

MEASUREMENT OF GAS AND LIQUIDS

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MEASUREMENT OF LPG

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INTRODUCTION

The static measurement of Light Hydrocarbon gases is usually associated with the loading of ships where the liquid is normally refrigerated in order to maintain the vapour pressure at near atmospheric conditions.

Tank Design

This constraint is normally imposed because it is not economically feasible to build either ship or shore tanks as pressure vessels.

In the refrigerated condition the gas liquids will normally be stored in vertical cylindrical storage tanks which are thermally insulated to reduce the boil off of vapour from the liquid. The tanks will be equipped with both vapour input and output lines. (See Figure 1)

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 tank as the liquid level falls. This could lead to a tank 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.

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 most cases there is a vapour return line from the ship into tank vapour space.

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:

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 . Values for various temperatures are given in Table 1.

$\theta_c(t_v)$ = Change in length per unit length of the tank metal between 15°C and the average vapour temperature t_v . Values for various temperatures are given in Table 1.

$\theta_c(t_l)$ = 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.

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.

Pressure Measurement

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

Vapour Measurement

Insertion type Vortex meters are 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 are measured by Vortex meter.

Density Measurement

A densitometer 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" densitometer can be installed in the loading line near to the tank. 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 densitometer appears to see an apparently denser fluid.)

One method of eliminating this error is to calibrate the densitometer on a liquid similar to the liquid in service. A special "Density Reference System" designed by NBS allows the densitometer 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 $\pm 0.02\%$. (See Figure 2)

Uncertainty of Static Measurement

The estimation of the uncertainty of the quantity loaded out of a tank is shown in the attached sheet. (Figure 3)

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.

CONCLUSIONS

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.

If an insertion type vortex meter is used for measuring the vapour return its contribution will be negligible to the overall uncertainty.

In terms of mass a Vortex meter with an uncertainty of $\pm 5\%$ would only contribute $5/273 = \pm 0.02\%$ to the overall uncertainty (Gas/Liquid ratio for C3 = 273).

The uncertainty of the quantity loaded is not dependent on the flow rate.

The uncertainty given in the examples are:

Δh = 2M^φ

23m = 33817 tonne x 0.0023 = 78 tonne

21m = $\frac{30892}{2925}$ tonne x 0.0023 = 71 tonne

Uncertainty = $\frac{100 (78^2 + 71^2)^{1/2}}{2925}$ = ± 3.6%

Δh = 5M^φ

23m = 33817 tonne x 0.0023 = 78 tonne

18m = $\frac{26506}{7311}$ tonne x 0.0023 = 61 tonne

Uncertainty = $\frac{100 (78^2 + 61^2)^{1/2}}{7311}$ = ± 1.4%

Δh = 20M

23m = 33817 tonne x 0.0023 = 78 tonne

3m = $\frac{4569}{29248}$ tonne x 0.0028 = 13 tonne

Uncertainty = $\frac{100 (78^2 + 13^2)}{29248}$ = ± 0.27%

NOTE ^φ

The uncertainty of transfers representing small "Δh" are grossly overestimated due to the assumption that the uncertainty of the initial and final quantities in the tank are unrelated. Whereas in practice the conditions of measurement would be similar (i.e. little change in temperature for small Δh etc.) A figure of 2.0% rather than 3.6% for a Δh = 2m would probably be more realistic.

DYNAMIC MEASUREMENT (Based on Meters)

There are a number of metering systems which are currently employed to measure refrigerated light hydrocarbon liquids. These systems are described below:

1 Turbine Meters and Pipe Prover

Experience has shown that turbine meters used for measuring LPG usually have a very poor linearity - variation of K factor with flow rate.

Typically a 6" meter used on crude oil will have a change in K factor of $\pm 0.2\%$ over a 6:1 flow range whereas on LPG the same meter would have ± 0.3 to 1.0% over 4:1 flow range. Under these circumstances it is necessary to use a micro-computer based linearisation technique where K factor versus frequency (pulse rate) curve is stored in the computer and the appropriate K factor applied to the integration of the throughput as the frequency changes.

In order to minimise the error incurred by large variation in flow rate i.e. very low flow rates when topping up ships tanks, several meters in parallel can be used. The number of meters employed for any given flow rate will be such that all the meters are operating within the top end of their linear range.

Special consideration however, has to be given to the problems of two phase flow as vapour can create very high velocities which may over-spin the rotor and damage the blades and bearings when the lubricity of the product is poor.

Several meters are now available which are specially designed to overcome this problem.

One method of reducing the bearing friction is to produce a rotor which rotates on a tungsten carbide spindle which in turn rotates in two tungsten carbide end journals. By this arrangement the rotational velocity and hence friction is reduced. Also the rotor can be accurately machined out of solid bar so that it will stand up to sonic velocities without shattering.

Pipe provers employing pistons with suitable peripheral seals can be operated at low temperatures. These provers, which are fitted with proximity detector switches at distances of 10 metres apart give satisfactory repeatability i.e. 5 results within a spread of 0.05% .

Comparisons between metered quantities and ships quantities (the ships were specially calibrated to $\pm 0.2\%$ by volume) are shown in figure 4.

2 Ultrasonic Meters

a. Principles

High frequency soundwaves in the ultrasonic range are beamed across a pipe at an angle, usually 45° . The velocity of the ultrasonic beam is increased or decreased by the fluid velocity depending whether the beam is with or against the flow. (See Figure 5).

The transit time of the pulses in the flight path is expressed in two equations involving the path length between the two transducers (L_p), sonic velocity in the fluid (C) and the component of fluid velocity in the direction of the path (V_p).

Transit Time

$$T_{1-2} = \frac{l_p}{C+V_p} \quad \text{in the direction of the fluid flow}$$

$$T_{2-1} = \frac{l_p}{C-V_p} \quad \text{against the fluid flow}$$

As the path length and transit time can be measured precisely and the sonic velocity in the fluid can be assumed to be constant the equations can be simplified as follows:

$$\Delta T = T_{2-1} - T_{1-2}$$

$$V = \frac{V_p}{\cos\theta} = \frac{l_p \Delta T}{2T_{1-2}(T_{1-2} + \Delta T)\cos\theta}$$

b. Applications

The main advantages of the ultrasonic meter for the measurement of LPG are:

- no obstruction in the pipe and therefore no pressure drop through the meter.
- no moving parts that can be damaged during a "blown-down".
- high frequency beams can be used with LPG (see accuracy)

The main disadvantage is the problem of gas break-out when the liquid is near its' bubble point when being measured, as vapour can dissipate the signal especially if it collects in the small chambers where the transducers are located. (This can be overcome by having the transducers fitted horizontally). The upstream straight lengths with or without a flow straightener to eliminate swirl and pulsation are the same as those required for turbine meters.

c. Repeatability

The repeatability of the meter is mainly dependant on the resolution or discrimination of the measurement of the transit time and the number of individual measurements. A typical transit time for a 24" diameter meter would be 200 nano seconds at 490 m³/h with a resolution of 1 nano second. This represents 1/200 = 0.05% or say 2 m³/h.

