Correction for Temperature and Pressure of a Liquid (C_{TPL}). The temperature portion of this correction is termed the Correction for the effect of Temperature on Liquid (C_{TL}), also historically known as VCF (Volume Correction Factor). The pressure portion is termed the Correction for the effect of Pressure on Liquid (C_{PL}).

Although the temperature, pressure and density limits have been extended (crude oil base density at 15°C between 611.2 and 1163.8 kgm⁻³, temperatures between -50 and 150°C and pressures between 0 to 103.4 bar gauge), no new data have been incorporated into the current version of the standard. Any limitations arising from the restricted range of data used in deriving the original correlations will therefore remain.

2.2 Volume Correction Factor Calculation Procedure

The current version of the Standard [2] includes a number of methods for calculating VCFs for various applications. For the case of calculating a VCF from a density at base conditions (60°F and 0 psig) to alternate temperature and pressure conditions, a two-part procedure is used:

1. A thermal correction is applied to the liquid to account for the change from the base temperature (60°F) to the alternate temperature t (in °F) at a constant base pressure.
2. A pressure correction is applied to the liquid to account for the change from the base pressure (0 psig) to the alternate pressure P (in psig) at the alternate temperature t .

The correction factor due to temperature, C_{TL} , is calculated from

$$C_{TL} = \exp[-\alpha_{60}\Delta t(1 + 0.8\alpha_{60}\Delta t)] \quad (1)$$

where $\Delta t = t - 60$. The coefficient of thermal expansion at the base temperature of 60°F, α_{60} (in °F⁻¹), is calculated from

$$\alpha_{60} = K_2 + \frac{K_1 + \frac{K_0}{\rho_{60}}}{\rho_{60}} \quad (2)$$

where ρ_{60} is the density at base conditions (60°F and 0 psig) in kg m⁻³. The coefficients in equation 2 depend on the commodity group. For crude oils $K_0 = 341.0957 \text{ kg}^2 \text{ m}^{-6} \text{ °F}^{-1}$, $K_1 = 0 \text{ kg m}^{-3} \text{ °F}^{-1}$ and $K_2 = 0 \text{ °F}^{-1}$. The scaled compressibility factor, F_p , (in psi⁻¹) is given by

$$F_p = \exp\left(-1.9947 + 0.00013427t + \frac{793920 + 2326.0t}{\rho_{60}^2}\right) \quad (3)$$

and the correction factor due to pressure, C_{PL} , is calculated from

$$C_{PL} = \frac{1}{1 - 10^{-5} F_p P} \quad (4)$$

The VCF or combined temperature and pressure correction, C_{TPL} , is given by

$$C_{TPL} = C_{TL} C_{PL} \quad (5)$$

allowing calculation of the volume at alternate conditions from

$$V = \frac{V_{60}}{C_{TPL}} \quad (6)$$

In the above procedure, for any given crude oil the only parameter which is specific to that oil is its density at base conditions, ρ_{60} , since the coefficient of thermal expansion, α_{60} , is a generic one applicable to all crude oils within the allowed density range.

Within the UK and elsewhere, it has become common practice to incorporate the above procedure into commercial contracts and flow computer routines. However, within the context of the North Sea, the UK oil and gas regulator states that [3]

“3.2.5 It is essential, however, that the values of K_0 and K_1 input into the flow computer for use in the calculations are representative of the type of oil being measured.”

Furthermore, the Guidance Notes [3] make it clear that there are limitations with the use of generic values for the coefficients:

“3.2.6 Operators are encouraged to consider the use of specific constants derived for the particular application by laboratory analysis of representative samples.

It should be borne in mind that the ‘generalised’ values of K_o and K_l referred to in 3.2.5 were derived from the analysis of a relatively small sample of stabilised, non-North Sea, crude oils.”

In the derivation of the coefficients in the current API / IP / ASTM standard [2], the only data specific to North Sea crude oils were for two samples taken in the mid-1970s. Since then, in common with many mature oil fields, North Sea crude oils have tended to become more viscous and their water content has increased. Furthermore, as part of the measurement procedure used to determine the density of each crude oil in the 1974-79 API-funded programme, the samples were allowed to stabilise in an open container before testing; this may have affected the composition of the crude by vaporisation of the light hydrocarbon fractions. The standard also has a lack of dependence on thermophysical properties and crude oil composition - the database used in the derivation included density data but did not reflect viscosity, molecular weight or any other characterisation parameter.

