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Norske Sivilingeniørers Forening

Measurement of Gas and Liquids

7. - 10. juni 1982

Rogaland Distriktshøgskole, Stavanger

H B Danielsen
Oljedirektoratet

Ettertrykk kun etter skriftlig tillatelse fra
NIF og forfatteren

THE COST OF SYSTEMATIC ERRORS

In accordance with good and accepted principles of teaching, I will use my part of the opening lecture to motivate you for this course.

One of the ways to do this, may be to quantify the effects of metering errors in terms of money.

Let us look into the money-flowrate of the metering systems used in connection with production of oil and gas from the continental shelf:

- A typical flowrate for an oil metering system is 300 000 BPD. At current prices, this is equivalent to a gross flow of money of the order of 60 million Nkr/day or 22 billion Nkr/year
- A typical flowrate for a gas metering system is 40 million SCMD. At current prices this is equivalent to 40 million Nkr/day or 14 billion Nkr/year.

These flowrates of money can be combined with the relevant percentage of systematic error in the metering system, and the economical effects of bad metering will become evident. Let us have a quick look at potential sources of systematic error in gas measurement. I will limit myself to two examples:

Errors in density metering

A little simplified, the flowrate through an orifice meter for gas is calculated as:

$$Q_m = \text{Constant} \times \sqrt{\text{Diff.pressure} \times \text{Density}}$$

Normally, the conditions are so that it is necessary to measure density by a densitometer. One of the problems with densitometers are that you cannot locate its sensor in the gas stream at the upstream pressure tapping, where it should be ideal. So the sensor of a densitometer is located somewhere else, in a by-pass or downstream of the orifice plate. A limited number of different arrangements are in use in custody-transfer metering systems. Experience has shown, however, that at least one of these arrangements can change the temperature of the gas to such an extent that a systematic metering error of 0.2 - 0.3% can occur. This will be an ever lasting error in the metering system and will represent up to 4,3 million Nkr/year for the gas metering system above.

Errors caused by meter-tube and orifice plate

Shown on the overhead projector is a table of errors caused by non-standard conditions of orifice plate or metertube for a 4-inch meter. Since the exact basis for the numbers are not given, the % difference must not be regarded as generally correct for all metering systems. It should be noted, however, that a liquid build up in the metertube may cost the seller of gas the sum of 4,5 million Nkr/day.

Norwegian Society of Chartered Engineers
Rogaland Regional College.

MEASUREMENT OF GAS AND LIQUIDS

Rogaland Regional College, Stavanger

7.-10. June 1982

Review of Mass Measurement

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1. Introduction

The ability to measure quantities of gas and liquids accurately is very important to the industry. In the case of process measurements, accuracy is the basis for good production optimisation. Its perhaps not obvious that absolute accuracy is necessary in custody transfer measurements, since the errors would tend to average out during the long run. The problem is however that nobody argues when he gets more than he is paying for, so there will be a systematic error in favour of the bayer.

In the case of hydrocarbons the interesting property is normally weight (mass) or energycontent.

This lecture is restricted to mass determination by static or dynamic methods. Of course the time is too short to cover all available methods. Emphasis is put on the necessity to determine the density accurately, because one can see many mass determination systems where tremendous effort is put into deteurmining the volume, where as the density measurement is not very accurate.

2. Density

As already stated in the introduction, determination of any measurable quantity is a combination of measuring it with a suitable measuring equipment and the calibration of this measuring equipment itself. The calibration is a comparison with higher accuracy equipment preferably traceable to international standard units i.e. meter, kg a.s.o.

In the case of density there is no standard unit. Density is defined as mass per unit volume and the only way to determine true density is by measuring mass and volume of a sample of the gas or liquid in question.

Since we are interested in the density only as a means of determining the mass, there is an absolute need for measuring the density at the same conditions as the volume.

In practice this is not always possible and one has to compromise by measuring at other than normal operating conditions and establish the density by calculation.

2.1 Areometer

The use of an areometer is shown in fig. 1. The method is suitable for liquids under atmospheric conditions. The principle is that the mass of the displaced liquid is the same as the total mass of the areometer. The areometer mass is known and the scale is really an inverse volume scale. The volume of the areometer has to be temperature compensated.

2.2 Displacement density meter

Fig. 2 shows a density meter working with a fully immersed body. The buoyancy (weight of displaced fluid) of this body is measured by an automatic measuring system. This meter can be used on line both for liquids and gases. Its mainly used on atmospheric gas. On high pressure gas

application the mass of the immersed body tend to be to large compared with the buoyancy.

2.3 Gravimetric

Fig. 3 shows a density meter where a length of pipe is weighed. The fluid, liquid or high pressure gas, flows through the pipe at normal operating conditions. After temperature and pressure compensation the pipe volume is known and the density can be calculated at operating conditions.

2.4 Vibrating tube/cylinder

Fig. 4 shows an example of the most widely used principle for on line measurement of high pressure gas density. The measuring element consists of a cylinder vibrating at its resonant frequency in vacuum. If the cylinder is surrounded by gas, the gas will also vibrate and the total vibrating mass is changed. This cause the vibration frequency to change and the frequency can therefore be used as a measure of gas density.

Exactly the same principle applies to the vibrating tube used for liquids, shown in fig. 5.

2.5 Calibration

All the densitometers mentioned above have to be calibrated against gases or liquids of know densities. Most of the densitometers are also influenced by other properties than the density alone, often in a way that is not very well predictable. This means that the densitometers should be calibrated under the same conditions as normal operating conditons.

In the case of hydrocarbons this is often almost impossible. Hydrocarbons consists of a nearly infinite number of

components in different combinations, which is practically impossible to simulate in a calibration laboratory.

The normal procedure is to calibrate with some well known "standard" gas like nitrogen or pure methane.

The PT ρ relations for nitrogen and pure methane are well established so it is easy to calibrate a densitometer against one of these gases at different pressures and temperatures. The temperature can be held at the normal operating temperature and the pressure effect on the densitometer is normally small.

With other hydrocarbons however, the only way of accurately determining the density is to go back to the basis of mass and volume. Of course, it is impractical to do this for all combinations of hydrocarbon mixtures. The practical solution is to make a set of experiments and interpolate mathematically for other compositions.

So far the world database for these relations is not extensive enough to give very accurate prediction of the density of a given mixture. Different organisations are using equations giving different results by as much as 1 or 2% for a given composition.

For a given installation where the accuracy is important, for inst. in the North Sea it would be worthwhile calibrating the densitometers with the actual process gas in the laboratory using the basic method of mass and volume. For high pressure gas this can be done with an accuracy of a few tenths of a percent by weighing a "bomb" filled with gas. Fig. 6.

Up till now the normal procedure has been to use either the nitrogen calibration curve directly, the methane curve directly or a calibration curve based on calculated compensation for gas properties. The use of one or the other method is really based on belief or disbelief in the validity of the compensation.

In this type of compensation the characterizing property of the gas is taken to be speed of sound in the gas at operating conditions. The reason is that changing the speed of sound changes the coupling between the cylinder and the gas.

Fig. 7 shows a typical curve for this type of compensation. The shape of the curve is totally dependent on the design of the densitometer. That means that every new design has to go through an extensive calibration program to establish the curve.

2.6 Pressure and temperature compensation

Where the composition of the gas mixture is known to be nearly constant, pressure and temperature compensation is often used instead of densitometers. The validity of the results depends on the accuracy of the base density of the mixture and the accuracy of the Z-factor, the gas compressibility. If the base density is calculated rather than measured the results will also depend on the analyser accuracy.

As said earlier the world data base for the calculation of density and Z-factor is not very well established. Fig. 8 shows typical differences between different sources of ethylene data.

3. Volume determination

The volume can be determined by static or dynamic methods. In theory the static method is the easiest and also the most accurate. In practice the dynamic method is the most accurate, when the necessary calibration are done correctly.

Fig. 9 shows a tank filled with liquid. The volume of the liquid is determined by the tank dimensions and the level. The dimensions of the tank is influenced by the temperature of the tank wall, nearly the same as the temperature of the liquid along the wall, and also by the static pressure of the liquid. In theory, therefore, one must know the pressure and temperature at any point on the tank wall, and also any sinking of the bottom due to liquid pressure. Again in practice one have to go for a compromise and just make a few temperature measurements in the liquid.

The pressure influence is normally calculated when the tank volume is determined and the calibration tables are worked out based on the pressure exerted by the liquid normally stored in the tank and based on a reference temperature. For atmospheric tanks the pressure correction is normally too small to be accounted for if the density of the liquid changes slightly.

On the other hand the temperature correction is very important. The tank itself changes dimensions and the density of the liquid changes with temperature. The mass is the volume x density, both measured at the same conditions. Normally the tank volume table is referred to a standard temperature (0° C, 15° C, 20° C), and the density of a sample of the liquid is also referred to a standard temperature. Since the tank and liquid have different coefficients of expansion, both volume and density have to be calculated at operating conditions.

The level of the liquid is normally measured by a tape, manually or automatically. The reference point is in practice the roof of the tank or the top of a standpipe fixed near the bottom of the tank. The position of this reference point is influenced by both the liquid temperature and the vapour phase temperature, which could be quite different especially on cooled tanks. The length of the tape itself is influenced by the vapour temperature.

Fig. 10 shows a typical temperature profile along the vertical center axis of a spherical tank filled with liquified gas. It is obvious that a large number of temperature measurements are necessary to determine this profile under conditions of varying level:

With the temperature profile established, quite extensive mathematics is necessary to establish the volume of the sphere, the corrected level measurement and the average densities of the liquid and gas phase (horizontal temperature gradients are not taken into account).

Fig. 11 shows a method where the level and density measurement are combined. The differential pressure between top and bottom of the tank represents the mass per unit area and the only other parameter required is the tank area. For a true cylindrical tank only the average temperature is required to establish the average area.

For spherical tanks both temperature profile and level are required to calculate average volume and density based on the differential pressure method.

3.1 Ships tanks

Additional problems with ships tanks are caused by the peculiar shapes necessary to contain the tanks efficiently in the ships hull. It is also necessary to compensate for trim and list variations, the ships vertical axis deviation from vertical position due to uneven loading.

4. Calibration of tanks

4.1 Strapping

Calibration of tanks is normally done by measuring the dimensions of the tanks with a tape and calculate the volume at a reference pressure and temperature. This method

is used both on spherical and cylindrical tanks and even on ships tanks of different shapes.

4.2 Volume comparison

Small tanks are calibrated by direct comparison with other tanks of known volume. By using small calibration tanks and measuring the level after each filling, the volume table can be established as a function of level.

4.3 Photogrammetric

Large tanks and in particular caverns can be calibrated by photogrammetric methods. The method is basically the same as used when making land maps from air photos. The position of a large number of points on the wall are determined by stereo photo method and the volume is calculated by interpolation. This method is practicable even on rough blasted rock surfaces.

4.4 Master meter

Tank volume tables can also be established by filling or emptying the tank through a master meter. This method also requires accurate level and temperature measurements.

4.5 Level calibration

The reference of the tank level measurement is normally at the bottom end of the stand pipe. This reference position is measured when the tank itself is calibrated.

This reference is used only during manual dipping or calibration of the automatic level gange. The practical reference for the automatic level ganger is the top of the stand pipe where it is mounted. For accurate measurements this position should be temperaturecompensated.

5. Volume determination by meters

The most common meters for both liquid and gas are the turbines and the positive displacement meter.

The only meter that measures true volumetric is the positive displacement (pd-) meter. Fig. 13. This meter passes a certain volume of liquid or gas per revolution. The actual volume is affected by temperature and pressure of the meter itself. There are also corrections to be made due to leakages in or out of the defined volumes, but the basic principle is a true volumetric measurement.

