

Norske Sivilingeniørers Forening

NORTH SEA FLOW METERING WORKSHOP

Stavanger, 5 - 7 November 1985

Static measurement of LPG

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Lecturer: Mr. P. A. M. Jellfs
Moore, Barrett & Redwood Ltd

STATIC MEASUREMENT OF LPG

1. Introduction

This paper reviews the instrumentation and calculation procedures for loading and discharging ships carrying Light Hydrocarbon gas liquids.

2. Tank Design

Due to the pressure limitations for large storage containers the LPG is refrigerated in order to obtain a vapour pressure at near atmospheric conditions. Some LPG, however, is stored under pressure at ambient temperatures in horizontal cylinders or spheres both on shore and in ship's. Where gas liquids are refrigerated it is necessary to thermally insulate the tanks in order to reduce the boil-off of vapour from the liquid. These tanks will be equipped with both vapour input and output lines.

3. Safety Considerations

As the loading flow rate to the ship increases there can be a situation where the vaporisation rate is too slow to fill the space left in the tanks as the liquid level falls. This could lead to a tank's shell failure (implosion) and facilities for importing vapour from another tank are required. Alternatively the heat input through the insulation and from totally immersed pumps (for loading line circulation) can generate more vapour which must be exported.

4. Ship Loading Lines

The ship loading pipeline from the tank to the jetty is usually kept full of liquid and although insulated is usually continuously circulated to prevent formation of vapour. The circulation return line to the tank is also a further source of heat. In some cases there is a vapour return line from the ship into tank vapour space.

5. Static Measurement - (Based on Shore Tank)

The accurate measurement of quantities transferred from a vertical cylindrical storage tank is achieved by using a number of individual instruments which are described below:

5.1 Tank Level Measurement

Tank level gauges of the servo-operated displacer type are normally used for measuring the level of the liquid in the tanks. These gauges have no significant hysteresis and are able to determine the level to $\pm 2-3$ mm.

As the vapour space will have a varying temperature profile with very cold gas liquids such as ethane it is necessary to make corrections for the contraction of the tape wire and tank shell height. The tape correction is:

$$\Delta h = \theta_g(t_v) (H-h) - \theta_c(t_v) (H-h) - \theta_c(t_l)h \quad (1)$$

Where:

Δh = Correction to the gauge readout in mm.

H = Total height of the tank in mm at ambient temperature 15°C (from tank data).

h = Height of the liquid in the tank in mm (from gauge readout)

$\theta_g(t_v)$ = Change in length per unit length of the gauge wire or tape metal between 15°C and the average vapour temperature t_v .

$\theta_c(t_v)$ = Change in length per unit length of the tank metal between 15°C and the average vapour temperature t_v .

$\theta_c(t_l)$ = Change in length per unit length of the tank metal between 15°C and the liquid temperature t_l .

NOTE In theory h is the corrected height. However, in practice the gauge readout is used in the equation without introducing significant errors.

Usually two gauges are recommended so that there is a redundancy factor in the event of a failure of one device also the mean of two gauges will give an improved precision in determining the level.

5.2 Temperature Measurement

As the thermal expansion of LPG is of the order of 0.3% by volume for 1°C it is necessary to measure the vertical temperature profile in both the gas and liquid space. A three or four wire platinum resistance multi-sensor system is normally used. The sensors spaced every 1 to 2 metres apart are often housed in a flexible stainless sheathed cable. For digital systems the uncertainty of n sensors (immersed in the liquid) is $0.4/\sqrt{n}$ °C.

5.3 Pressure Measurement

The measurement of pressure in the vapour space is necessary in order to calculate the mass of the vapour above the liquid.

5.4 Vapour Measurement

Insertion type Vortex meters may be used for measuring the quantity of vapour entering into the tank from the compressor, adjacent tank or ships vapour return line during loading. Also vapour quantities discharged from the tank during loading at very low flow rates may be measured by Vortex meter.