3 CRUDE OIL CHARACTERISATION AND MODELLING

Taken together, the limitations outlined in Section 2 highlight the need for better methods to characterise crude oils and the use of more physically realistic models to predict the effects of temperature and pressure on their volumetric properties.

3.1 Crude Oil Characterisation

Crude oils are complex mixtures of hydrocarbons. Although there is no such substance as a standard oil, most are comprised of general hydrocarbons (typically 80-85% carbon, 10-15 % hydrogen), with small amounts of sulphur, inert gases, salts and metallic compounds. Depending on the process, there will be varying amounts of water and oxygenated compounds. A full understanding of the effects of temperature and pressure on the properties of a crude oil (in this context in particular its volumetric properties) would therefore require a knowledge of its composition.

In chemical process engineering the problem of evaluating the properties of mixtures is generally tackled by assuming that it is possible to accurately identify the individual components in the mixture and also their relative amounts (i.e. the mixture composition). However, this approach is not practical for crude oil and its derivatives. To characterise this type of fluid, an alternative approach is used which employs the concept of a ‘pseudo-component’. Having represented the ‘oil’ as one or more pseudo-components, normal chemical thermodynamic methods can be applied to calculate phase equilibrium and physical properties.

This has led to a number of developments, including the establishment of approved standards, for the measurement, characterisation and calculation of the phase behaviour and physical properties for such systems.

Unlike a defined component which is a single compound with clearly defined constant properties such as molecular weight and normal boiling point, a pseudo-component is typically a mixture with a boiling range, but considered as a single compound with some ‘average’ property value representing the molecular weight and normal boiling point.

A key input parameter for many of the property calculation methods is an average boiling point. Whilst there are a number of 'average' boiling points (the most common of which are volume, weight, molar, cubic and mean), the most convenient method for determining an average boiling point is via a distillation curve.

A distillation curve is usually produced as the result of a laboratory experiment, typically according to an approved standard issued by ASTM. The basic experiment determines the temperature at which prescribed amounts (volume percentage) of the oil have been boiled off from the reservoir. The minimum data set is at 10, 30, 50, 70, 90% points. Determination of the initial boiling point is fairly straightforward, but problems can occur in determining the final boiling point, particularly for heavy oils, due to thermal cracking (decomposition).

From this curve it is straightforward to get the average volume boiling point. The other boiling points can be obtained from correlations (either graphical charts or software). Correlations also exist for inter-conversion of the complete curves from one form to another.

The ASTM standards are relatively quick and cheap to undertake. D86 is generally used for light to medium oils and D216 is used for gasolines. Both of these methods are usually at atmospheric pressure. D1160 is typically used for heavy oils and depending on the oil may be required to be run at low pressure (down to 10 mmHg).

Much more use is now made of analytical methods using gas chromatography (GC) and mass spectroscopy (MS), although these need additional calibrations using a set of defined components. In many situations crude oil samples are routinely analysed by GCMS, providing, in principle, the data necessary to allow the use of physically realistic models that capture the complexity of such fluids.

3.2 Thermophysical Property Modelling

As noted in Section 3.1, crude oils are complex mixtures of hydrocarbons and hence models for predicting their thermophysical properties must be capable of describing both the phase behaviour and the thermodynamic properties of the mixture.

Modelling methods for prediction of the phase behaviour of fluid systems have tended to concentrate on answering the questions

- How much of a phase is present?
- What are the compositions of the components in a phase?

Some of the models are capable of delivering the thermodynamic properties of the individual phases in addition to phase behaviour. Many of the common models fall in to the category of equations of state. An equation of state (EOS) provides a mathematical relationship between the pressure, temperature and volume (or density) of a fluid:

$$\phi(P, V, T) = 0 \quad (7)$$

There are a couple of advantages to using an EOS method for obtaining thermodynamic properties:

- The properties derived from an EOS are self-consistent from a thermodynamic viewpoint.
- The equations cover a wide range of conditions in temperature and pressure.

On the minus side, the equations themselves are mathematically complex and are not explicit in volume or density and so must be solved by computer with iterative methods. Since the equations are usually couched in terms of V^3 or higher order, there is also the problem of returning the correct root for a given input of T and P . The methods were originally developed to

solve phase equilibria problems and whilst many of them give good results for phase splits and resulting phase compositions, they can have short-comings in the liquid phase, especially for liquid volume calculations.

For complex liquid mixtures such as crude oils, a much better approach is to use the COSTALD (corresponding states liquid density) method [4]. This method explicitly relates the liquid volume of a pure component to its reduced temperature and a second parameter termed the characteristic volume. It has been adopted as an API standard [5] and is widely used in process simulation software packages such as PRO II, HYSYS and Aspen Plus.