A typical calibration curve is shown in fig. 14. The pd-meter is sensitive to particles passing through and it can easily get stuck, thereby blocking the line completely. Large diameter pd-meters are very heavy and expensive, especially high pressure versions.

All other meters measure flow velocity in one way or other. The turbine is already mentioned. For accurate measurements of large quantities of liquids it is almost exclusively used. It is a lot less mechanically complicated than the positive displacement meter and the repeatability is approximately the same, in the order of 0.01% for a good meter within its normal operating range.

This repeatability is referred to constant conditions in terms of flow velocity and liquid properties, especially the viscosity.

Under conditions of varying viscosity the calibration factor can vary with a few percent. These variations are theoretically predictable, and manufacturers normally supply "typical" curves showing the viscosity dependence. However, machining tolerances are so critical that two meters never give identical curves, and thus every meter needs calibration.

Fig. 15 shows "typical" curves supplied by a manufacturer and fig. 16 shows actual curves for one of the meters from the same manufacturer.

One of the biggest advantages of both positive displacement meters and turbine meters is the digital output. The number of pulses are a direct representation of the number of revolutions of the meter, and there is no need for a transducer that is inevitably introducing additional errors in the metering system.

The pulse output is also well suited for the digital electronics supplied to do the necessary calculations of mass flow.

6. Calibration of meters for liquids

6.1 Master meter

The easiest method of calibration is to run the meter against a master meter. As said before there are a few factors affecting the calibration making it necessary to consider the calibration conditions very carefully. Ideally the master meter should have been calibrated under exactly the same conditions as the meter to be calibrated. This is of course impracticable, especially if the master meter is going to be used under a lot of various conditions. Since the primary factor affecting the liquid meter, is the viscosity, the normal procedure is to calibrate the master meter with different liquids covering the range of viscosities.

6.2 Tank

Calibration against a calibrated tank is the most common method for small meters. The volume of the tank should be at least equal to the meter throughput in one minute to make sure that the errors caused by standing start/stop are small. With this method the meter can be calibrated under normal running conditions.

Larger meters can be calibrated off-line by the same method. The reason for off-line calibration is that it is not practical to move big calibration tanks around to the meter installation.

Fig. 17 shows a calibration laboratory at Norsk Hydro in Porsgrunn, intended for off-line calibration at a throughput of up to $1000 \text{ m}^3/\text{h}$.

6.3 Mechanical displacement provers

Large meters are calibrated against moving piston/ball meter provers. A typical ball prover is shown in fig. 18. The volume between detectors 1 and 2 is determined to an accuracy of $\pm 0.5\%$ or better. The volume of the pipe should as a minimum represent 10.000 pulses from the meter under test, 1:200 of the hourly flow rate, and the velocity of the ball should not exceed 3 m/s.

These requirements lead to large size provers. Typical figures for a prover for 8" turbines are 20" diameter and a length of 30 m between the detectors.

By this method the meter is calibrated under normal operating conditions and where the accuracy is very important i.e. in the North Sea the provers are installed permanently so that the meters can be calibrated at arbitrary short intervals.

The pressure and temperature control of the system is just as important as when calibrating against a tank, only in this case it is much easier. With the prover lagged the total length of the calibrated volume have a constant temperature equal to the liquid temperature.

Typical repeatability of the ball prover is in the order of 0.01%. The calibration of the prover itself is done by the water draw method. The prover is filled with water

and when the ball is passing detector 1 the water is routed to a calibration tank until the ball is passing detector 2. The water in the tank is measured by volumetric or gravimetric methods.

The piston prover works by the same principle as the ball prover, but the use of a piston requires the proving pipe to be one straight length.

A very short piston prover is shown in fig. 19. This prover does not comply with the requirement that the volume is 1:200 of the hourly flow rate or that the number of pulses are minimum 10000. This is compensated by the fact that detection of the position of the piston is accurate enough to facilitate splitting of the pulses from the meter. Fig. 20 shows the principle. Over the calibration period both the total travel time and the travel time for a whole number of pulses is measured. Then the number of pulses corresponding to the calibrated volume, are calculated to an accuracy of a few decimals.

A typical size of this prover for an 8" turbine is 18" diameter and 2 m length of the cylinder.

Calibration of the prover is done by the same method as with a ball prover.

The most critical part of this prover is the sealing of the piston and the valve inside the piston.

7. Gas metering

7.1 Orifice plate

The most widely used meter for large gas quantities is the orifice plate/differential pressure transducer. The main reason for using the orifice plate is that there is no absolute need for calibration. Laboratory calibration shows that the meter factor can be calculated based solely

on measurement of mechanical dimensions and knowledge of the physical properties of the gas.

One of the problems is, however, that different standards give different results. Also even the latest ISO-standard (ISO 5167-80) issued in 1980, differs as much as 0.4% from some recent laboratory results. The standard itself gives a tolerance of $\pm 0.6 - 0.8\%$ under specified conditions. The requirements include very long straight lengths of pipe, up to 88 pipe diameters, and eccentricity of orifice installation less than 0.1 - 0.2% of the pipe diameter. The last requirement is almost impossible to fulfill in practice. Outside the limit an additional uncertainty of $\pm 0.3\%$ applies.

The quadratic relationship between flowrate and differential pressure gives a range of only 1:3 for each d/p-transducer, so that for larger ranges 2 or more transducers in parallel are required. On the other hand, since the massflow is proportional to the square root of the density, a 1% error in the density determination will give only 1/2% error in the massflow.

The calibration factor of an orifice plate is very sensitive to wear, deposits or deformation. It is also worth mentioning that most of the factors affecting the calibration, tend to give low readings.

It is necessary to inspect the orifice plate at frequent intervals. If the pressure cannot be taken off the line for this inspection, an orifice fitting similar to the one shown in fig. 21 is used. With this type of installations it is of course difficult to make sure the eccentricity specification is fulfilled.

7.2 Calibration

Even though one of the prime advantages of the orifice measurement is the ability to calculate the calibration factor, it is quite common to calibrate the orifice installed in its straight length of pipe. In this way one can avoid any errors due to inaccuracy of the standard, machining tolerances or mechanical installation.

Such calibration is dependent on very accurate knowledge of the properties of the calibration gas, especially if the comparison is referred to a volumetric measurement, since in the orifice calculation one needs the gas density.

The problem is reduced when calibrating against a sonic nozzle. Fig. 22. The mass flowing through the sonic nozzle, is mainly proportional to the square root of the gas density, as is the case with the orifice plate, and the square root of the inlet pressure.

The main advantages of the sonic nozzle is that there is no need for a differential pressure measurement and that it is by far less sensitive to mechanical wear than the orifice. Main disadvantages is the pressure loss of 10 - 20% of line pressure.

7.3 Gas turbine meters

A typical custody transfer gas turbine meter is shown in fig. 23. Typical calibration curves for atmospheric air and high pressure gas (in this case 8 bar natural gas), is shown in fig. 24. The repeatability is in the same order as for the liquid turbine, $\pm 0.01\%$. As can be seen from these curves the linear range is approximately 1:20 on high pressure gas.

The calibration shift from 8 bar to 60 bar is very small. Since we are in the same situation as with the liquid turbine, that the machining tolerances are very critical every single turbine has to be calibrated.

7.4 Calibration

Gas turbine meters are sensitive to gas density rather than viscosity. Therefore if the turbine cannot be calibrated under normal operating condition, the gas density should be as near normal as possible.

The only practical way of calibrating a large turbine meter is comparison with a master meter or a sonic nozzle.

Fig. 23 shows in principle the stepwise procedure used to calibrate a master meter at 60 bar referred to an atmospheric bell prover. In this procedure it is essential to know the PTZ-relations for the calibration gas. Ideally the gas should be the same as the process gas under normal operating conditions.

Gas meters are also calibrated with water, but this should be regarded more as a function check than a real calibration.

7.5 Positive displacement meter

The Pd-meter is extensively used for small gas quantities especially at or near atmospheric pressure. This reflects the fact that the meter is very accurate, but that meters for large quantities and/or high pressure are very heavy and expensive.

Repeatability figures are in the same order as for the turbine meter.

8. Vortex meter

Over the last few years there have been quite a development of the Vortex meter, fig. 25. The meter consists of a straight pipe with a bluff body generating vortices in the liquid or gas. The frequency of the vortex shedding is proportional to flow velocity.

Because of the mechanical simplicity of the meter the calibration factor can be calculated, based on dimension measurements. In the same way as with the orifice plate the Reynolds number is the characterizing fluid property. Therefore the vortex meter can be calibrated with water even if it is going to be used for gas metering. Depending on fluid properties the turndown ratio could be from 1:5 to 1:100.

Rounding of the corners of the bluff body is less critical than rounding of the inlet of an orifice bore. The bluff body is also normally very robust.

Pressure loss is about the same or less than for an orifice.

There are a lot of different ways of detecting the vortices like mechanical deflection of the bluff body, pressure transducers, termistors and ultrasound. Irrespective of detector principle the pulses generated are not exactly regular. This causes a problem during calibration because a large number of pulses are required to average out the irregularities.

9. Ultrasonic flow meter

An ultrasonic flow meter is shown in fig. 25. An ultrasound pulse is transmitted through the fluid at an angle to the pipe centerline. The difference in time required to cross the pipe in both directions is a direct measure of the flow velocity. It represents the average flow velocity along the path of the ultrasound pulse. This means that the flow profile must be known for the average flow velocity over the whole pipe area to be calculated.

At the sophistication level of today's electronics, the limiting factors on the accuracy is mainly related to the mechanical part of the meter. The accuracy of installation of the pulse transmitters and receivers are very critical, especially on multipass meters. These meters are intended to calculate the flow profile based on measurements along 2, 3 or 4 ultrasound paths.

The best results are of course obtained when the meter have been calibrated. The long term stability is very good since there is virtually no possibilities for mechanical changes affecting the accuracy, except, corrosion or deposits.

The ultrasonic flow meter represents no pressure loss whereas most of the other meters could require 100 kW in a large metering station.