However, as the number (n) of individual measurements is increased so the resolution is improved by $\frac{1}{\sqrt{n}}$.

Also there is random variation in a single measurement of flow due to the turbulence in the fluid flow. The maximum variation has been found to be of the order of 20%. This error can similarly be reduced by averaging a large number of individual measurements. However, there is a limit to the length of the averaging period or update time.

In the case of the 24" meter the update time is 4 seconds involving 200 individual readings (100 per path in a two path meter).

The uncertainty due to this random variation in ΔT is:

$$(E_R)_{95} = \frac{W}{\sqrt{n}} = \frac{20}{\sqrt{200}} = \pm 1.4\%$$

Combining the uncertainty due to the discrimination of the transit time with the uncertainty due to the random variation for a single measurement, in terms of the total update time the equation for the overall random uncertainty % is:

$$(E_R)_{95} = \frac{1}{\sqrt{t/n}} \left\{ \left(\frac{100 \cdot tr}{\Delta T} \right)^2 + 1.4^2 \right\}^{1/2}$$

Where: t = throughput time

n = update time

ΔT = transit time

tr = resolution of transit time (smallest discrimination)

d. Accuracy

The accuracy of the meter is dependant on the frequency of the emitted pulses. As the leading edge of the first emitted pulse is used to trigger the electronic timing processor it must be very sharp or nearly vertical. However, certain absorptive fluids such as crude oil or viscous liquids are difficult to penetrate so that the frequency has to be reduced with the result that the location of the leading edge is less clearly defined. This factor reduces the accuracy of determining ΔT .

Also the number of ΔT 's that can be measured and integrated to derive volume throughput is a function of processing capability. It is often a trade-off between micro-computer cost and sophistication.

The shorter the update period and the larger the number of individual ΔT measurements the less sensitive will be the meter to changes in flow rate.

Another factor is the accurate determination of the average velocity over the pipe cross-section.

The velocity distribution in a pipe is a function of the Reynolds Number (Re) and of the pipe configuration in the vicinity of the metering section. In the acoustic path crossing the pipe there will be a variation in the velocity of each local element of flow. Although it can be assumed therefore, that the average velocity over the acoustic path is measured it will be necessary to compute the overall velocity profile across the whole pipe cross-section. As the flow varies so the profile will vary as well.

The systematic error for an ultrasonic meter is a function of the accuracy of the determination of the path length l_p , $\cos\theta$, number of acoustic paths and the formula for integrating the flow profile and the volume flow rate. In practice the errors due to the determination of the physical dimensions of the large diameter ultrasonic meters would be less than 0.01%.

However, there is evidence of a systematic error of the order of $\pm 0.4\%$ from tests carried out on a 24" two path meter.

e. Linearity

It can be seen that the linearity of the ultrasonic meter is dependant on the number of paths, the flow profile and the repeatability. (See Figure 4).

In practice there is evidence that at Reynolds numbers below 500,000 the curve of K factor versus Reynolds number becomes non-linear. This is probably due to boundary layer effects. Typical "linearities" for 12" and 24" 2-path meters are given below:

Meter Size	Viscosity cSt at 15°C	Flow Range m ³ /h	Maximum Variation % in K factor
8"	0.7	400-1200	$\pm 0.05\%$
	0.7	100-1200	$\pm 0.08\%$
24"	0.9	1000-5500	$\pm 0.15\%$
	0.9	550-5500	$\pm 0.80\%$
24"	0.9	1600-5500	$\pm 0.24\%$
	0.9	550-5500	$\pm 0.83\%$

f. Calibration (On Site)

There is some evidence that it is possible to estimate K factor of an ultrasonic meter without carrying out a physical calibration. However, from tests where meters have been compared to gravimetric systems (weigh tanks in laboratories) or turbine meters used as a transfer standard against pipe provers the estimated uncertainty is seldom better than $\pm 0.4\%$

Also there is evidence that the repeatability of the ultrasonic meter improves with time (throughput period) (see figure 7). Under these circumstances, therefore, it is possible to use a shore tank as described in the first section of this paper to calibrate the ultrasonic meter on site.

The tank uncertainty is dependant on the Δh and as this is a function of loading time, a continuous comparison can be made with meter during periods of steady flow rates.

g. Data Processing and Integration

A micro-computer can be employed to continually integrate the mass of product in the tank by reading the level gauges, temperature sensors, vortex (vapour) meters and densitometer readings. The mass of both liquid and vapour in the tank is constantly up-dated and compared with the up-dated ultrasonic meter. (See Figure 8).

3 VORTEX METERS

a. Principles

A solid body usually known as a bluff body when placed in a stream of flowing liquid at Reynolds number of approximately 30,000 gives rise to a number of vortices. This phenomenon known as "vortex shedding" has the characteristic that the vortices are shed from alternate sides of the bluff body. As the volume flowing through a meter can be related directly to the frequency of the vortices it is possible to measure the volume throughput by counting the total vortices. The ideal shape of the bluff body is a triangle which induces strong vortices. (See Figure 9). A "shuttle" senses the small reduction in pressure created by each vortex.

The ideal relationship is given by the equation:

$$S = \frac{f \cdot d}{v}$$

where S is known as the Strouhal number (constant)
d is the bluff body diameter
f is the frequency
v is the velocity

b. Application

Due to the bluff body there is a pressure drop through the meter which can cause cavitation which in turn can effect the accuracy of the meter.

The API formula for defining the back pressure required to eliminate cavitation is similar to that required for turbine meters in LPG service.

A number of straight lengths in the upstream pipe with or without a flow straightening device is similar to that required for the turbine meter. The advantage of no moving parts means that the meter can be blown-down without damage.

c. Repeatability

The frequency becomes lower for a given fluid velocity as the meter diameter increases as shown below:

<u>Meter Size (Diameter)</u>	<u>Pulses/m³</u>
2"	9600
12"	544
24"	11

Tests have shown that there is a maximum variation of 20% between the largest and smallest interval between consecutive vortices, which is mainly due to the random velocity effects in the flowing liquid. The repeatability therefore, will be mainly a function of the throughput period.

d. Accuracy

In theory there should be no effect on the accuracy of the meter throughput due to variation in viscosity or density of the measured liquid. However, in practice there can be significant differences in the K factor between water and LPG, especially in the smaller size of meters less than 2".

e. Linearity

The variation of K factor with flow rate decreases as the meter diameter increases as shown below:

Meter Size (Diameter)	Flow Range (m /h)	Maximum Variation % in K Factor
2"	5- 40	0.8 to 1.1%
12"	80- 900	0.3% (See figure 10)
24"	500-1600 [∅]	0.1%

[∅] Unable to test meters over wider flow range.

f. Calibration

There is no direct evidence that it is possible to estimate the K factor of a Vortex meter without carrying out a physical calibration.

However, it is necessary with the large diameter meters (>12") to compare them with a transfer standard such as a turbine meter in order to achieve the required repeatability.

A shore tank as described in the section on Ultrasonic meters can also be used to determine a K factor over a long loading period on site.