Relatively easy to use and applicable to a wide variety of liquids, the COSTALD method is flexible and consistent, requiring only reduced temperature, acentric factor, and a characteristic volume for each pure compound. The method includes a set of rules to allow it to handle mixtures.

When applied to liquid mixtures at elevated pressures, COSTALD is implemented as the Tait-COSTALD or HBT (Hankinson-Brost-Thomson) method [6]. Using only critical temperature, critical pressure, Soave-Redlich-Kwong acentric factor, saturation pressure, and saturated volume for a pure liquid or the corresponding mixture parameters for a liquid mixture, the method has been found to be accurate for temperatures in the range -223°C to 327°C and pressures ranging from saturation pressure to 689.5 bar [6]. Keskinen and Aalto [7] carried out research on the parameters required for a variety of compressed mixture density models for many different compounds and found the Tait-COSTALD method to be "one of the most accurate and most general models for saturated liquid density prediction".

3.3 PPDS Petroleum Fractions Package

When calculating the effects of temperature and pressure on fluid volume for crude oils, the COSTALD method would be expected to be more accurate than simple, generic correlations that do not account properly for fluid composition. This premise has been tested by making use of an implementation of the COSTALD method developed by TUV NEL Ltd.

TUV NEL's Physical Property Data Service (PPDS) is a software suite used extensively in the chemical production, oil & gas, pharmaceutical, fine chemical, power generation and process industries. It allows users to perform calculations and simulations across a wide range of process and engineering applications. Backed by a comprehensive package of databases, PPDS includes key information on more than 1,500 chemicals, constant and variable physical properties, thermodynamic model sets and flash calculations. A number of special packages including equations of state and a petroleum fractions module are also available.

The PPDS Petroleum Fractions package has been developed to estimate the physical properties of a mixture when detailed composition information is not known, or too impractical to obtain. It uses pseudo-components to characterise fluid mixtures and an extension of the COSTALD method to calculate the density of hydrocarbon mixtures.

The minimum input requirements into the PPDS Petroleum Fractions package for the calculation of liquid mixture densities are distillation curve information and API/ specific gravity of the mixture. To ensure greater accuracy in liquid density predictions it is also possible to enter viscosity, Watson K-factor, and average molecular weight data. The critical temperature, critical pressure, and acentric factor for the mixture are estimated from these data for input into the Tait-COSTALD model.

To provide a check on the accuracy of the COSTALD implementation within the PPDS Petroleum Fractions package, it was used to calculate densities for a range of diesel fuel samples at temperatures between 25 and 75°C at pressures up to 2000 bar. Although a diesel fuel is not as complex as a full crude oil, it is nevertheless a very good test fluid since it contains a wide range of hydrocarbon components, typically from C₉ to C₂₄. Experimental density data for the fuel

samples were measured using a MicroPVT apparatus. This equipment uses a variant of the piston-in-bottle method to determine the pressure developed as the volume of a sample of liquid is changed by forcing a piston in to a closed bottle [8-9]. For the fluid samples used in this study, the overall uncertainty in elevated pressure density was estimated at 0.23% (at k=2) [9].

Full, reduced pressure and simulated distillation curves were obtained for the fuel samples, to provide the necessary inputs to the PPDS Petroleum Fractions package. Each fuel sample was then modelled as a six pseudo-component mixture and the results compared with the experimental data. Liquid densities were also estimated using the Lee-Kesler-Plöcker (LKP) and the Peng-Robinson (PR) equations of state, based on the full, reduced pressure and simulated distillation curves. Figure 1 shows the deviations between the calculated and experimental densities for one fuel sample at 25°C. Results for the other samples are similar, with the PPDS Petroleum Fractions package values typically having an average absolute deviation of less than 0.05% from the measured values.