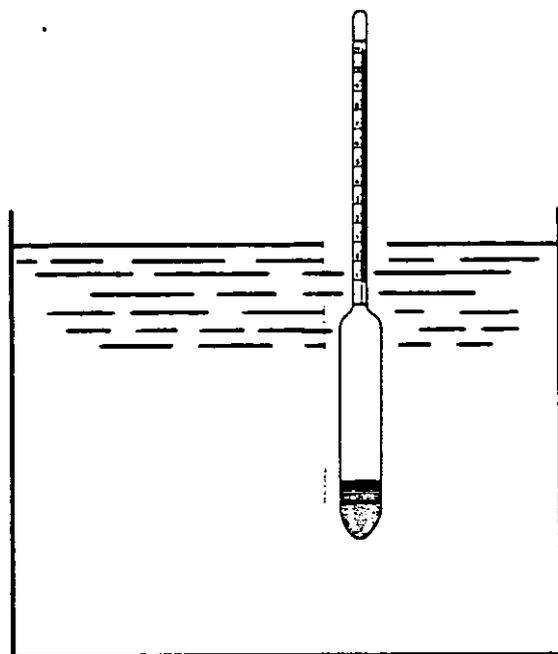


Fig. 1 Areometer

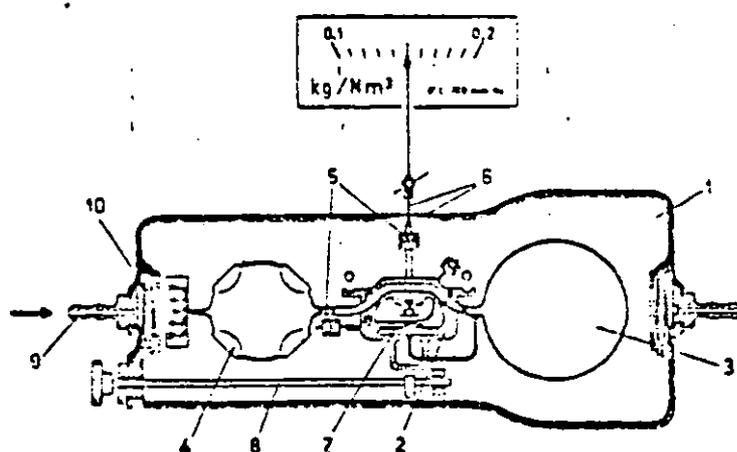


Fig. 2 Density meter for atmosphere gas

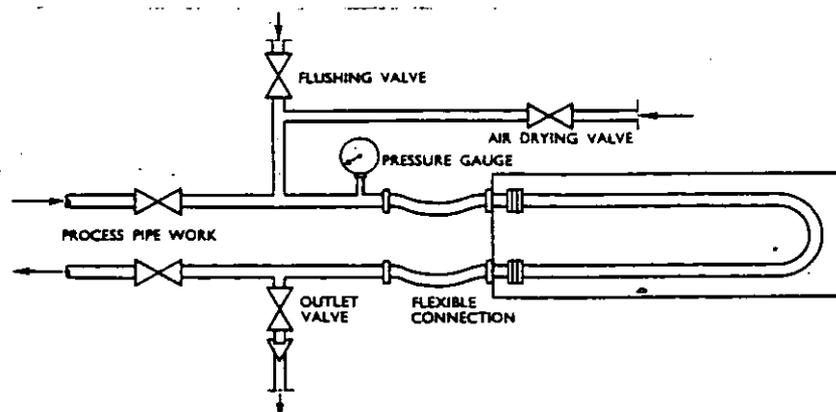


Fig 3 Gravitrol

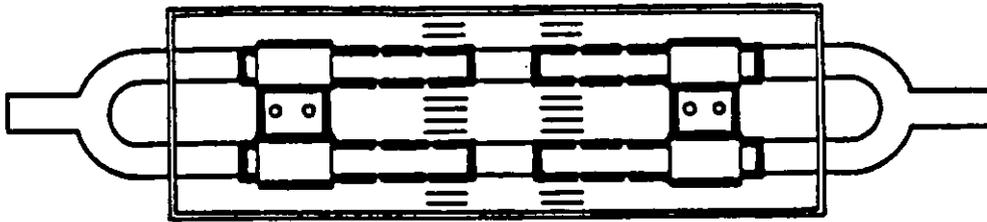


Fig.5 Vibrating tube densitometer

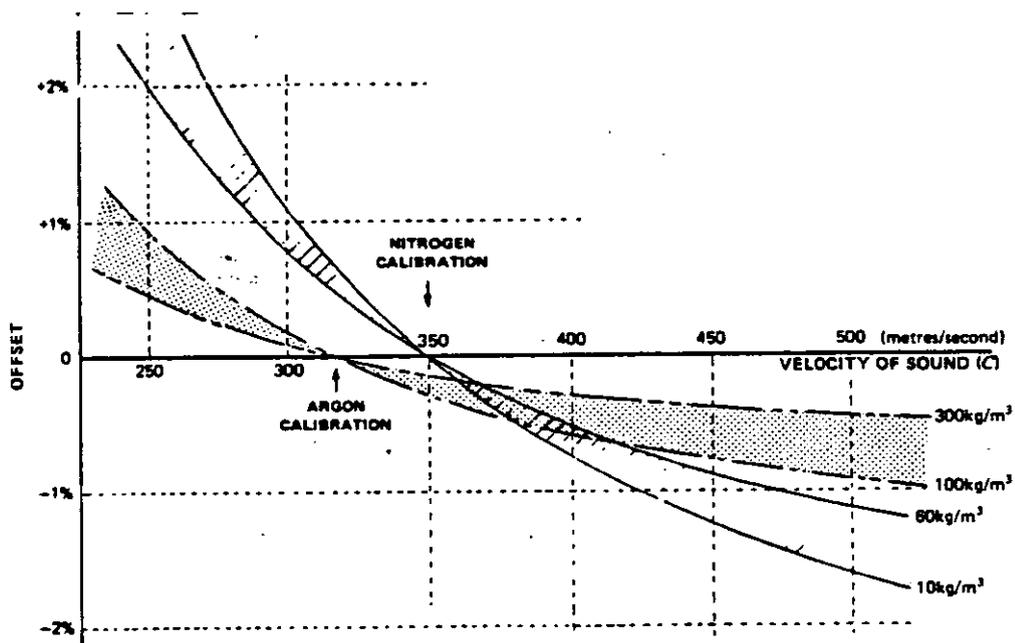


Fig. 7 Velocity of sound corrections

1 T °C	2 P bar abs	4 (P/Pn)ARG	5 (P/Pn)K-372	6 (P/Pn)K-372 (P/Pn)ARG
7.400	19.870	22.321	22.080	.98920
9.500	19.810	21.965	21.529	.98015
10.500	20.550	22.819	22.541	.98782
13.900	20.570	22.399	22.212	.99165
14.400	20.560	22.232	22.123	.99104
36.800	20.530	19.890	19.899	1.0005