5.5 Density Measurement

A density meter of the displacement or Archimedes type consisting of a float that is weighed by an electronic balance can be used for measuring liquid density in the tank. Alternatively a "vibration type" density meter can be installed in the shore tank which can be lowered through the liquid in order to obtain a density profile. Another method is to install a density meter in the loading line adjacent to the tank and to measure the average density of the liquid transferred to the ship. The device should be flow proportional.

A new single tube transducer has been recently developed which has a very low resonance frequency with a high Q value (a peaked resonance with large amplitude over a small range of frequency variation). This low resonance frequency is essential in order to minimise a systematic error which can occur when measuring densities of liquids such as LPG which have a low velocity of sound. (The low velocity of sound fails to disperse the pressure waves set up by the transducer vibration so that the density meter appears to see an apparently denser fluid.)

One method of eliminating this error is to calibrate the density meter on a liquid similar to the liquid in service. A special "Density Reference System" designed by NBS allows the density meter to be immersed in a refrigerated liquid in a vacuum insulated container. A silicon crystal suspended from the arm of an automatic balance allows the density to be measured to $\pm 0.02\%$.

5.6 Tank Calibration

The uncertainty of a vertical cylindrical shore tank is of the order $\pm 0.05\%$ expressed in terms of the cross sectional liquid area. It is necessary also to measure the total capacity of the tank in order to obtain the vapour volume.

Ships tanks which have been individually calibrated by physical measurement or water meter methods have an uncertainty of the order of ± 0.1 to $\pm 0.2\%$ (by volume).

Although accurate ship measurement is difficult due to the effects of liquid movement and trim/list, the overall uncertainty is usually better than the individual tank uncertainty due to the randomisation effect.

$$\text{Overall uncertainty} = \frac{\text{Tank uncertainty}}{\sqrt{\text{number of tanks}}}$$

5.7 Uncertainty of Static Measurement

The estimation of the uncertainty of the quantity loaded out of a tank is shown in the attached sheet. (Figure 1)

A number of assumptions are made which are detailed below:

- i. The uncertainty of the calibration of the shore tank has been estimated as $\pm 0.05\%$ (area) including the effect of the liquid head expansion of the shell plates.
- ii. The diameter of the tank does not vary by more than 1% from top to bottom.

- iii. Most of the bottom floor movement takes place between 2-3 metres and the empty condition.
- iv. The uncertainty of the vapour return quantity from the ship has not been included in the calculations.

5.8 Summary

The uncertainty improves as the Δh of the transfer quantity increases, which is the reason why long term contracts based on several transfers have until recently been acceptable. However, with the advent of the "spot cargo" market, single transfer quantities to small vessels (where the Δh is less than 10 m) can incur significant errors and is one of the reasons why meters are being installed (retrospectively in some cases).

6. Calculation Routines

- 6.1 To achieve a set of accurate calculation routines for certifying LPG cargoes which could be internationally standardised is now the main task of the Institute of Petroleum Liquid Gas Panel PM-F.

At present there are two main calculation routes as shown in the attached "Calculation Algorithm":- Fig 2

Mass (or weight) from volume and density at tank temperature

Mass (or weight) from volume and density at 15°C

These two calculation routes are partly covered by ISO Codes.

- 6.2 ISO DIS 6578 which covers density by calculation from composition is now being revised to accept the COSTALD density routine as an alternative to the Francis Routine. At present the code will only derive saturated densities (equilibrium conditions). A future revision would cover densities under pressure (which exceed the vapour pressure). Also there will be a need to include vapour density calculations for LPG under pressure by introducing a compressibility factor Z.
- 6.3 ISO 91/1 tables 53 and 54 below 650 k/m³ will be re-published. These tables are based on "orthobaric" densities which are determined at their saturated or vapour pressure. In order to calculate densities at pressures in excess of the vapour pressure it will be necessary to have equations and/or tables which will give vapour pressure and compressibility factors corresponding to values of density at 15°C respectively. The API have recently re-issued the original LPG compressibility tables in API Standard 1101 Table II and propose to supersede these with a new set in Chapter 11.2.2 covering the density range 350 to 636 k/m³. However these tables require the vapour pressure of the LPG mixture in order to derive the compressibility factor.