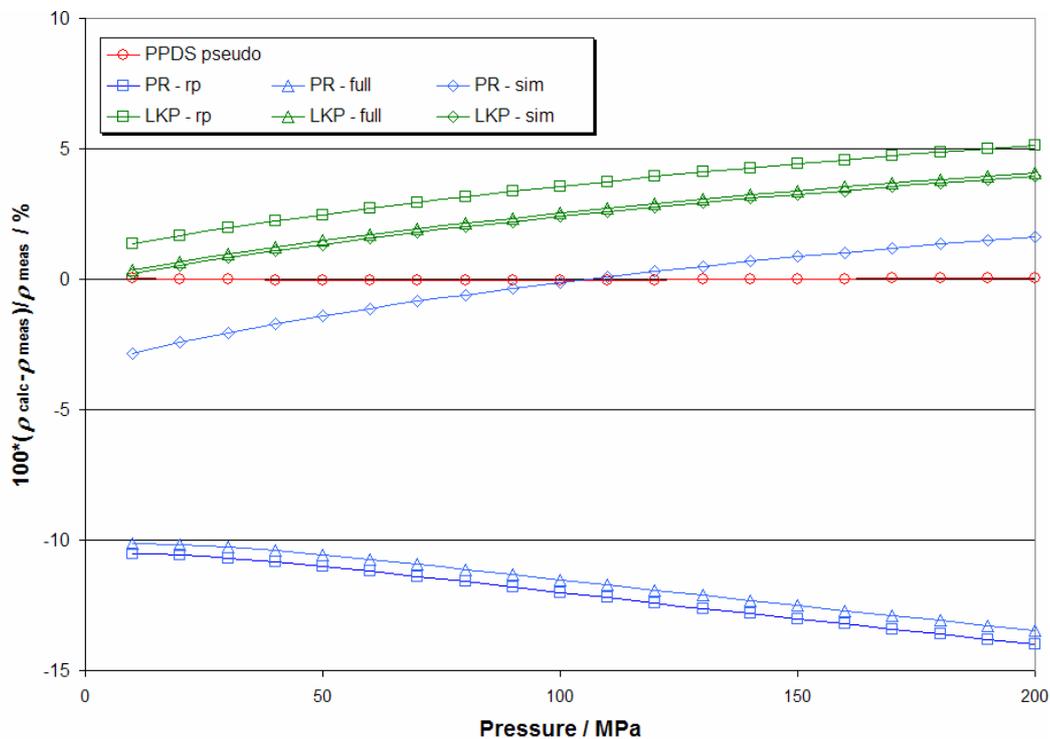


Fig 1. Error in calculated density for various models

Similar experience with other fluid samples, its use in a range of process simulators and incorporation into an API standard [5] reinforce the view that the Tait-COSTALD method should be capable of more accurately representing the effects of temperature and pressure on crude oil volume than a simple expansion coefficient method with generic coefficients that do not take account of oil composition.

4 RESULTS

To provide a proper examination of the impact of the use of generic crude oil expansion coefficients would require experimental data on the effects of temperature and pressure on fluid volume from base conditions (60°F and 0 psig or 15°C and 101325 Pa) across the full range up to the temperatures and pressures at the flow metering stations plus compositional information for the oils to be modelled. As noted previously, there is a dearth of data and such data as there are may not be truly representative [2]. However, an initial estimate of the magnitude of the problem can be obtained by comparison of volumes calculated from the API / IP / ASTM method and the Tait-COSTALD method.

Assay data were gathered for eight different North Sea crude oils for analysis using both the Tait-COSTALD and expansion coefficient methods. Crude oils with a wide range of API gravities and viscosities were chosen to give an extensive comparison of different North Sea reserves. The assay data were obtained from North Sea oil and gas operator Chevron; the names and physical characteristics of the crude oils are shown in Table 1.

Table 1 – North Sea Crude Oil Physical Characteristics

Name	Average Molecular Weight	Viscosity at 210°F cP	API Gravity
Alba	385	14.567	19.3
Brent	209	1.483	38.5
Captain	370	11.526	19.2
Draugen	197	0.970	40.8
DUC	210	2.245	34.8
Ekofisk	217	0.990	37.8
Forties Blend	183	0.976	42.1
Statfjord	207	1.322	39.4

For the initial analysis, each crude oil was characterised as a single pseudo-component within the PPDS Petroleum Fractions package. However, to show the effects of characterising the crude oils in a more detailed manner, the analysis was repeated by characterising each crude oil as a mixture of six different pseudo-components. It was anticipated that characterising the crude oils as a greater number of pseudo-components would give a more accurate representation of composition, and produce density results which would be even closer to experimental values.

From the detailed assay data for the crude oils shown in Table 1, a weight percentage average boiling point was calculated for each crude and input to the PPDS Petroleum Fractions package, (along with the average molecular weight, API gravity and viscosity at 210°F), to create the single pseudo-components. For the six pseudo-component mixtures, the detailed assay data were used to construct true boiling point distillation curves, based on the cuts (light naphtha, medium naphtha etc). In addition to the molecular weight, API gravity and viscosity at 210°F of each cut, the distillation curves were used to create the pseudo-components. Each crude oil was then modelled as a mixture of its six pseudo-components, based on the mole fraction of each cut in the crude.