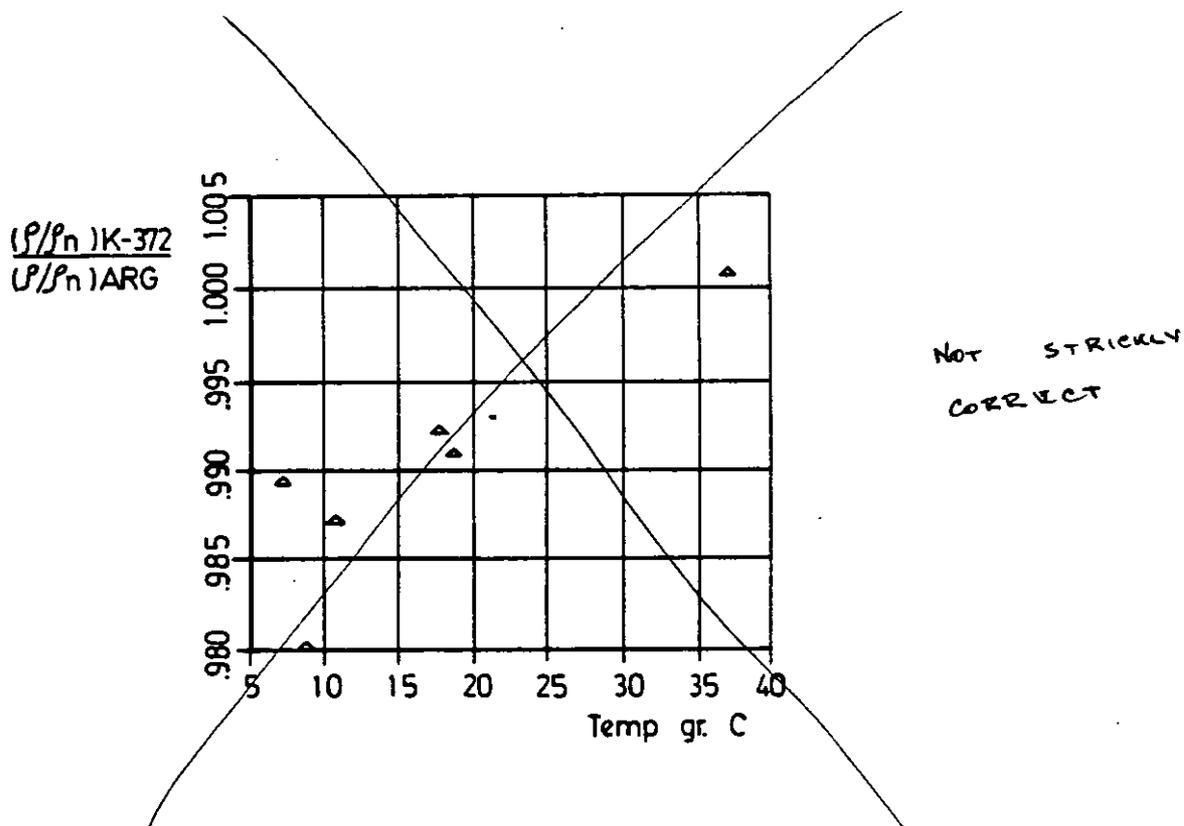


Fig. 8 Differences in ethylene data

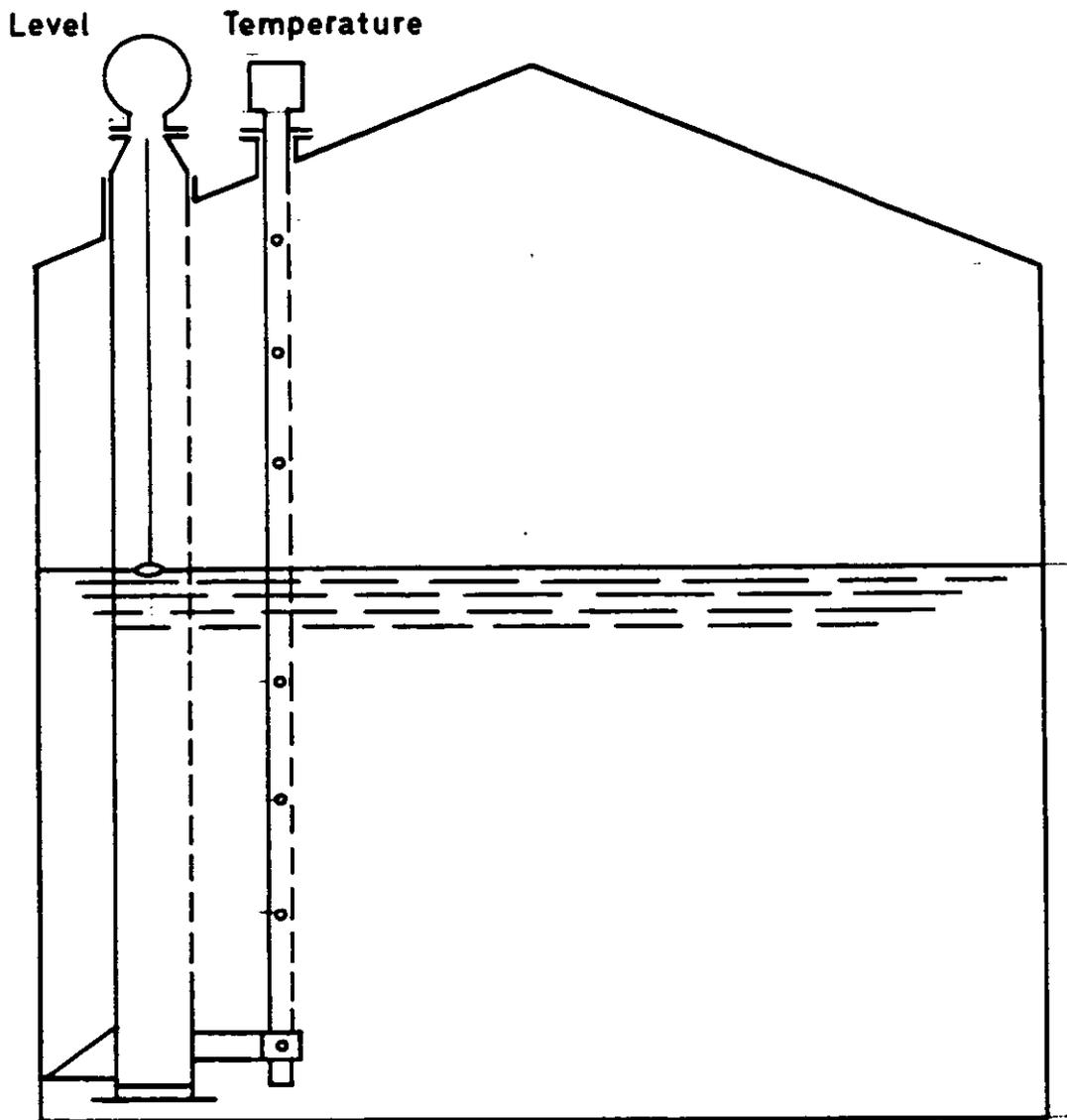


Fig.9 Atmospheric tank for liquids

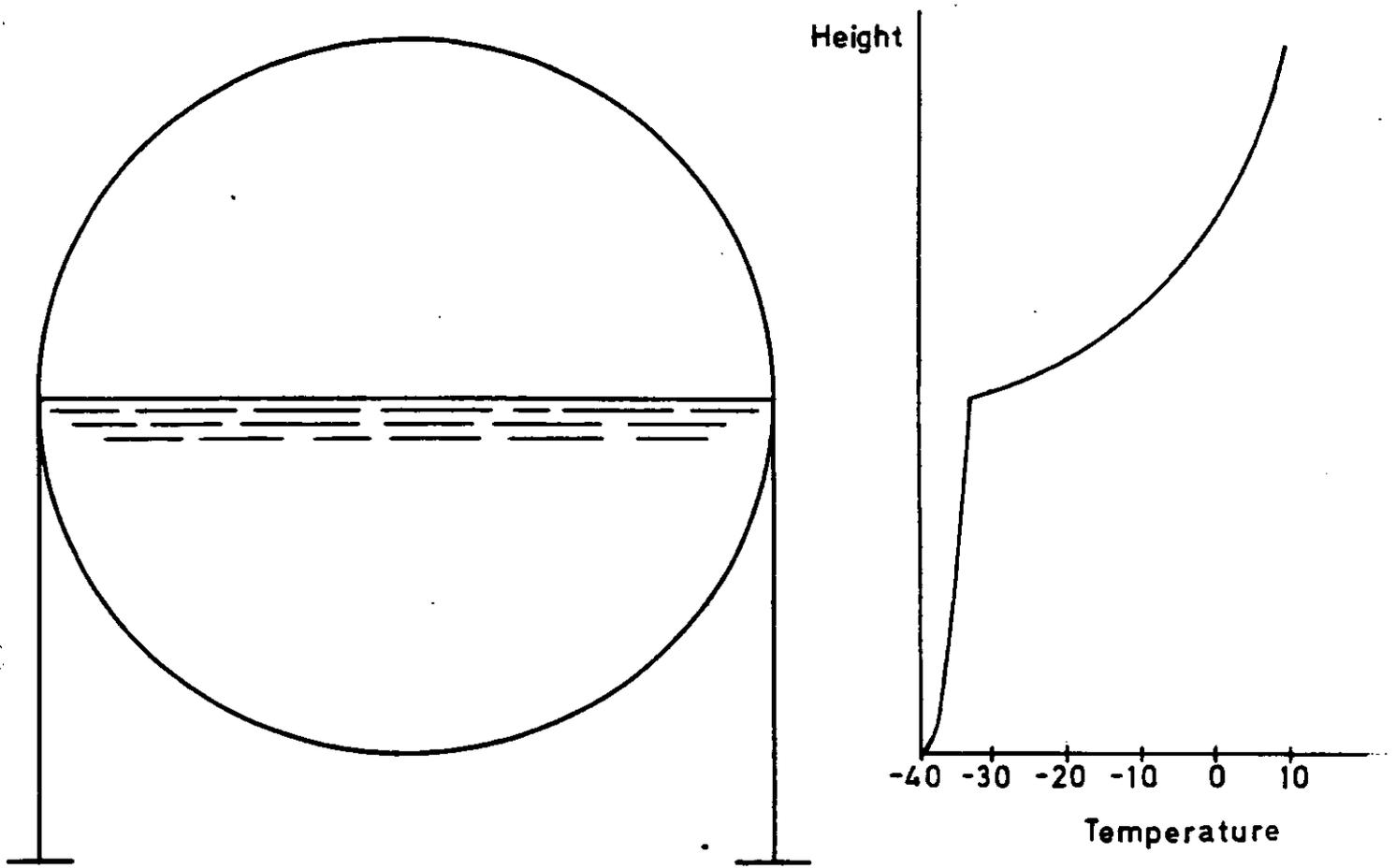


Fig.10 Temperature profile in cryogenic sphere tank

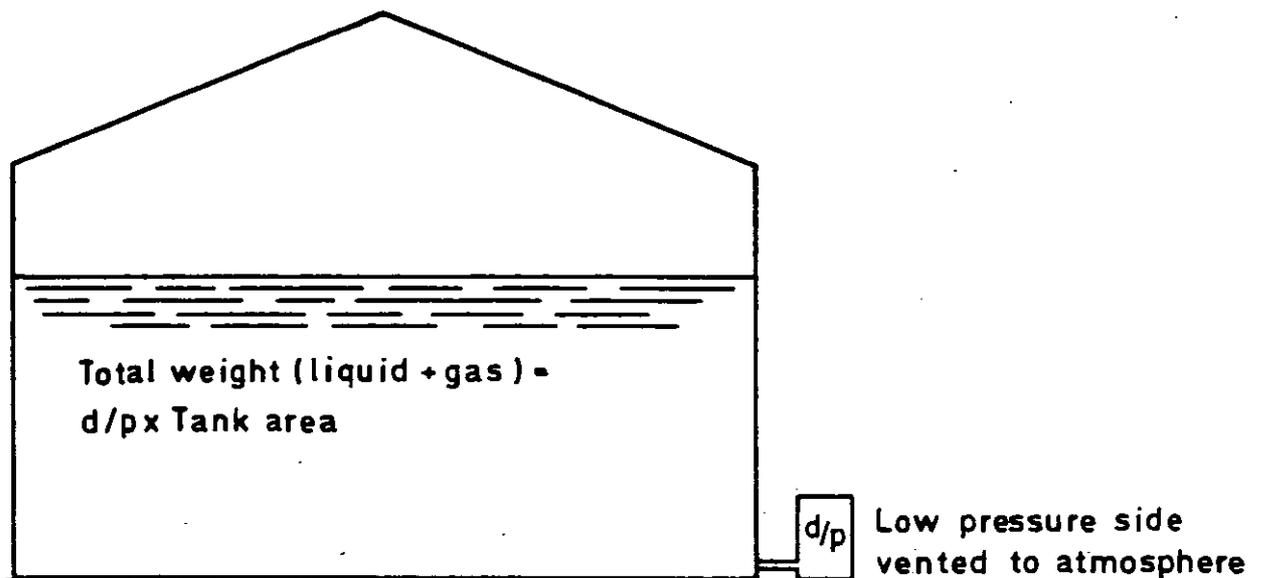
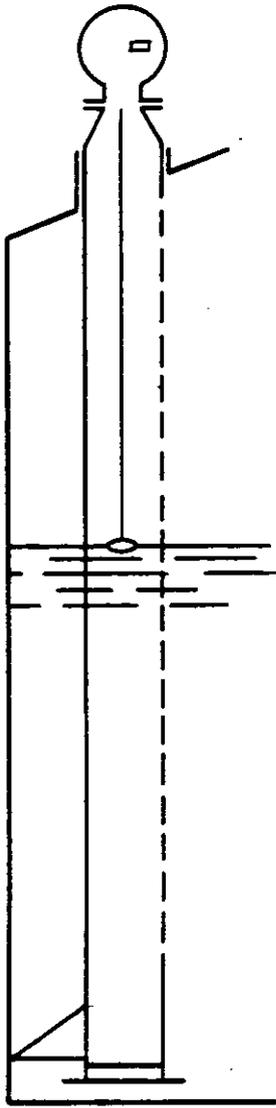
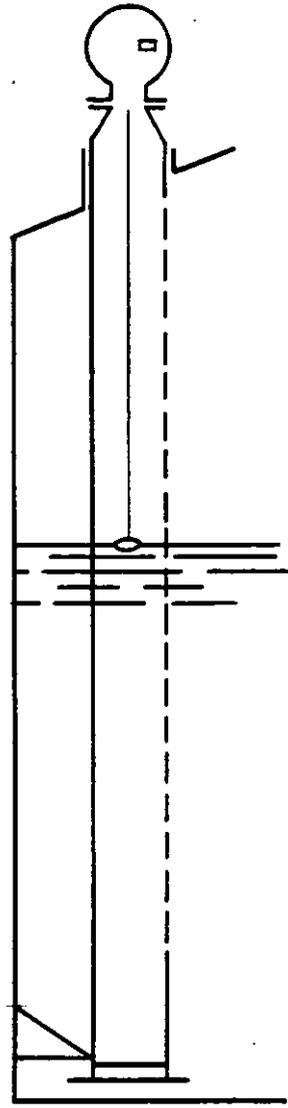


Fig.11 Mass determination by tank
bottom pressure measurement



Right



Wrong

Fig.12 Installation of level gauge

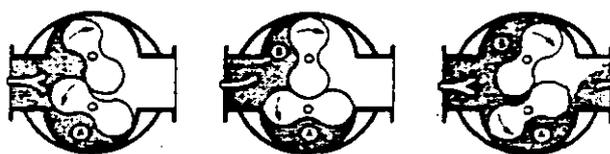
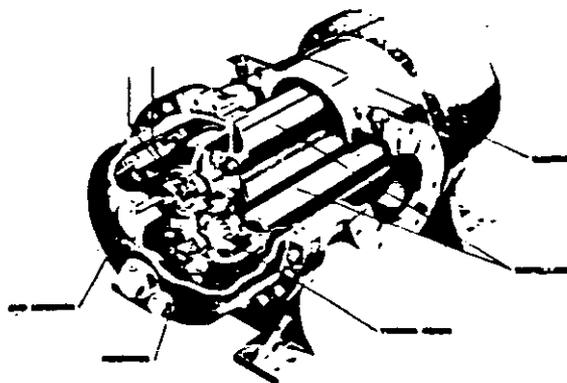