TABLE 1

THERMAL CONTRACTION CORRECTION FACTORS FOR

LEVEL GAUGE AND TANK METALS

(See Note at end of Table)

This table gives values of $\theta(t)$ or $\theta(T) = \frac{\Delta L}{L_{15}}$ for various temperatures $t^{\circ}\text{C}$ or $T^{\circ}\text{C}$. For derivation refer to A.1.2. When used in the formulae given in 3.3 the general symbols are used in subscript form as follows to indicate the application

$$\theta_g(t_v), \theta_c(t_v), \theta_c(t_\ell), \theta_g(t_\ell), \theta_c(t_a), \theta_c(T)$$

Temperature t or T $^{\circ}\text{C}$	$\theta(t)$ or $\theta(T)$ values			
	Al Alloy	36% Ni/Iron	Ferrous Alloy AISI 301/306	Stainless Steel
+ 15	Zero	Zero	Zero	Zero
0	0.00035	0.00004	0.00023	0.00015
- 10	0.00058	0.00006	0.00039	0.00024
- 20	0.00080	0.00009	0.00053	0.00034
- 30	0.00102	0.00012	0.00067	0.00042
- 40	0.00124	0.00014	0.00083	0.00051
- 50	0.00145	0.00017	0.00097	0.00059
- 60	0.00166	0.00019	0.00111	0.00067
- 70	0.00186	0.00022	0.00125	0.00075
- 80	0.00205	0.00024	0.00138	0.00083
- 90	0.00234	0.00027	0.00148	0.00090
- 100	0.00263	0.00030	0.00163	0.00098
- 101	0.00264	0.00030	0.00164	0.00099
- 102	0.00265	0.00031	0.00165	0.00100
- 103	0.00266	0.00031	0.00166	0.00101
- 104	0.00267	0.00031	0.00167	0.00102
- 105	0.00268	0.00032	0.00168	0.00102
- 106	0.00269	0.00032	0.00170	0.00103
- 107	0.00270	0.00032	0.00171	0.00103
- 108	0.00271	0.00032	0.00173	0.00104
- 109	0.00272	0.00033	0.00174	0.00105
- 110	0.00273	0.00033	0.00176	0.00106
- 111	0.00274	0.00033	0.00177	0.00107
- 112	0.00275	0.00033	0.00179	0.00108
- 113	0.00276	0.00033	0.00180	0.00108
- 114	0.00277	0.00033	0.00181	0.00109
- 115	0.00277	0.00034	0.00183	0.00110
- 116	0.00278	0.00034	0.00184	0.00111
- 117	0.00279	0.00034	0.00185	0.00112
- 118	0.00280	0.00034	0.00187	0.00112

Temperature t or T °C	$\theta(t)$ or $\theta(T)$ values			
	Al Alloy	36% Ni/Iron	Ferrous Alloy AISI 301/306	Stainless Steel
119	0.00281	0.00034	0.00188	0.00113
120	0.00281	0.00035	0.00189	0.00113
121	0.00283	0.00035	0.00190	0.00114
122	0.00284	0.00035	0.00191	0.00115
123	0.00286	0.00035	0.00193	0.00115
124	0.00287	0.00035	0.00194	0.00116
125	0.00289	0.00036	0.00195	0.00116
126	0.00291	0.00036	0.00197	0.00117
127	0.00292	0.00036	0.00198	0.00118
128	0.00294	0.00036	0.00199	0.00119
129	0.00295	0.00036	0.00201	0.00120
130	0.00297	0.00037	0.00201	0.00121
131	0.00299	0.00037	0.00202	0.00122
132	0.00300	0.00037	0.00203	0.00122
133	0.00302	0.00037	0.00205	0.00123
134	0.00304	0.00037	0.00206	0.00124
135	0.00306	0.00037	0.00208	0.00125
136	0.00307	0.00038	0.00209	0.00125
137	0.00309	0.00038	0.00210	0.00126
138	0.00311	0.00038	0.00212	0.00127
139	0.00312	0.00038	0.00213	0.00127
140	0.00314	0.00039	0.00214	0.00128
141	0.00316	0.00039	0.00215	0.00128
142	0.00318	0.00039	0.00216	0.00129
143	0.00319	0.00039	0.00218	0.00130
144	0.00320	0.00039	0.00219	0.00130
145	0.00322	0.00040	0.00220	0.00131
146	0.00324	0.00040	0.00222	0.00132
147	0.00325	0.00040	0.00223	0.00133
148	0.00327	0.00040	0.00224	0.00134
149	0.00328	0.00040	0.00225	0.00135
150	0.00330	0.00041	0.00227	0.00136
151	0.00332	0.00041	0.00228	0.00137
152	0.00333	0.00041	0.00229	0.00137
153	0.00335	0.00041	0.00230	0.00138
154	0.00336	0.00041	0.00231	0.00138
155	0.00338	0.00042	0.00234	0.00139
156	0.00339	0.00042	0.00235	0.00140
157	0.00341	0.00042	0.00236	0.00141
158	0.00342	0.00042	0.00237	0.00142
159	0.00344	0.00042	0.00238	0.00142
160	0.00345	0.00043	0.00239	0.00143

Temperature t or T °C	$\theta(t)$ or $\theta(T)$ values			
	Al Alloy	36% Ni/Iron	Ferrous Alloy AISI 301/306	Stainless Steel
- 161	0.00346	0.00043	0.00240	0.00143
- 162	0.00346	0.00043	0.00241	0.00144
- 163	0.00347	0.00043	0.00242	0.00144
- 164	0.00348	0.00043	0.00243	0.00145
- 165	0.00349	0.00044	0.00245	0.00146
- 166	0.00349	0.00044	0.00246	0.00146
- 167	0.00350	0.00044	0.00248	0.00147
- 168	0.00351	0.00044	0.00249	0.00148
- 169	0.00352	0.00044	0.00250	0.00149
- 170	0.00352	0.00045	0.00251	0.00150
- 180	0.00358	0.00047	0.00265	
- 190	0.00379	0.00048	0.00277	
- 200	0.00385	0.00050	0.00287	

NOTE The values given in this table are typical for alloys and stainless steels and it is recommended that in order to achieve the highest level of accuracy the coefficient should be determined on a sample of the material used.