As part of a current Joint Industry Project aimed at addressing issues associated with the characterisation and traceable calibration of densitometers used in the North Sea oil industry [10], TUV NEL surveyed operators to obtain information on typical densitometer operating conditions.

The results of the survey have been used to set the range of temperatures and pressures over which to carry out the comparison. Normal operating temperatures of North Sea densitometers are around 20 to 80°C but in extreme circumstances can be as low as 10°C or as high as 110°C. There is also a wide range of pressure conditions over which the densitometers must operate; from around 5 to 120 bar although most operate at pressures in the 20 to 40 bar region.

On the basis of the above data, three pairs of conditions were examined:

- low 10°C and 5 bar
- average 50°C and 30 bar
- high 110°C and 120 bar.

The phase equilibrium models within PPDS are capable of calculating the effects of adding water to a complex hydrocarbon mixture. Initial calculations indicated that, for the conditions examined, water contents of the order of a few percent by weight resulted in the formation of separate water and hydrocarbon phases. In practice, whilst the phases may be separate, the conditions at the densitometer are such that an “average” density will be measured. For this initial study it was therefore decided to only investigate the effects of temperature and pressure.

At each of these pairs of conditions the effect of varying one parameter while the other was held constant was investigated. Thus, for example, at the average condition, calculations were done for temperatures between 0 and 150°C at 30 bar and between 1 and 150 bar at 50°C.

For each temperature and pressure examined, the PPDS Petroleum Fractions package was used to calculate the volume of the crude oil at those conditions and the volume $V_{0,PPDS}$ at the corresponding base conditions (15°C and 101325 Pa). The expansion coefficient method was then used to calculate the volume $V_{0,API}$ of the crude oil at base conditions from the elevated temperature and pressure volume determined by the PPDS Petroleum Fractions package and the differences compared. In the following sections, the difference between the methods is expressed as

$$\delta = 100 \left(\frac{V_{0,API} - V_{0,PPDS}}{V_{0,PPDS}} \right). \quad (8)$$

4.1 Single Pseudo-component

Figure 2 shows the effect of increasing temperature at medium conditions. Similar results are obtained at low and high conditions. As the temperature is increased at any pressure, the deviations between the expansion coefficient and PPDS Petroleum Fractions methods increase, with significant differences between the standard volumes except at conditions very close to the base conditions.

Figure 3 shows the effect of increasing pressure at medium conditions, with similar results at low and high conditions. As the pressure is increased at any temperature, the deviations between the expansion coefficient and PPDS Petroleum Fractions methods increase.

It is worth noting that, except at conditions very close to the base conditions, there are always significant differences between the standard volumes calculated by the two methods. Furthermore, as the temperature or pressure is increased, two of the oils appear to exhibit anomalous behaviour.

26th International North Sea Flow Measurement Workshop
 21st – 24th October 2008

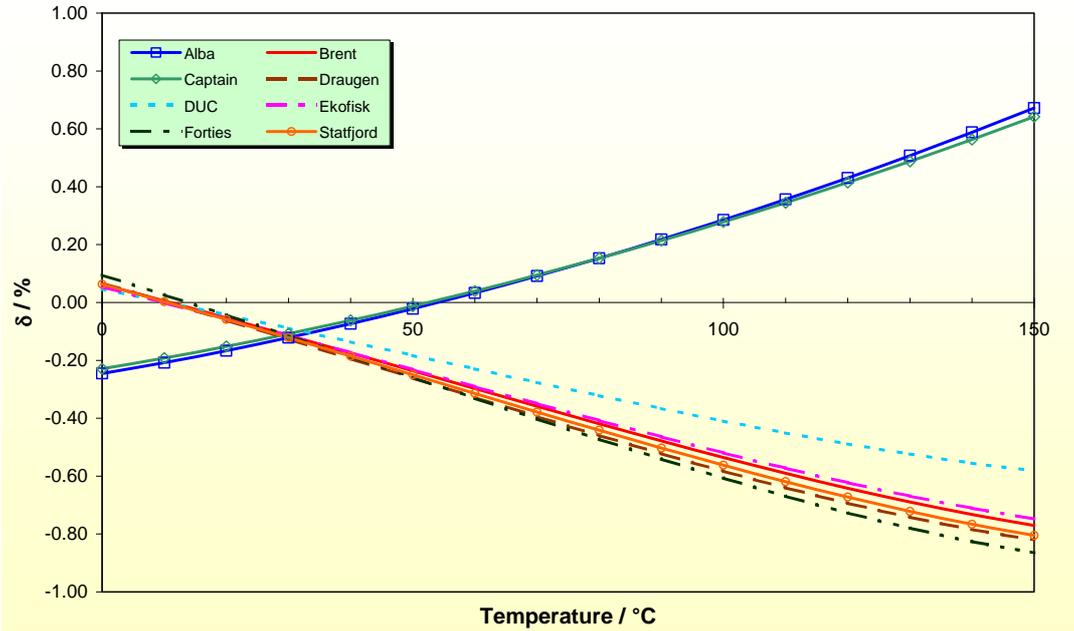


Fig 2. Volume deviation with increasing temperature at average pressure (one pseudo-component)