Fig. 13 Positive displacement meter

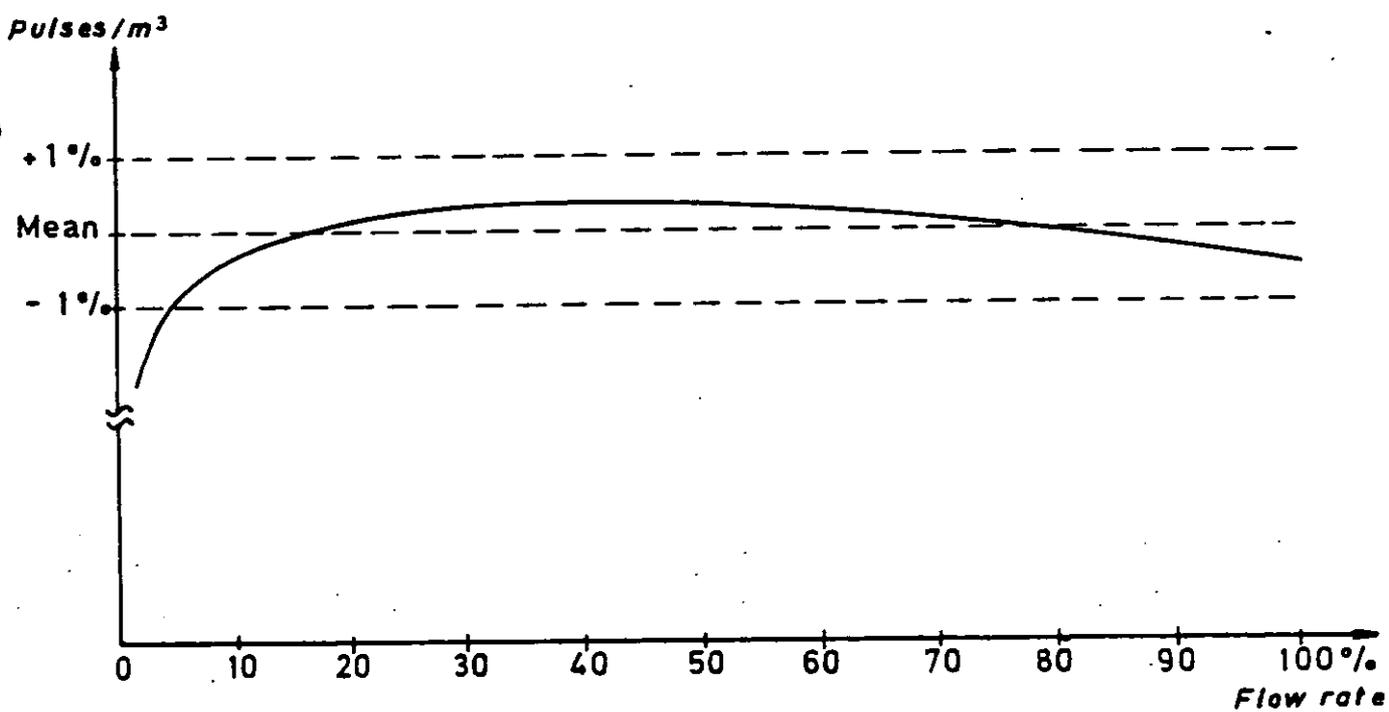
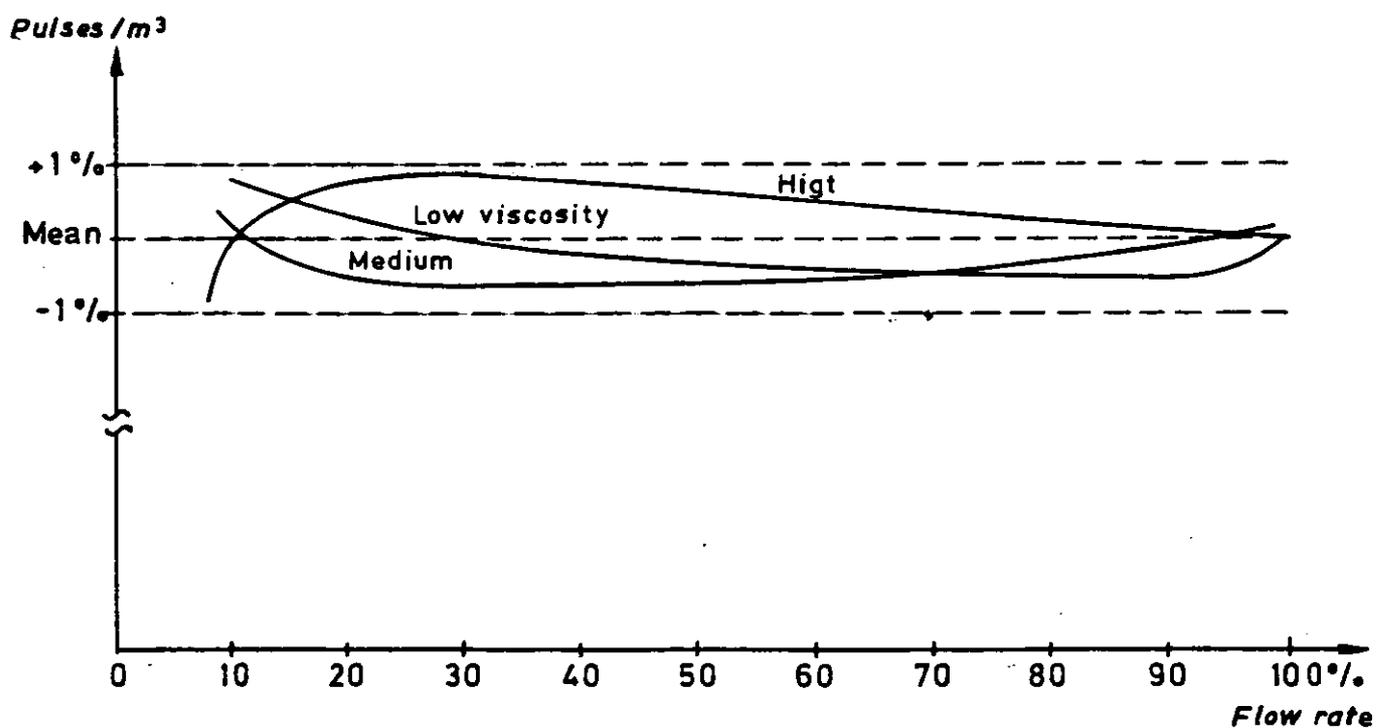
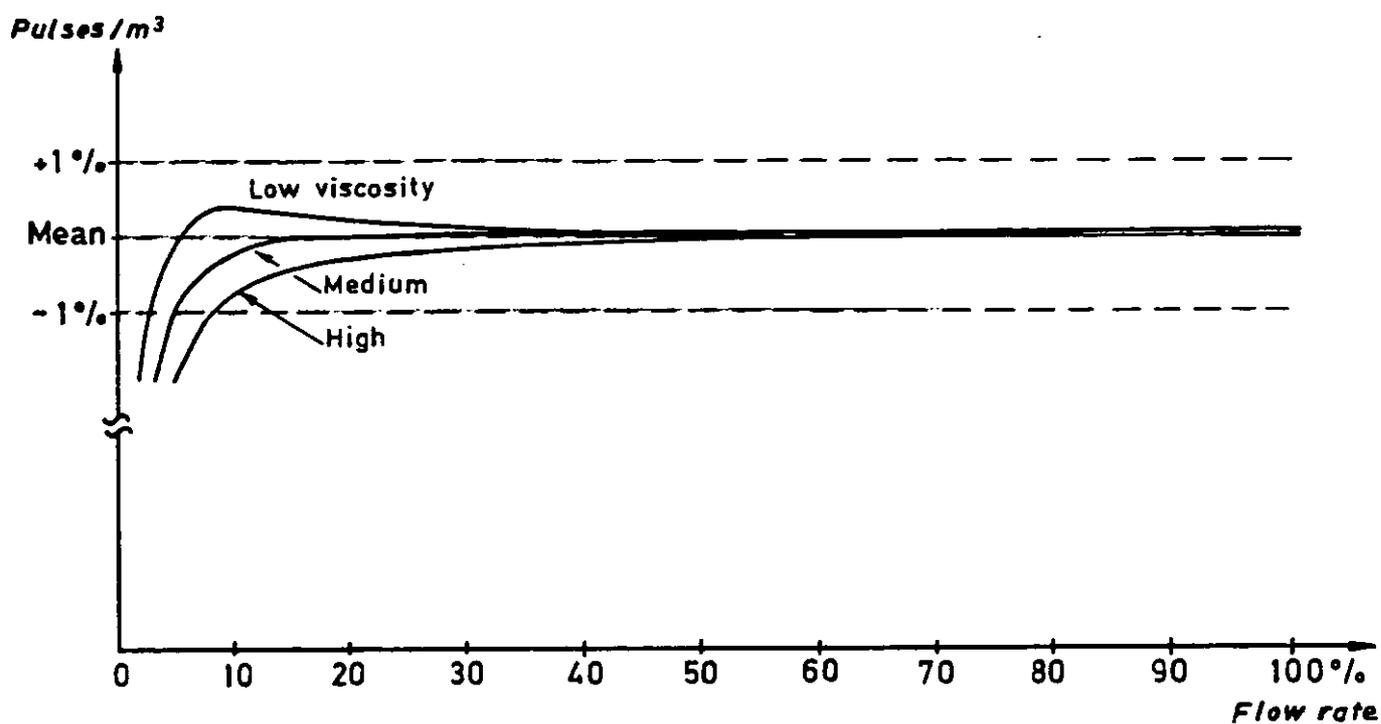