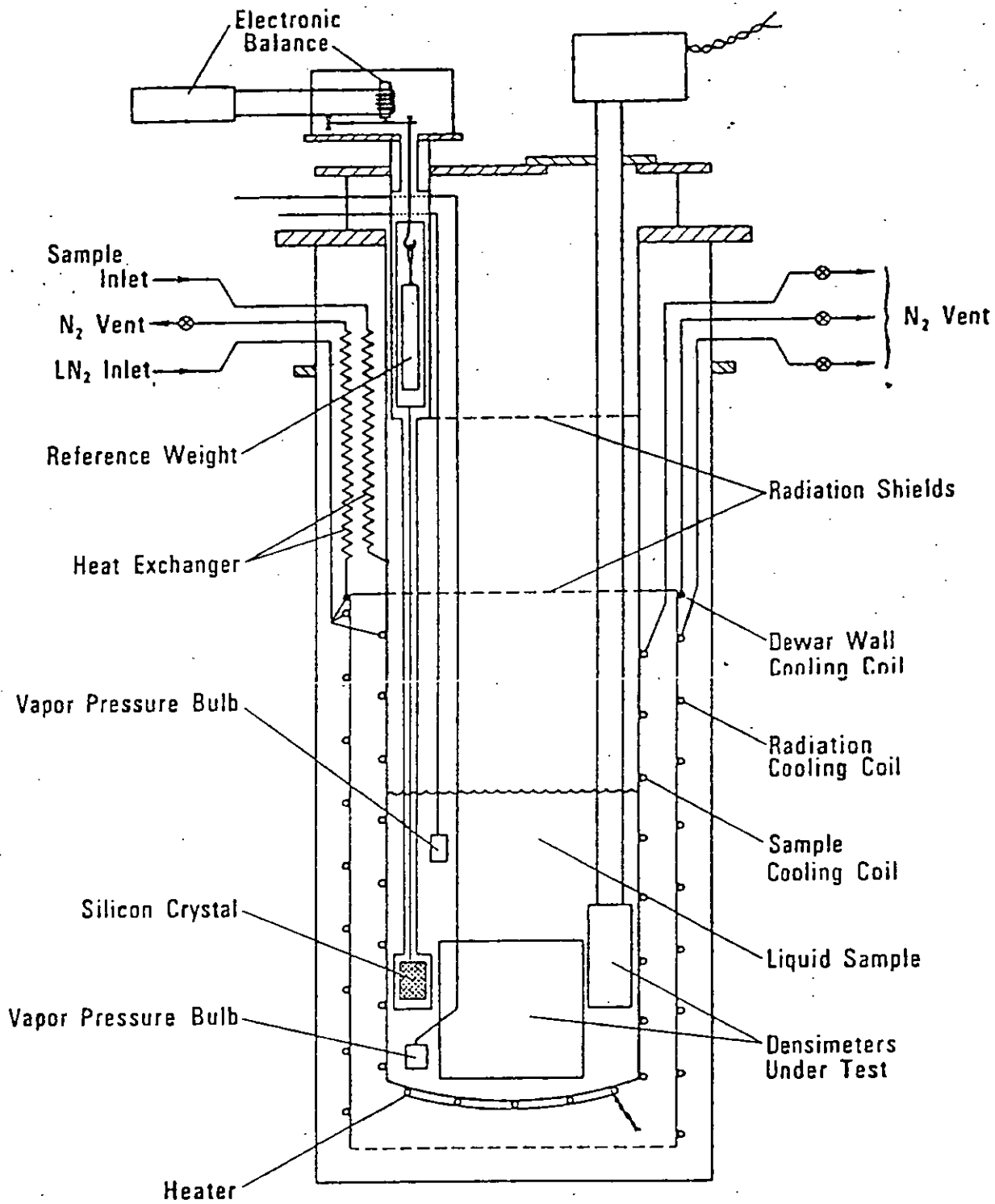
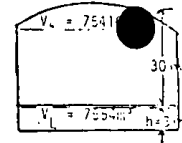
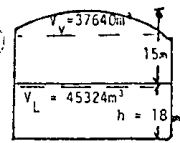
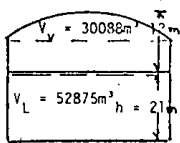
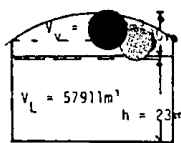