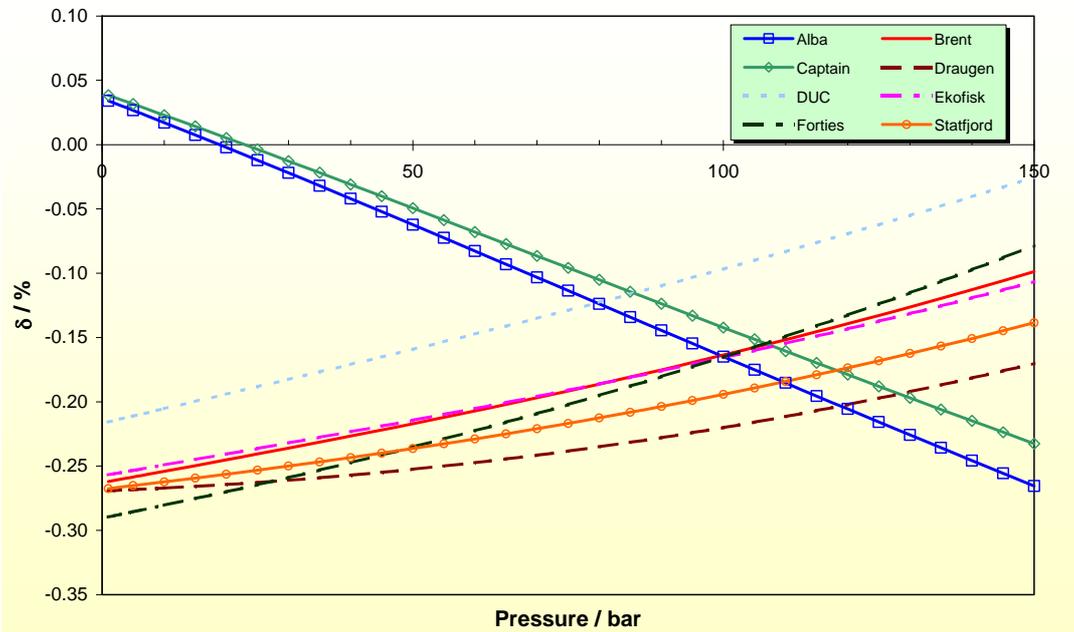


Fig 3. Volume deviation with increasing pressure at average temperature (one pseudo-component)

The two samples that exhibit anomalous behaviour are high molecular weight, high density, high viscosity crude oils. Attempting to characterise them as single pseudo-component fluids is unlikely to capture the complexity of the oils and it is likely that the calculations from both the expansion coefficient and PPDS Petroleum Fractions methods are equally wrong. Analysis of these results suggested the need to characterise high density, high viscosity crude oils as multiple pseudo-component fluids.

4.2 Six Pseudo-components

Each of the eight oil samples listed in Table 1 were characterised as mixtures of six pseudo-components and corresponding entries were created in the PPDS Petroleum Fractions package. Figures 4 and 5 show the volume deviations with increasing temperature at average pressure and increasing pressure at average temperature respectively for these mixtures and clearly demonstrate that the two high density, high viscosity crude oils now exhibit similar behaviour to the other oils.

This is almost certainly due to the more realistic physical characterisation of the crude oils when using multiple pseudo-components within the PPDS Petroleum Fractions package. Furthermore, although all eight oils now display similar behaviour, there is a trend in the differences, in approximate API gravity order. In Figure 4 the deviations increase as the API gravity increases, i.e. as the crude oil becomes lighter, with a higher proportion of more volatile components. Whilst this may seem counter-intuitive, it is consistent with the derivation of the crude oil expansion coefficients; the samples used to derive the coefficients in the 1980 API Standard were allowed to stabilise in open containers before being measured [2]. As a result, the samples as tested would all have been more representative of low API gravity crude oils and hence the generic K_0 derived for use in equation 2 would reflect this. Modelling the crude oils in a more physically realistic manner, based on accurate assay data, will therefore show increasing deviations as the API gravity increases.