Fig. 14 Calibration curve for Pd - meter



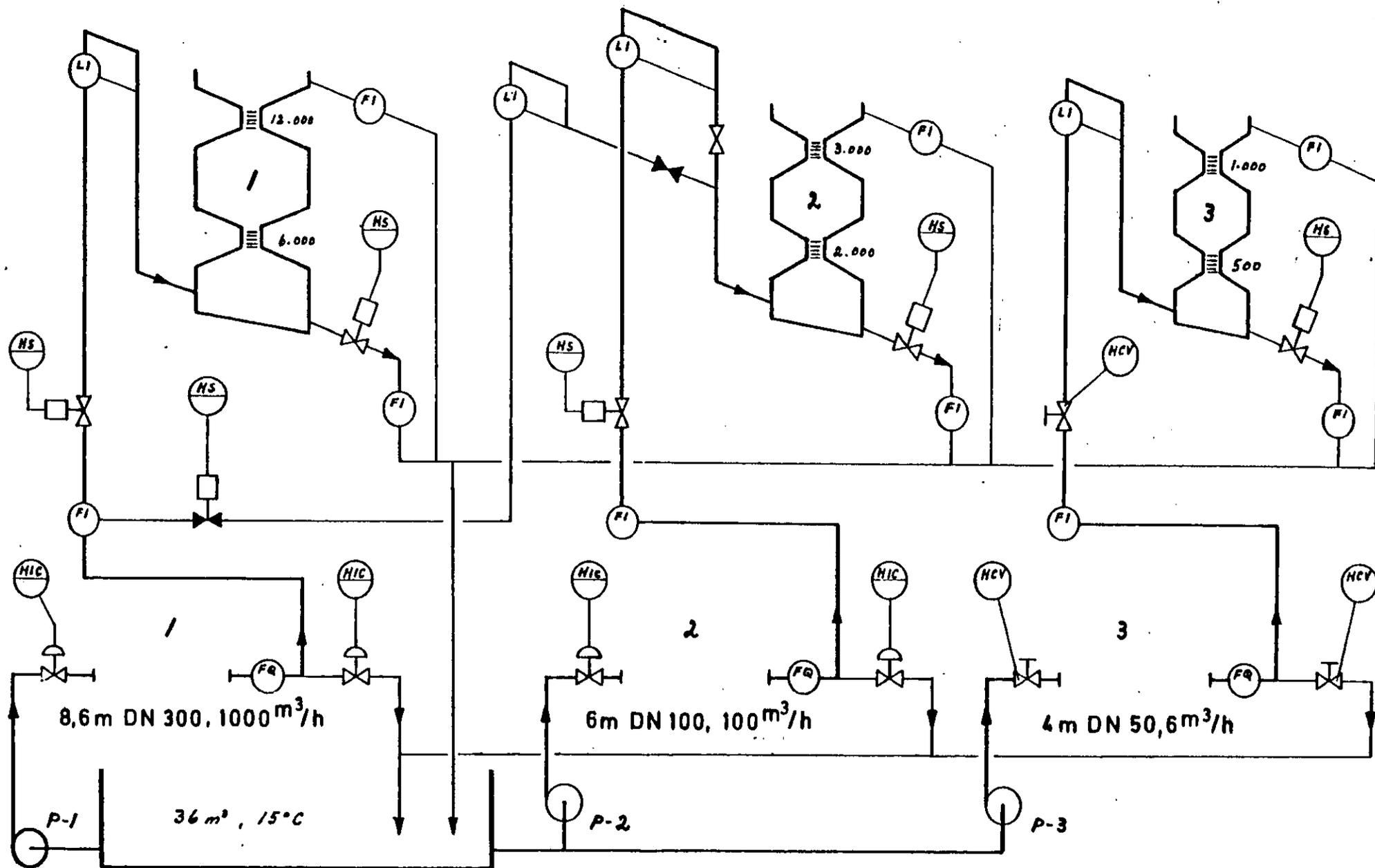


Fig. 17 Calibration laboratory
at Norsk Hydro Porsgrunn

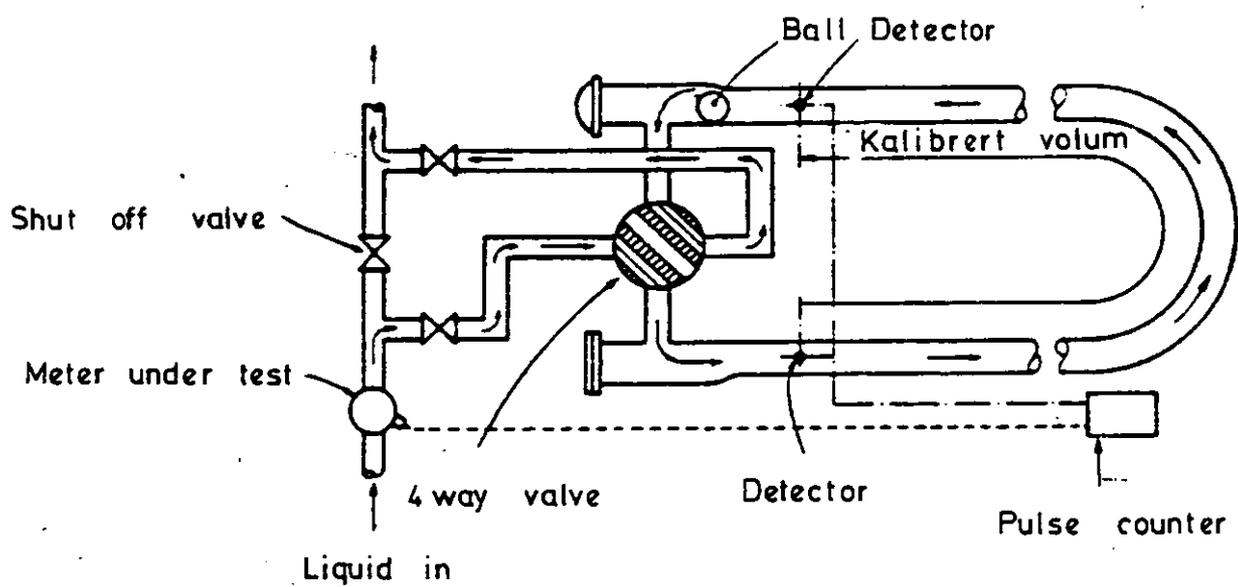


Fig. 18 Mechanical displacement prover

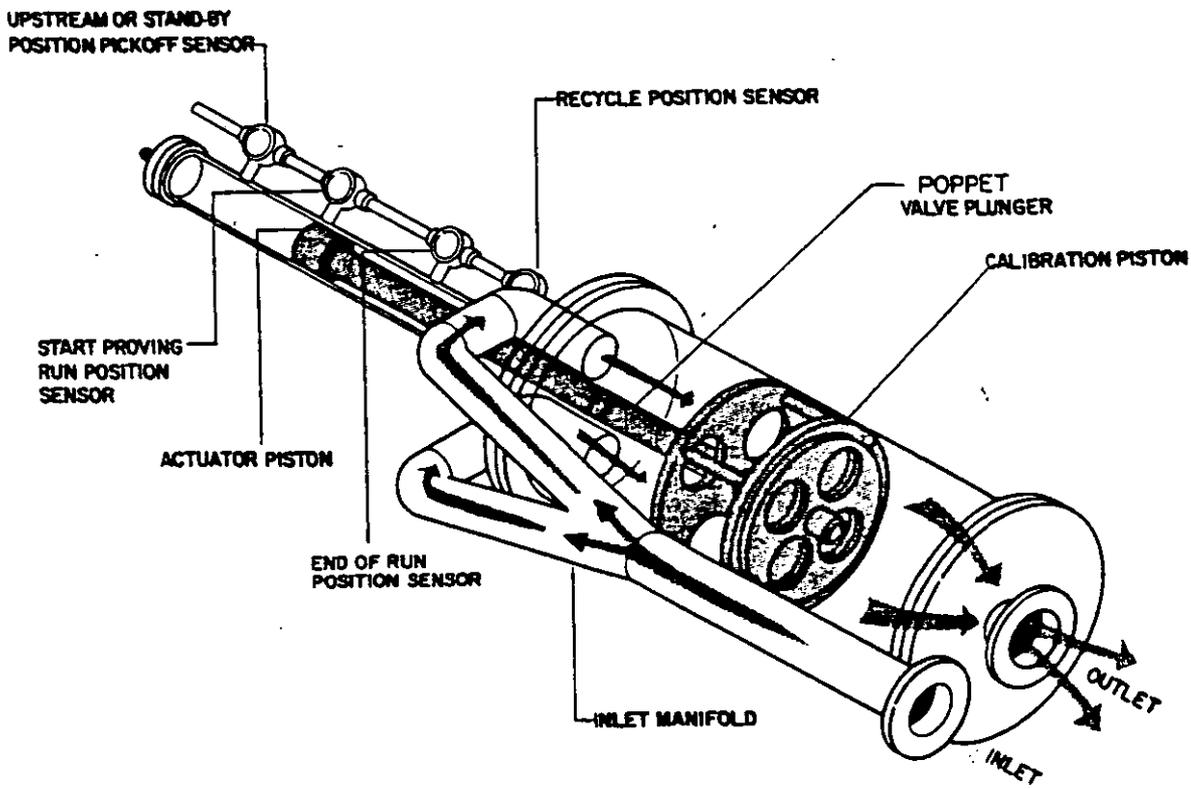
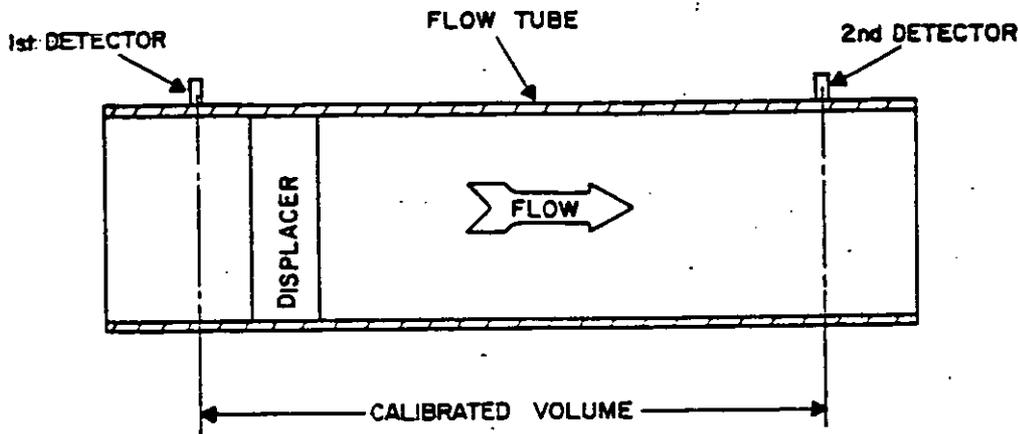


Fig. 19 INLINE BALLISTIC FLOW PROVER



SIMPLIFIED PROVER DIAGRAM

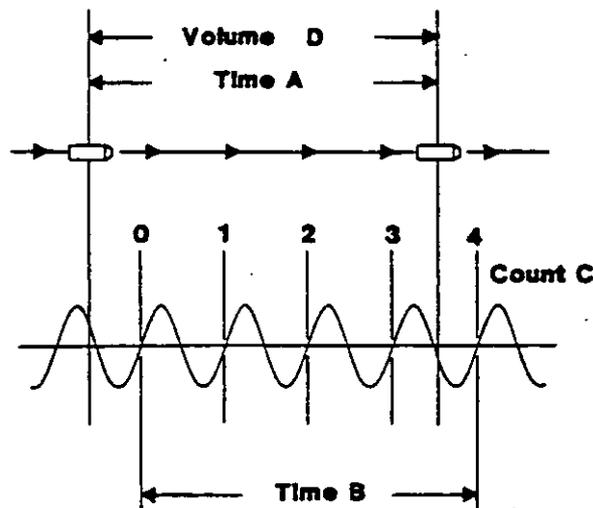


Fig. 20 Pulse splitting

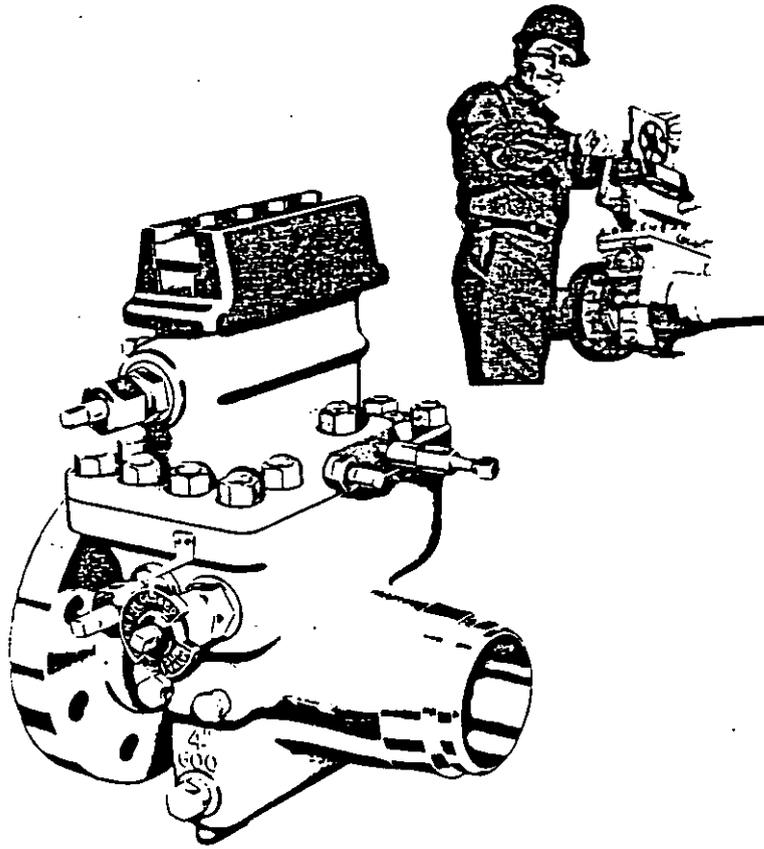
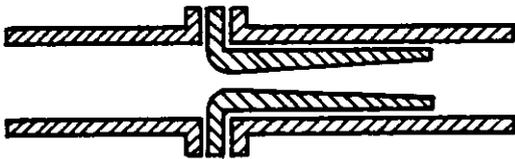


Fig. 21 Orifice inspection with
line pressurized.