At present, however, there is no official correlation of LPG vapour pressures with density at 15°C. It can be shown that in theory there is no unique relationship between these two parameters due to the fact that it is possible to have mixtures of varying composition with the same density but different vapour pressures. However in practice the uncertainty due to this problem would probably be insignificant.

Provisional correlations, therefore, for both compressibility factors and vapour pressure with density at 15°C have been derived by my company.

6.4 Vapour Density Calculations

ISO DIS 6578 provides a density equation for vapour which uses the liquid molecular mass but does not include a correction for compressibility (Z). At refrigerated conditions for both commercial butane and propane the vapour pressure seldom exceeds 1 bar gauge. It is the intention therefore to recommend that where the pressures are less than say 2 bar gauge the Z factor will be taken as 1.0000.

For pressures above 2 bar gauge the Z factor will be calculated by using the Standing & Katz graph which is entered with Reduced Temperature T_R and Reduced Pressure P_R which can be calculated from composition ISO 91/1 has at present no vapour density routine but uses a "liquid equivalent of vapour". However as the latter makes no allowance for variations in composition or compressibility the I.P. are recommending the use of the same density routine as employed in ISO DIS 6578.

Graphs giving molecular weight, critical temperature and critical pressure for values of density at 15°C have been produced to enable the density calculation to be performed to the same order of accuracy as that derived from composition.

In theory the vapour composition and thus the molecular mass are not the same as the liquid for LPG mixtures. However where the main component in commercial butane and propane is greater than say 0.85% then the error incurred in calling the vapour molecular mass the same as the liquid is less than 0.1%.

6.5 Accuracy of the New Routines

It is evident that the original methods for calculating LPG under pressure employed density prediction equations which did not take into account the shrinkage - or negative volume of mixing which become significant at higher temperatures (i.e. ambient). ISO DP 4267/1.3 Calculation of Oil Quantities (Static). Also the mass of vapour should include a correction for compressibility which can be significant.

A table showing typical uncertainties illustrates the need for improved calculation techniques. (Fig 3)

6.6 Selection of Methods

The choice of the method is usually determined by the size and facilities of the installation.

Compositional analysis requires a fairly sophisticated laboratory with the analytical equipment and a competent technician.

In many situations the only apparatus for determining density is a pressure hydrometer which has a fairly low precision.

Even where there are facilities for obtaining the composition the density at 15°C is calculated using a Molar Volume Routine such as Francis or API Project 44 which does not allow for the negative volume of mixing.

7. Conclusion

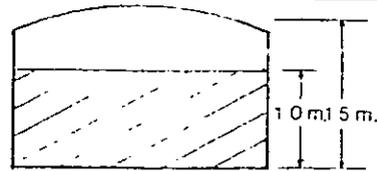
Improved instrumentation and standardised calculation procedures are now essential if voyage losses in LPG are to be accurately monitored between loading and discharge ports.

Note. An alternative routine is the Redlich Soave Vapour equation.

The mass of LPG at ambient temperatures and under pressure will be under-estimated when the calculation routines do not take into account the negative volume of mixing in the liquid phase and the compressibility in the vapour phase. Typically calculations utilizing Table 54 are under-estimated by 0.3% and utilizing composition by 0.5%.

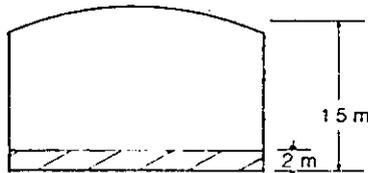
Fig 1 ESTIMATED UNCERTAINTY OF A PROPANE LIQUID TRANSFER

Assumptions
 1mm depth = 1m³
 Density at 15° C = 507.4kg/m³
 at -43° C = 582.0kg/m³