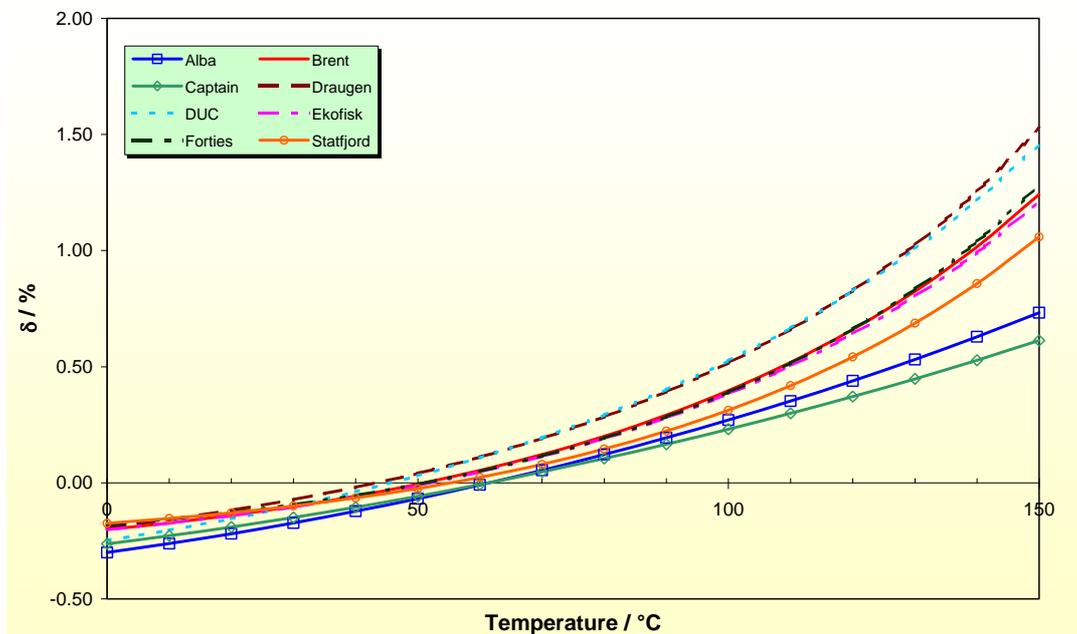


Fig 4. Volume deviation with increasing temperature at average pressure (six pseudo-components)

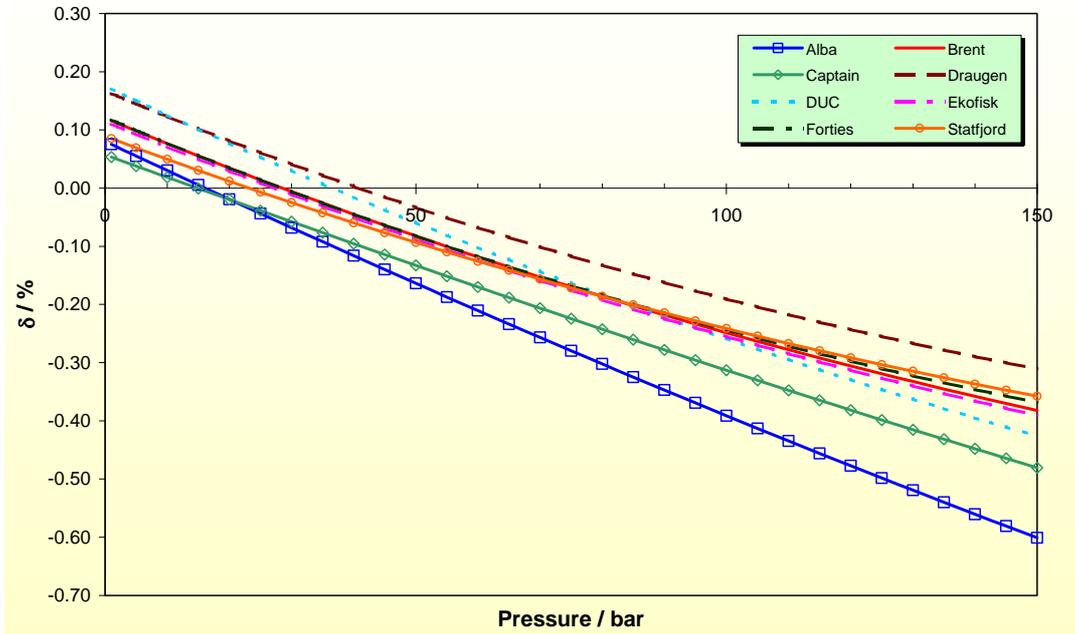


Fig 5. Volume deviation with increasing pressure at average temperature (six pseudo-components)

Using detailed assay data to characterise crude oils and a more physically realistic model such as the Tait-COSTALD method clearly provides a much better representation of the effects of temperature and pressure on fluid volume. The results shown in Figures 4 and 5 reinforce the need for a more rigorous approach to volume correction in the North Sea and elsewhere, to ensure that all crude oils are characterised effectively and to ensure economic losses are minimised.