Sonic nozzle

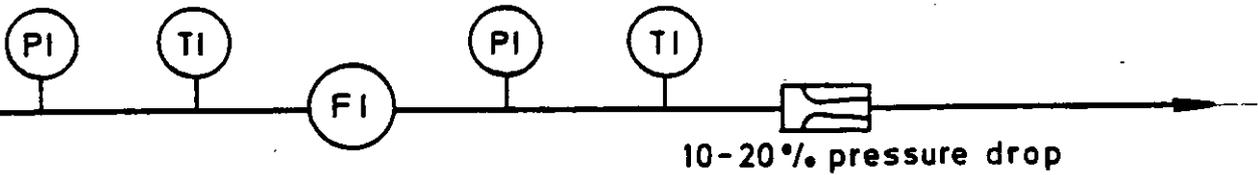


Fig. 22 Calibration against sonic nozzle

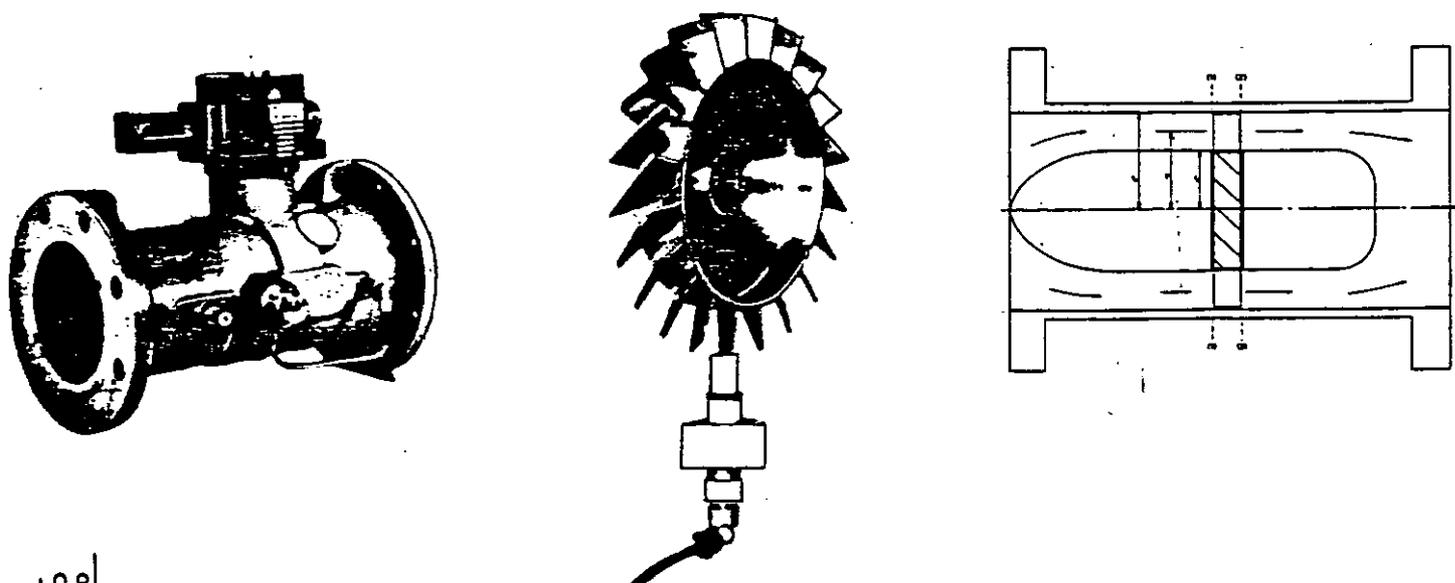


Fig.23 Gas turbine meter.

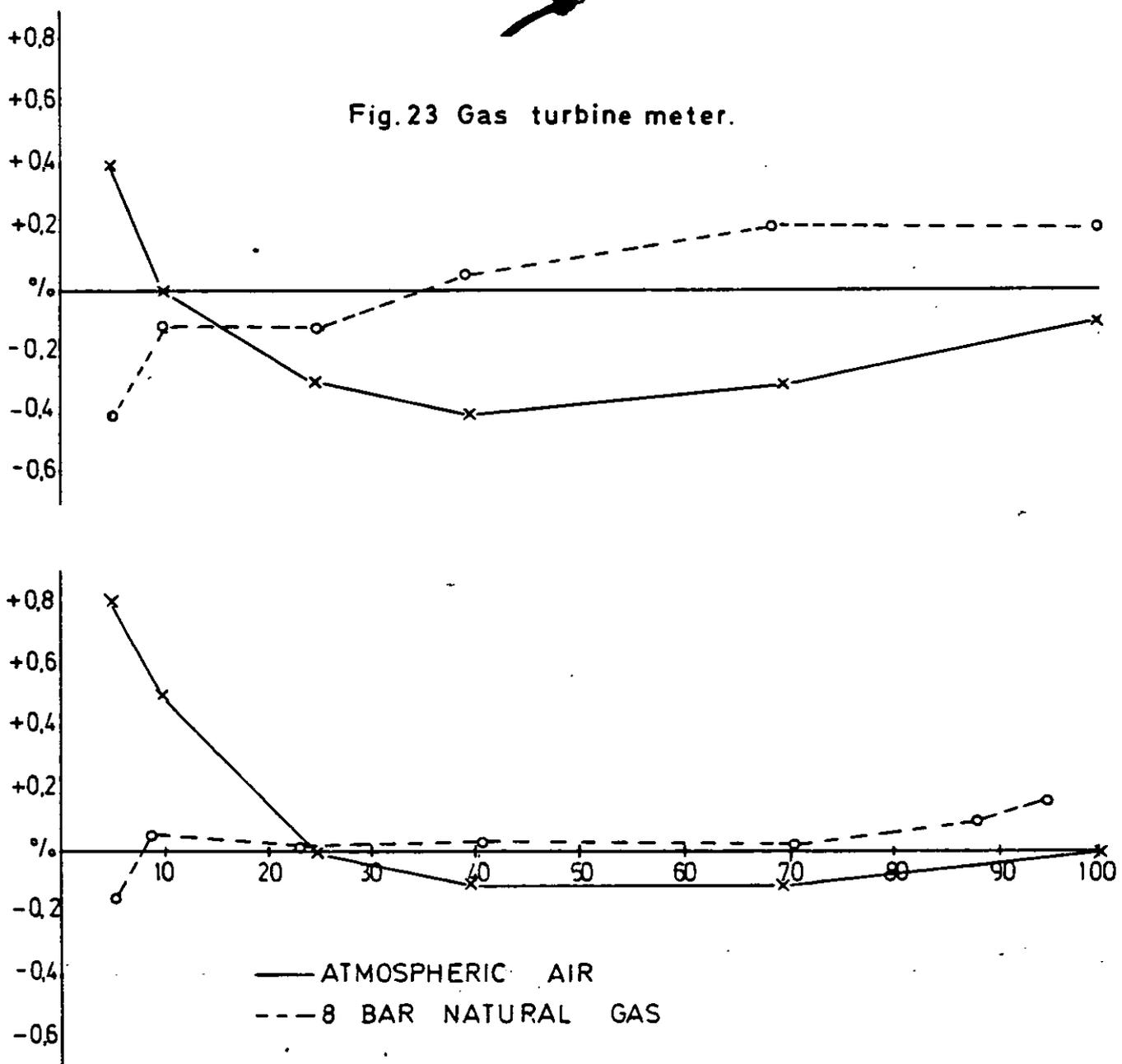


Fig. 24 Calibration curves for two identical meters.