<u>LIQUID</u>			
Level	± 2mm	2 x 1 mm	= 2.00m ³
Temp. ° C	± 0.15° C	Plate shrinkage = 0.000022 x 0.15 x 10 000	= 0.03m ³
Tank Calibration	± 0.05%	0.0005 x 10 000	= 5.00m ³
		Vol. Uncertainty = $\sqrt{(2.00^2 + 0.03^2 + 5.00^2)}$	= 5.39m ³
		Mass = 5.39 x 582 / 1 000	= 3.14 tonnes
Density	± 0.22%	0.0022 x 582 x 10 000 / 1 000	= 12.80 tonnes
		Liq. Mass Uncertainty = $\sqrt{(3.14^2 + 12.80^2)}$	= 13.18 tonnes
<u>VAPOUR</u>			
Tank Calibration + Contraction.		Ignore	
Density	{ Temp. ± 1° C Composition ± 0.12% Mol. Wt. ± 0.10% Press. 5mb ± 0.30%	± 1.0%	2.400 x 0.01 x 5 000 / 1 000 = 0.12 tonnes
<u>LIQUID & VAPOUR</u>		Uncertainty (Mass) = $\sqrt{(13.18^2 + 0.12^2)}$	= 13.18 tonnes
<u>TOTAL MASS</u>		- 582 x 10 000 / 1 000 + 2.4 x 5 000 / 1 000	= 5 832 tonnes

Assumptions
 1mm depth = 1m³
 Density at 15° C = 507.4kg/m³
 at -43° C = 582.0kg/m³