Fig 2 Density Reference System

Estimation of Uncertainty of a Propane Liquid Transfer from a Storage Tank

FIG 3



Component Uncertainties	h = 23M	h = 21M	h = 18M	h = 3M
LIQUID				
Level (a) contraction of stillwell for $\pm 5^\circ\text{C}$ = $5 \times 0.000014 \times 33000 = \pm 2.3\text{mm}$ (1)				
(b) = repeatability of $\pm 3\text{mm} = \pm 2.1\text{mm}$ $\sqrt{2}$ = $(2.3^2 + 2.1^2)^{1/2} = \pm 3.1\text{mm}$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$
Volume - Calibration table = $\pm 0.05\%$	$0.0005 \times 2518 \times 23$ = $\pm 29\text{m}^3$	$0.0005 \times 2518 \times 21$ = $\pm 26\text{m}^3$	$0.0005 \times 2518 \times 18$ = $\pm 23\text{m}^3$	$0.0005 \times 2518 \times 3$ = $\pm 4\text{m}^3$
Shell Shrinkage = $0.000028 \times 5 \times 100 = 0.01\%$ (2)	$0.0001 \times 2518 \times 23$ = $\pm 6\text{m}^3$	$0.0001 \times 2518 \times 21$ = $\pm 5\text{m}^3$	$0.0001 \times 2518 \times 18$ = $\pm 5\text{m}^3$	$0.0001 \times 2518 \times 3$ = $\pm 0.8\text{m}^3$
Volumetric Uncertainty	volume = $8 + (29^2 + 6^2)^{1/2}$ = $\pm 38\text{m}^3$ mass (4) = 38×583 = 22.2 Tonnes (5)	volume = $8 + (26^2 + 5^2)^{1/2}$ = $\pm 34\text{m}^3$ mass (4) = $34 \times 583 = 19.8 \text{ Tonnes}$ (5)	volume = $8 + (23^2 + 5^2)^{1/2}$ = $\pm 32\text{m}^3$ mass (4) = $32 \times 583 = 18.7 \text{ Tonnes}$ (5)	volume = $8 + (4^2 + 0.8^2)^{1/2}$ = $\pm 12\text{m}^3$ mass (4) = $12 \times 583 = 7.0 \text{ Tonnes}$ (5)
Density - Calibration = 0.2% (6) - Correction = $\pm 0.1\%$ (6) $(0.2 + 0.1)^2 = \pm 0.22\%$	mass (4) = $\frac{2518 \times 23 \times 583 \times 0.0022}{1000}$ = 74.3 Tonnes	mass (4) = $\frac{2518 \times 21 \times 583 \times 0.0022}{1000}$ = 67.8 Tonnes	mass (4) = $\frac{2518 \times 18 \times 583 \times 0.0022/1000}{1000}$ = 58.1 Tonnes	mass (4) = $\frac{2518 \times 3 \times 583 \times 0.0022}{1000}$ = 9.7 Tonnes
Overall Liquid Uncertainty	mass = $(22.2^2 + 74.3^2)^{1/2}$ = $\pm 77.5 \text{ Tonnes}$	mass = $(19.8^2 + 67.8^2)^{1/2}$ = $\pm 70.6 \text{ Tonnes}$	mass = $(18.7^2 + 58.1^2)^{1/2}$ = $\pm 61.0 \text{ Tonnes}$	mass = $(7.0^2 + 9.7^2)^{1/2}$ = $\pm 12.0 \text{ Tonnes}$
VAPOUR				
Volume - calibration table - dome = 0.5% - tank = 0.05%	0.005×22711 } $\pm 114 \text{ m}^3$ 0.0005×2342 }	0.005×22711 } $\pm 117 \text{ m}^3$ 0.0005×7377 }	0.005×22711 } $\pm 121 \text{ m}^3$ 0.0005×14929 }	0.005×22711 } $\pm 140 \text{ m}^3$ 0.0005×52699 }
Shell Shrinkage = $0.000042 \times 5 \times 100 = 0.02\%$	volume = 0.0002×25053 = $\pm 5\text{m}^3$ = $(114^2 + 5^2)^{1/2}$ = $\pm 114\text{m}^3$	volume = 0.0002×30088 = $\pm 6\text{m}^3$ = $(117^2 + 6^2)^{1/2}$ = $\pm 117\text{m}^3$	volume = 0.0002×37640 = $\pm 8\text{m}^3$ = $(121^2 + 8^2)^{1/2}$ = $\pm 124 \text{ m}^3$	volume = 0.0002×75410 = $\pm 15\text{m}^3$ = $(140^2 + 15^2)^{1/2}$ = $\pm 141\text{m}^3$
Density (7) - Temp = $\pm 5^\circ\text{C} = 2.2\%$ Press = $\pm 5\text{mb} = 0.5\%$ } $\pm 2.8\%$ (8) Mole Comp = 1.6% }	mass = $114 \times 2.183^{(8)}$ = $\pm 0.25 \text{ Tonnes}$	mass = 117×2.183 = $\pm 0.26 \text{ Tonnes}$	mass = 121×2.183 = 0.26 Tonnes	mass = 141×2.183 = $\pm 0.31 \text{ Tonnes}$
Overall Vapour Uncertainty	mass = $\frac{25053 \times 2.183 \times 0.028}{1000}$ = $\pm 1.5 \text{ Tonnes}$ = $(0.25^2 + 1.5^2)^{1/2}$ = 1.5 Tonnes	mass = $\frac{30088 \times 2.183 \times 0.028}{1000}$ = $\pm 1.8 \text{ Tonnes}$ = $(0.26^2 + 1.8^2)^{1/2}$ = $\pm 1.8 \text{ Tonnes}$	mass = $\frac{37640 \times 2.183 \times 0.028/1000}{1000}$ = $\pm 2.3 \text{ Tonnes}$ = $(0.26^2 + 2.3^2)^{1/2}$ = $\pm 2.3 \text{ Tonnes}$	mass = $\frac{75410 \times 2.183 \times 0.028/1000}{1000}$ = 4.6 Tonnes = $(0.31^2 + 4.6^2)^{1/2}$ = 4.6 Tonnes
TOTAL	Liquid = $57911 \times 583 = 33762 \text{ Tonnes}$ Vapour = $25053 \times 2.183 = 55 \text{ Tonnes}$ $\frac{33817}{33817}$ $(77.5^2 + 1.5^2)^{1/2} = 77.5 \text{ Tonnes} = 0.23\%$	Liquid = $52875 \times 583 = 30826 \text{ Tonnes}$ Vapour = $30088 \times 2.183 = 66 \text{ Tonnes}$ $\frac{30892}{30892}$ $(70.6^2 + 1.8)^{1/2} = 70.6 \text{ Tonnes} = 0.23\%$	Liquid = $45324 \times 583 = 26424 \text{ Tonnes}$ Vapour = $37640 \times 2.183 = 82 \text{ Tonnes}$ $\frac{26506}{26506}$ $(61.0^2 + 2.3^2)^{1/2} = 61.0 \text{ Tonnes} = 0.23\%$	Liquid = $75410 \times 583 = 4404 \text{ Tonnes}$ Vapour = $75410 \times 2.183 = 165 \text{ Tonnes}$ $\frac{4569}{4569}$ $(12^2 + 4.6^2)^{1/2} = 12.9 \text{ Tonnes} = 0.23\%$

(1) Tank dia = 56.62mm ($\text{m}^3/\text{m} = 2518$)
 $\text{m}^3/\text{mm} = 2.518$
(2) Height of Stillwell = 33000 mm

(3) Av. temp error = $\pm 5^\circ\text{C}$
(4) Density of C3 = 583kg/m³

(5) Tonnes - kg/1000
(6) Uncertainty of densitometer = $\pm 0.2\%$

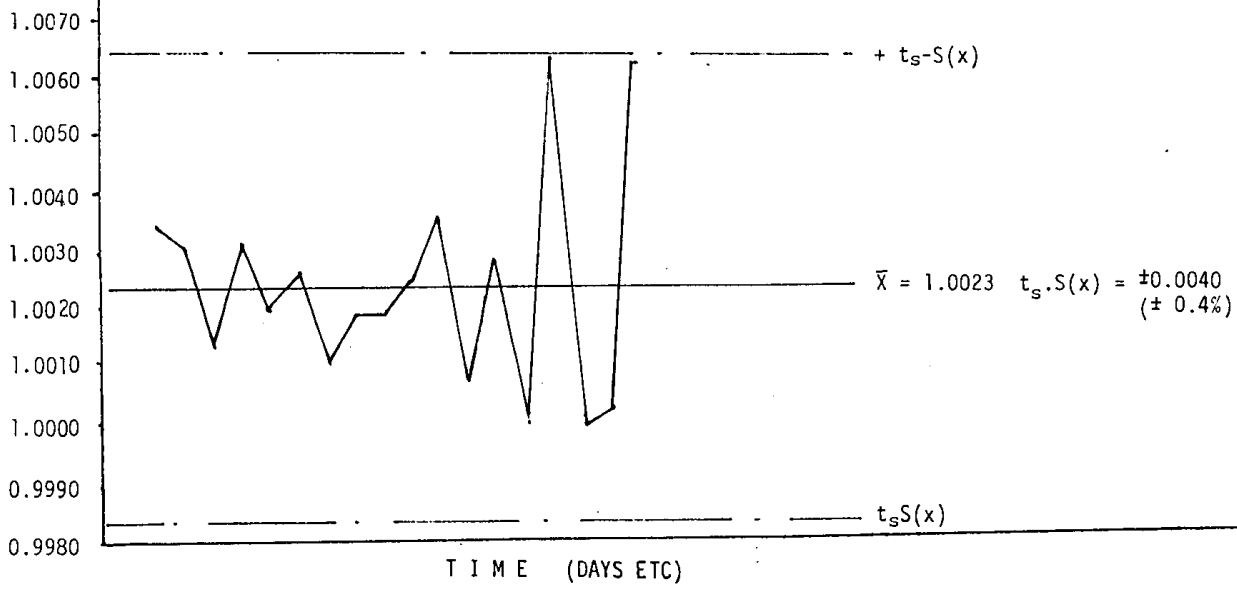
(7) Error in correcting densitometer to tank temperature
(8) Density of vapour C3 = 2.183kg/m³
(9) Density of Vapour = $\frac{288 \times P_v \times 44.097}{T_v \times 1.013 \times 27.6451}$
(10) = $(2.2^2 + 0.5^2 + 1.6^2)^{1/2} = 2.8\%$

FIG.4

COMPARISON BETWEEN METERS AND SHIPS (CALIBRATED)

IN COLD PROPANE SERVICE

(95 in 100 within $\pm 0.4\%$)