5 DISCUSSION

From Sections 4.1 and 4.2 it is clear that there are significant differences between the volumes estimated by the expansion coefficient method and a more physically realistic model such as the Tait-COSTALD method. This highlights the possibility of economic consequences when using inaccurate expansion coefficients.

For all the major North Sea pipeline systems, allocation is done on a mass basis, by using densitometer data to convert volumetric flowrate from the fiscal meter to mass flowrate. Although this still involves the use of crude oil expansion coefficients, to account for the differences in temperature and pressure between the densitometer and the volumetric flowmeter, as shown in equation 9,

$$\rho_m(t_m, p_m) = \rho(t_d, p_d) \frac{C_{TL,m} C_{PL,m}}{C_{TL,d} C_{PL,d}} \quad (9)$$

errors will tend to cancel, particularly in well-designed systems where temperature differences will typically be less than 2K and pressure differences less than 0.5 bar. However, for some systems these differences can exceed 10K and 20 bar respectively. Clearly the actual differences will

depend on the temperature and pressure differences, and the effects of crude oil composition (including water and gas content) on the effective expansion coefficient.

If it is assumed that the PPDS Petroleum Fractions package implementation of the Tait-COSTALD method is accurately replicating experimental data when estimating crude oil densities, the economic consequences of using inaccurate expansion consequences can be estimated using these results.

Table 2 shows the potential economic effects of misallocation per annum at current production rates for the eight fields listed in Table 1. Table 2 has been constructed using an average price of \$100 per barrel for Brent crude throughout 2008 and assuming that the temperature differences between densitometers and flowmeters are 2K (with pressure differences of less than 0.5 bar). Although the deviations are small, for these eight fields alone, potential misallocations run to several million dollars per annum. Extrapolation of these figures to worldwide oil production suggests that there are significant problems with the current use of generic crude oil expansion coefficients to correct fluid volumes to other conditions, particularly is significant temperature or pressure differences are involved.

Table 2 – Potential mis-allocation

Field	Annual production barrels	Deviation %	Misallocation \$
Alba	18,940,000	-0.023	-429,150
Brent	3,046,000	-0.002	-5,990
Captain	16,257,000	-0.019	-313,410
Draugen	28,615,000	0.014	392,040
DUC	109,500,000	0.010	1,094,190
Ekofisk	99,317,000	-0.004	-379,680
Forties	22,568,000	-0.002	-47,050
Statfjord	463,000	-0.008	-3,830
Total misallocation			2,665,340

Within the UK, under the Licensing Consent system, oil and gas operators are expected to use “good oilfield practice” in all their measurement procedures. Whilst the use of the API / IP ASTM standard is common practice throughout the industry, it could be argued that it is no longer good practice. By extension, the requirements of the US Sarbanes-Oxley Act imply that the limitations of the current expansion coefficient method may need to be justified. Misallocation between companies clearly has significant implications, both for the companies concerned but also for the tax revenues for the governments involved.

6 CONCLUSIONS

From the results shown in Section 4, it can be seen that water content, temperature and pressure all appear to have a significant effect on the applicability of the current crude oil expansion coefficient method. The greatest errors are at high temperatures and pressures but there are also noteworthy errors at a wide range of temperatures, pressures and water contents. The economic consequences of these errors will be significant.

Whilst this paper has concentrated on the issue of allocation in shared off-shore pipeline systems, the problems caused by the use of generic crude oil expansion coefficients also affects on-shore pipelines and tanker loading systems

It is believed that physically realistic liquid density models, such as the Tait-COSTALD method, accurately replicate the density results which would be obtained experimentally but this would have to be verified before the limitations of the current expansion coefficient method could be confirmed. TUV NEL have a range of unique measurement facilities that would enable the densities of live crude oils to be accurately measured as functions of temperature and pressure. In conjunction with their expertise in thermophysical property modelling, they intend to launch a Joint Industry Project to address the issues highlighted by this initial study.

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