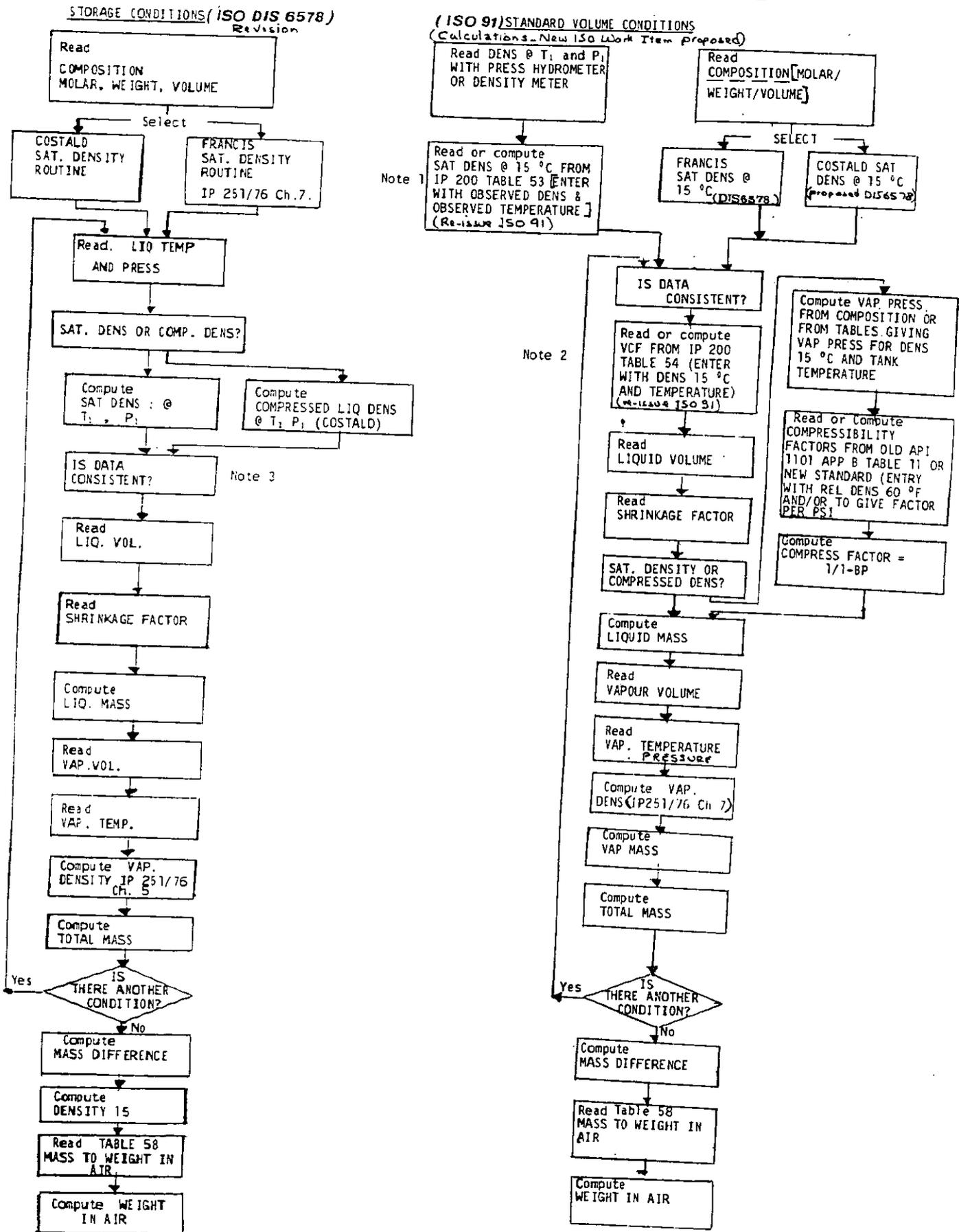
<u>LIQUID</u>			
Level	± 2mm	Same	= 2.00m ³
Temp. ° C	± 0.15° C	Plate Shrinkage = 0.000022 x 0.15 x 2 000	= 0.01m ³
Tank Calibration	± 0.05%	0.0005 x 2 000	= 1.00m ³
		Vol. Uncertainty = $\sqrt{(2.0^2 + 1.0^2)}$	= 2.24m ³
		Mass = 2.24 x 582 / 1 000	= 1.30 tonnes
Density	± 0.22%	0.0022 x 582 x 2 000 / 1 000	= 2.56 tonnes
		Liq. Mass Uncertainty = $\sqrt{(1.30^2 + 2.56^2)}$	= 2.87 tonnes
<u>VAPOUR</u>			
Tank Calibration + Contraction		Ignore	
Density	{ Temp. ± 1° C Composition ± 0.12% Mol. Wt. ± 0.10% Press. 5mb ± 0.30%	± 1.0%	
<u>LIQUID & VAPOUR</u>		Uncertainty (mass) = $\sqrt{(2.87^2 + 0.31^2)}$	= 2.89 tonnes
<u>TOTAL MASS</u>		= 582 x 2 000 / 1 000 + 2.4 x 13 000 / 1 000	= 1 195 tonnes

Uncertainty Before 13.18 x 100/5832 = 0.23%

Uncertainty After 2.89 x 100/1195 = 0.24%

SUMMARY Uncertainty of Mass Transfer = $\frac{100 \sqrt{(13.18^2 + 2.89^2)}}{(5832 - 1195)} = \pm 0.29\%$

Fig 2 LIQUID PETROLEUM GAS CALCULATION ROUTINES



NOTES :
 Note 1 Equation accepted by PM-C-2
 Note 2 " " " " PM-C-2
 Note 3 Equivalence between meas. density, meas. temp, and calculate pressure

Fig 3

Instrumentation and Calculation Uncertainties

Density (By Composition and Calculation).

$$\left. \begin{array}{l} \text{Composition } 0.08\% \\ \text{Costald } 0.20\% \end{array} \right\} \text{Uncertainty} = \sqrt{(0.08^2 + 0.20^2)} \\ = \pm 0.22\%$$

Francis [additional bias due to shrinkage] at $-43^{\circ}\text{C} + 0.20\%$
at $15^{\circ}\text{C} = +0.40\%$

Density (by direct reading)

	Pressure Hydrometer		
VCF	$\sqrt{(0.34^2 + 0.09^2)}$	=	$\pm 0.35\%$
Density Meter	not known	=	$\pm 0.20\% ?$

References

[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.