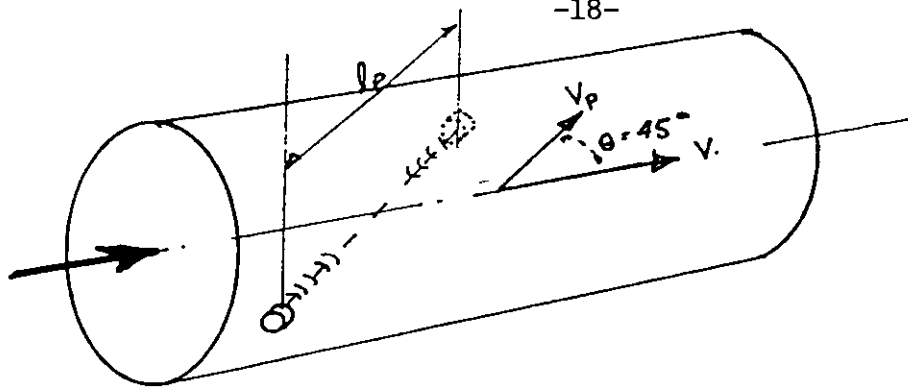
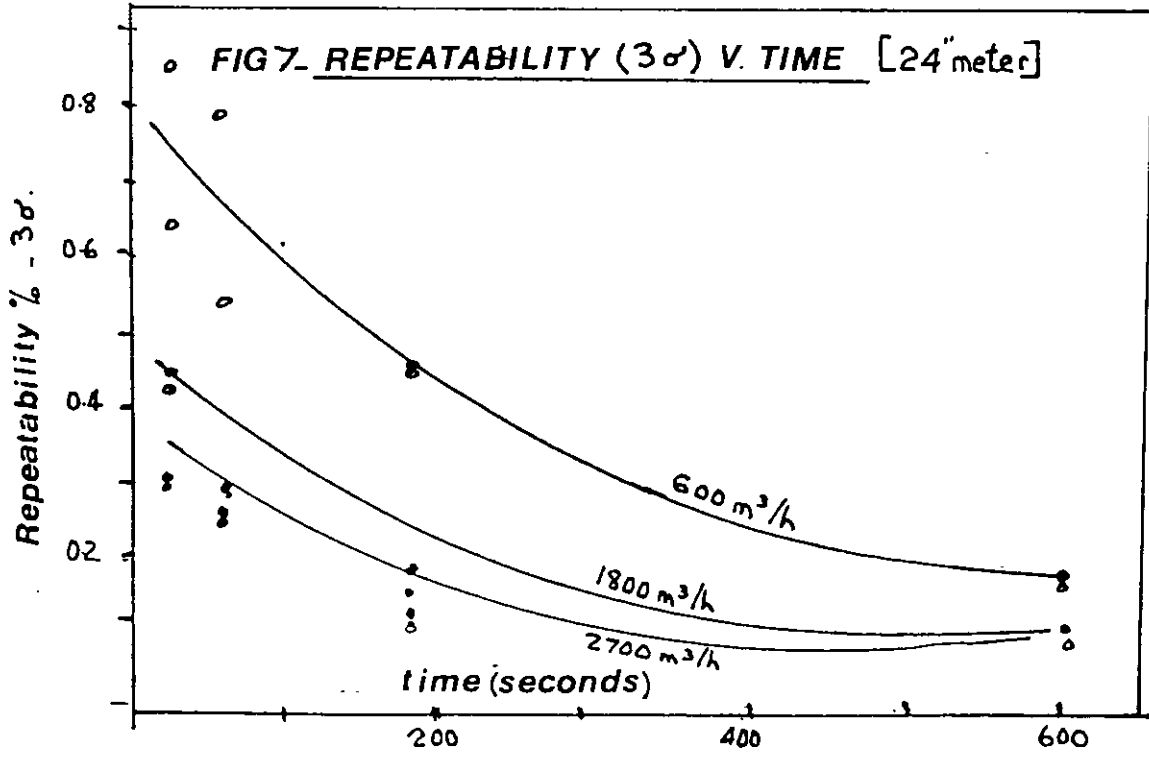
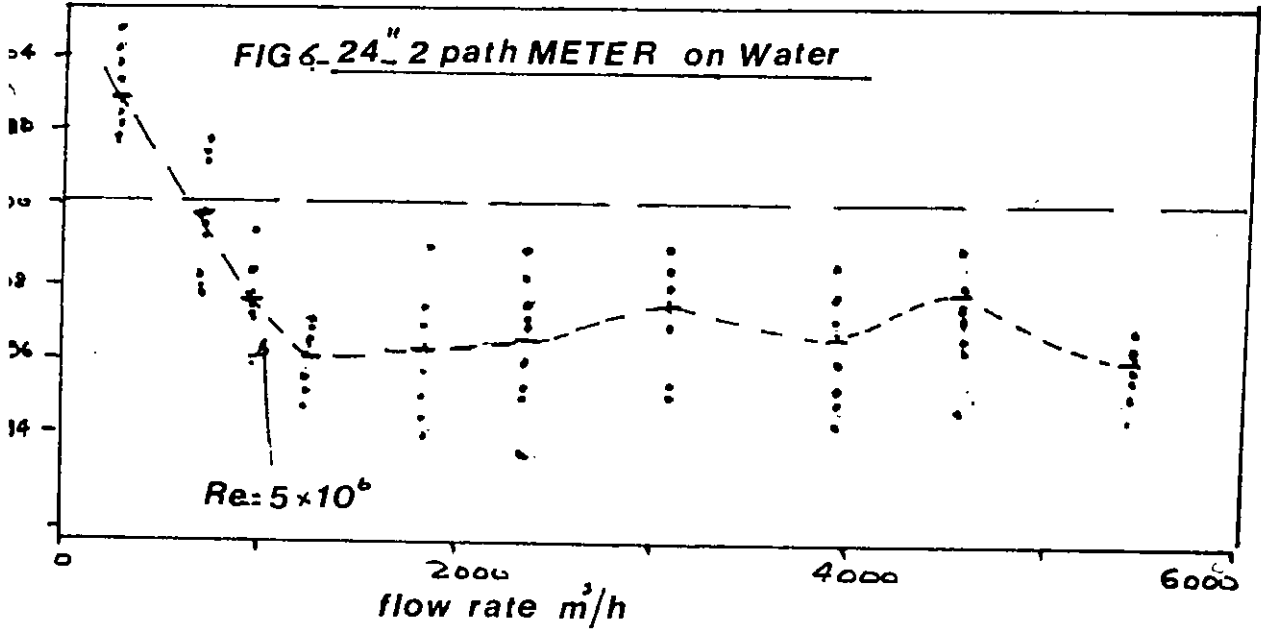


FIG. 5 Principles



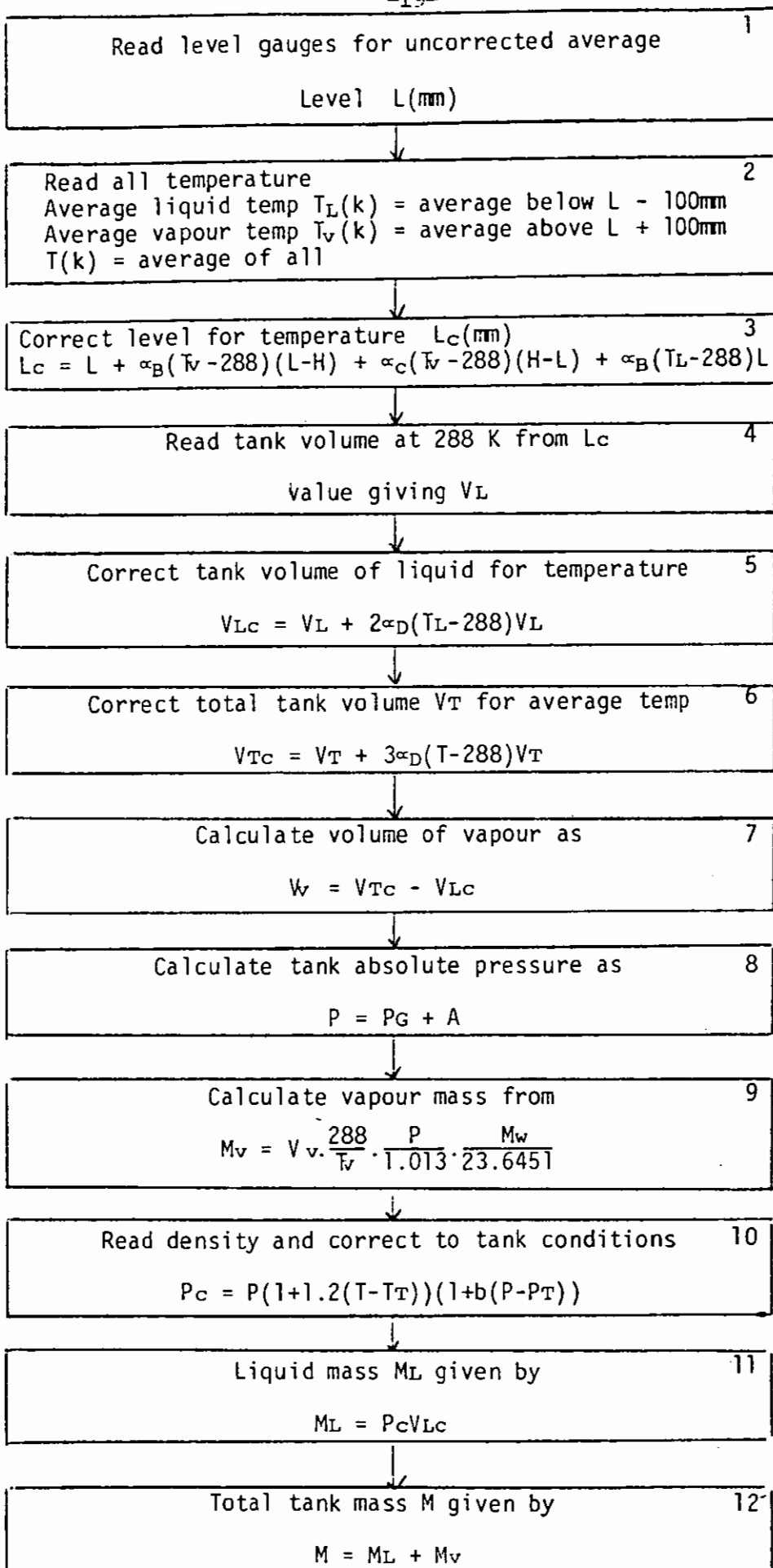


FIGURE 8

TANK INVENTORY CALCULATION

FIG 9. VORTEX METER - Bluff Body Design

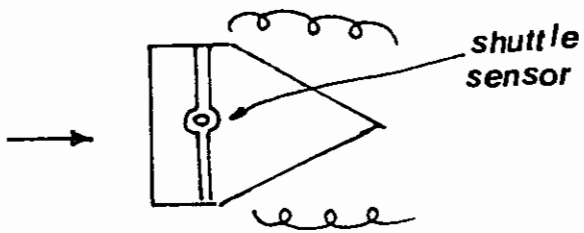
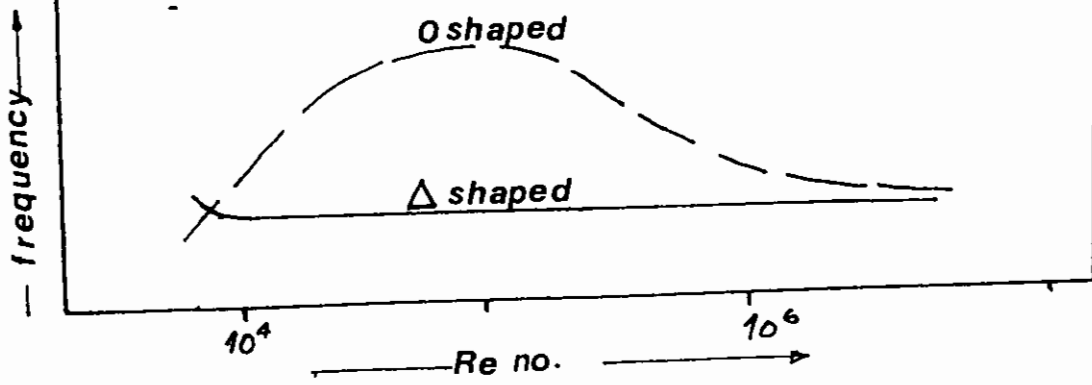


FIG 10 12" VORTEX METER - LINEARITY

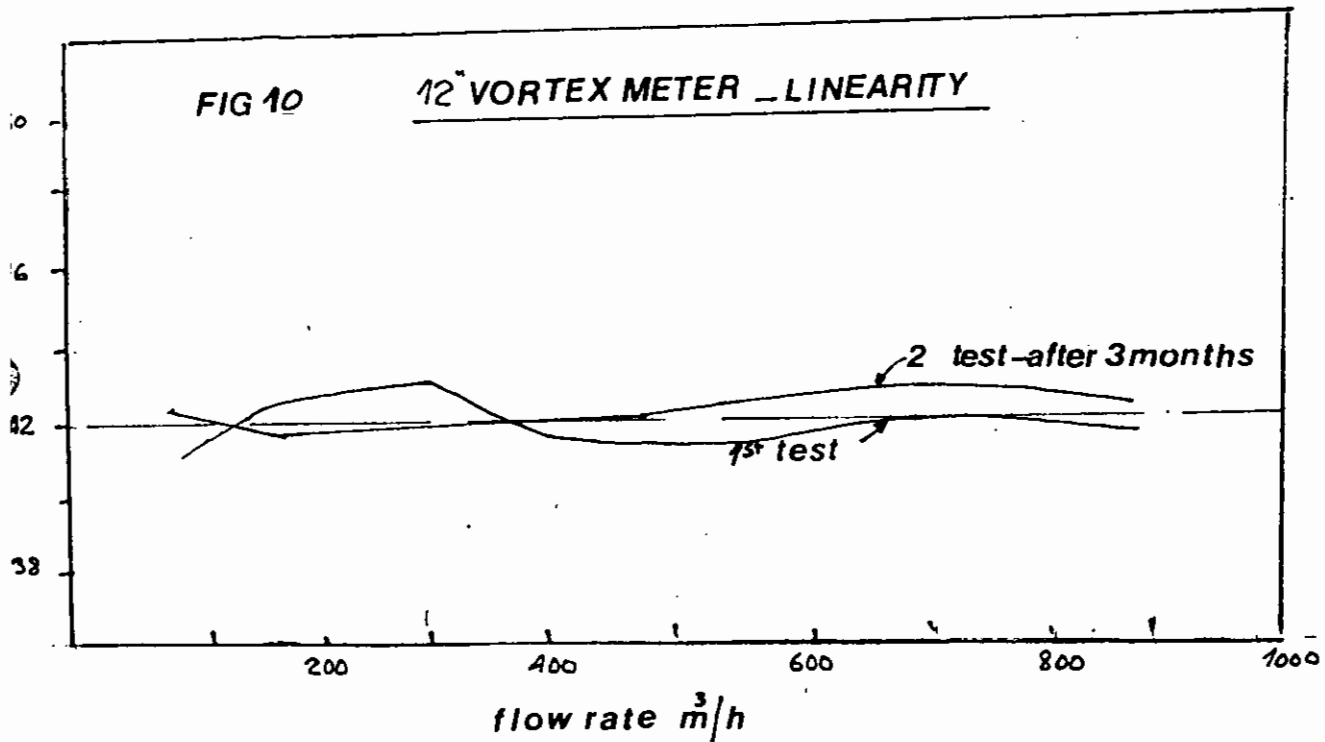


FIG 9. VORTEX METER - Bluff Body Design

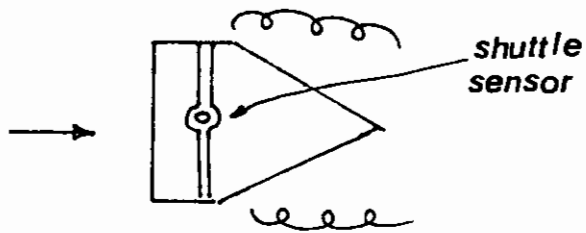
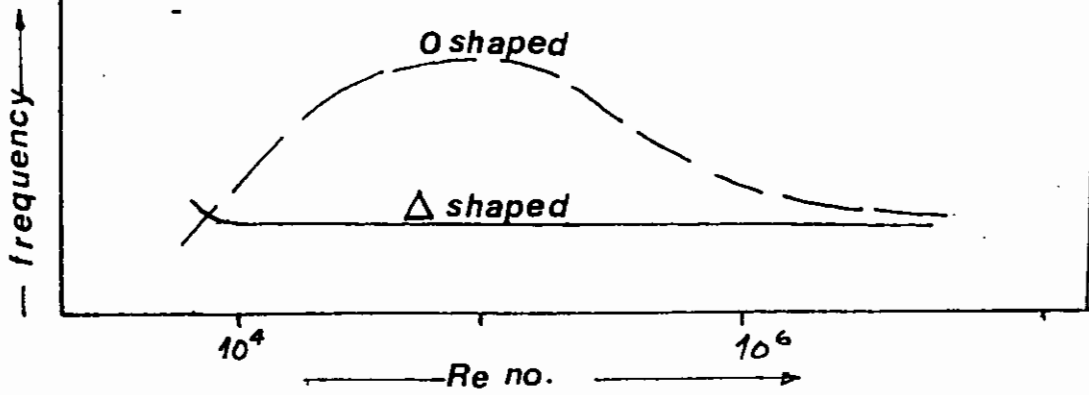
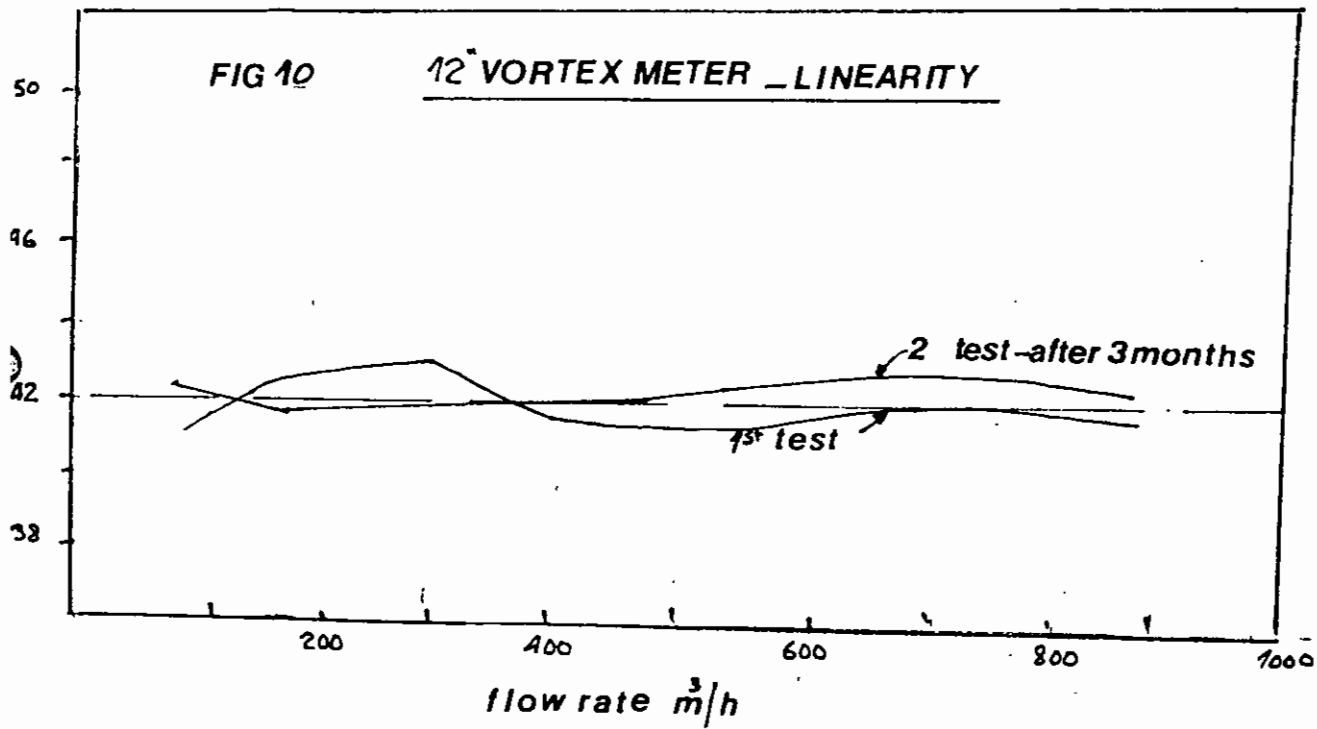


FIG 10 12° VORTEX METER - LINEARITY



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.