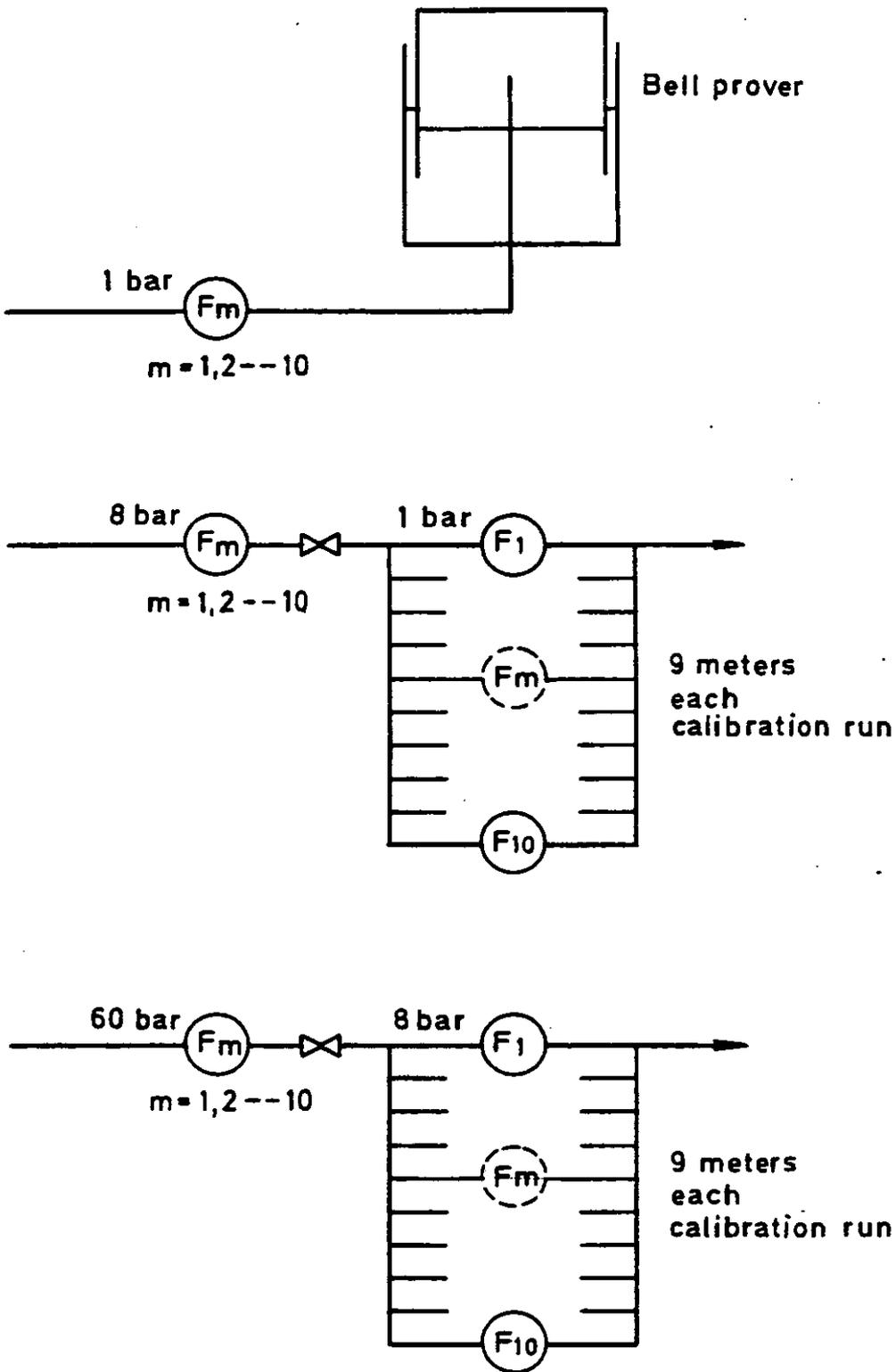


Fig.25 Stepwise calibration of high pressure turbine meter

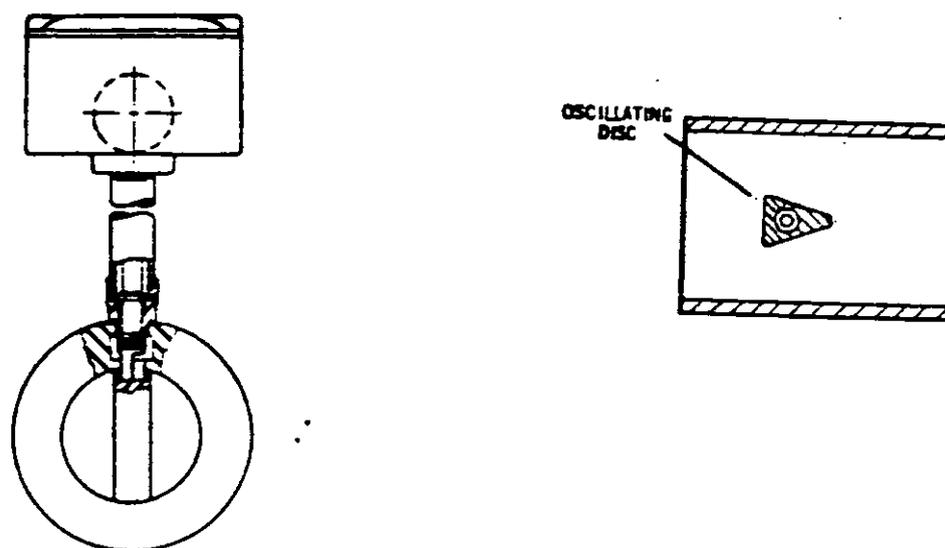


Fig. 26 Vortex meter

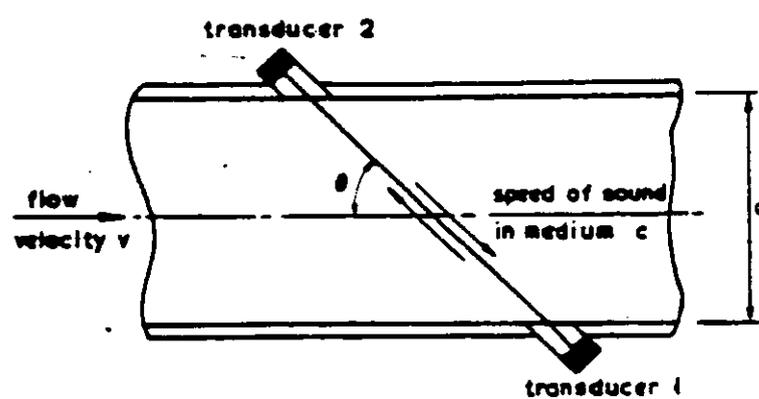


Fig. 27 Diagonal-beam
Ultrasonic Flow Meter

Norske Sivilingeniørers Forening
Rogaland Distriktshøgskole

MEASUREMENTS OF GAS AND LIQUIDS

Rogaland Distriktshøgskole, Stavanger
7.- 10. juni 1982

PHYSICAL PROPERTIES OF GASES AND LIQUIDS

Olav Vikane

Rogaland Distriktshøgskole/Rogalandsforskning

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
C1	17.0	40.5
C2	8.7	12.8
C3	6.5	9.9
iC4	1.4	0.6
nC4	3.7	1.6
iC5	1.7	0.9
nC5	2.0	1.1
C6	5.4	1.2
C7+	<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

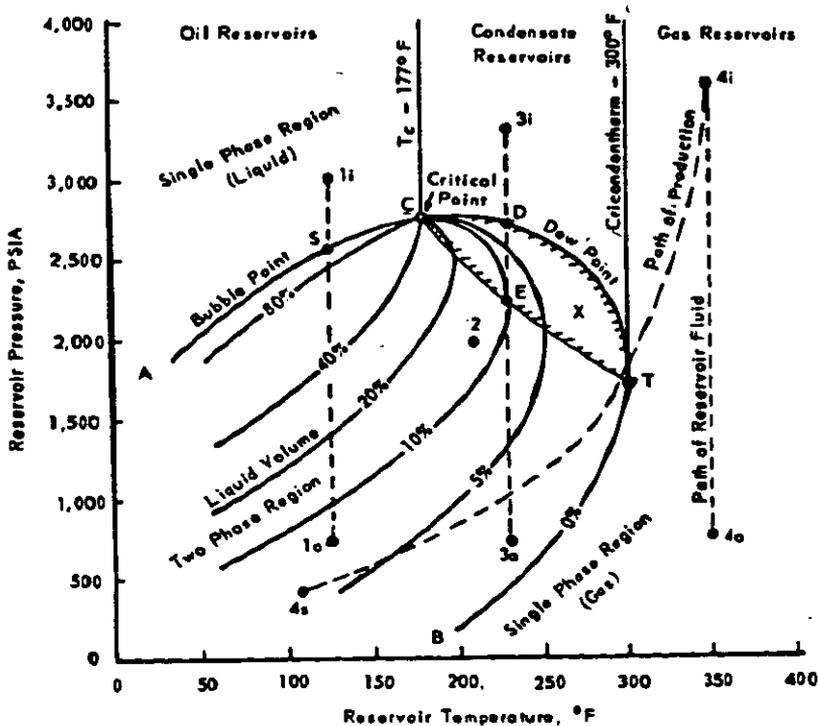


Figure 1. Pressure-temperature phase diagram of a petroleum fluid

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

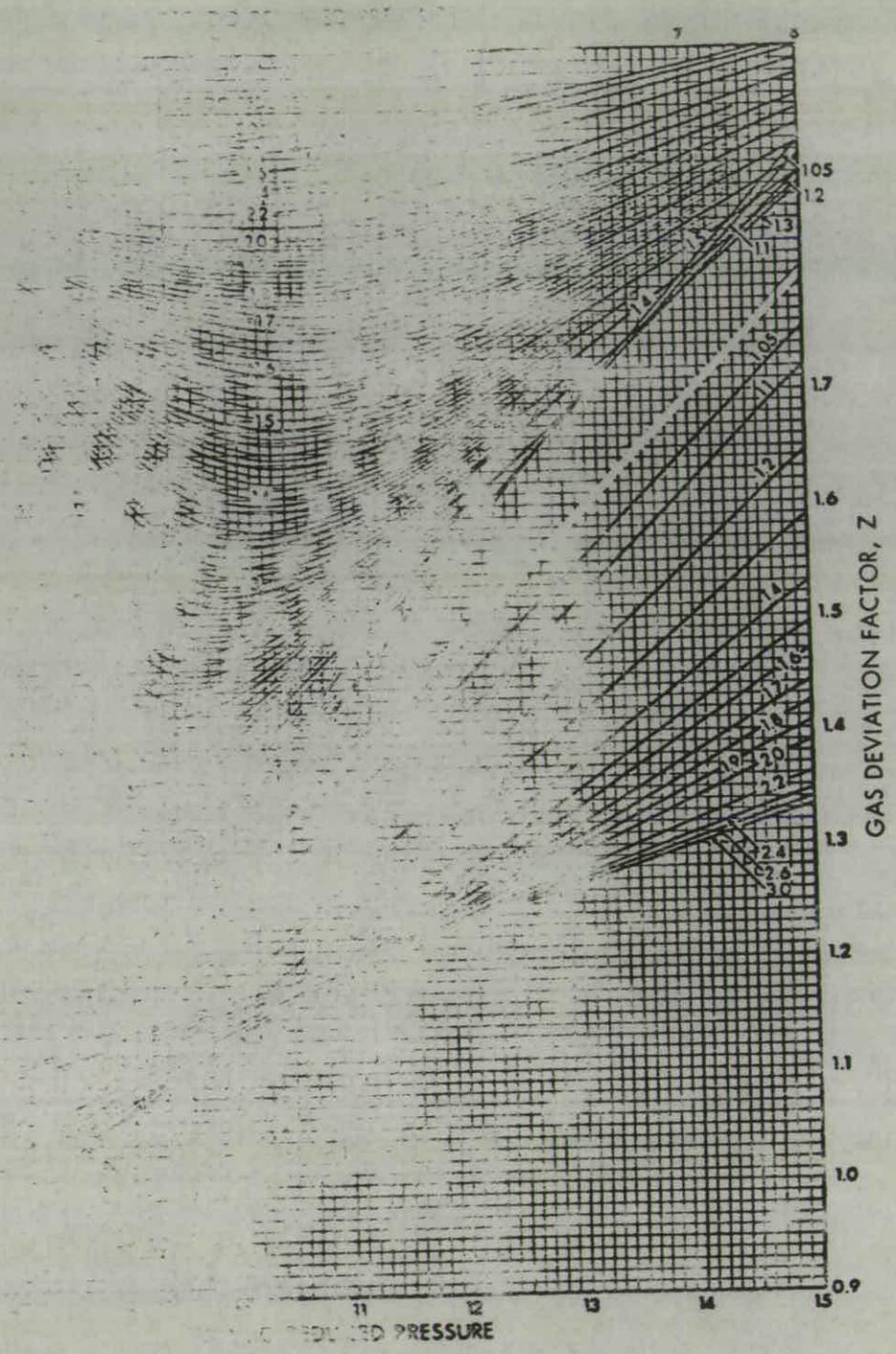
Beginning with his thesis in 1873,⁴ J. D. van der Waals proposed his theorem of corresponding states. Before stating the theorem the following terms will be defined: Reduced temperature, reduced pressure, and reduced volume are the ratios of the actual temperature, pressure, and specific volume to the 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



The gas deviation factor for natural gases (after \bar{z}^E)

whose properties are ⁶:

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

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

Specific gravity, $\bar{g}_e = \sum_i y_i \bar{g}_i$ (E)

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.

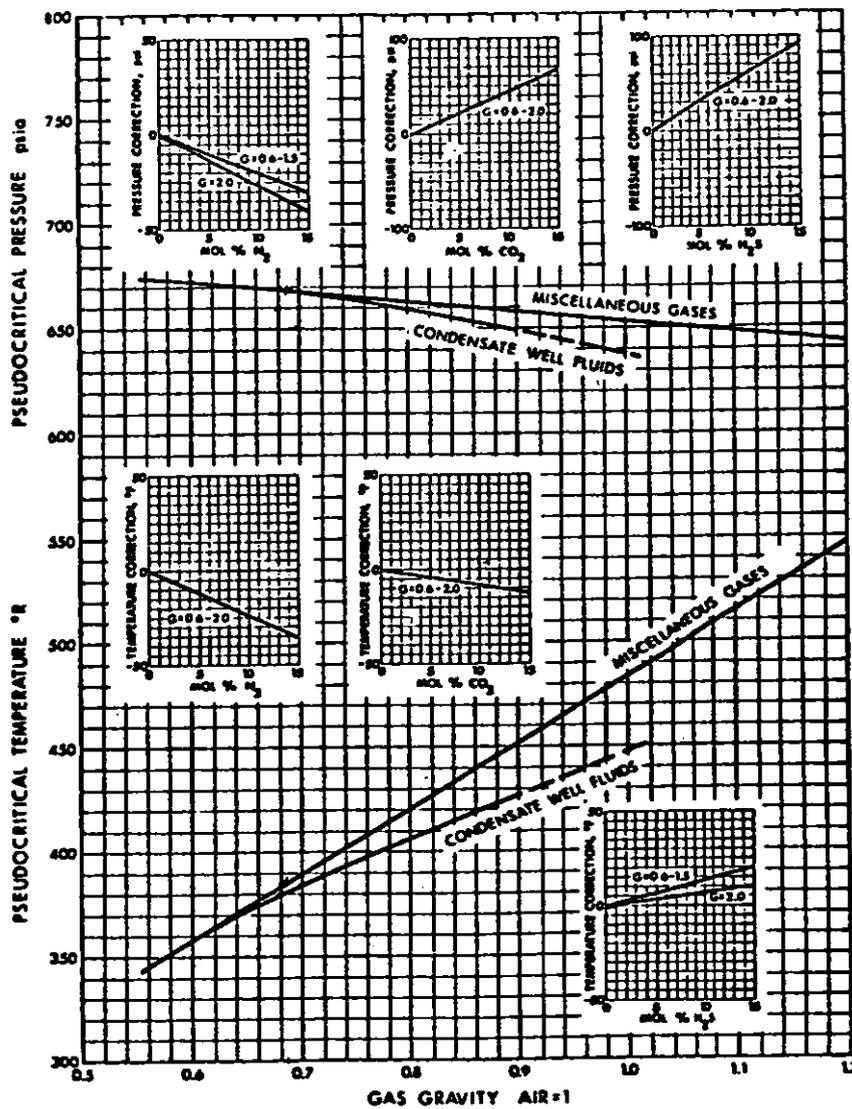


Figure 3. Pseudo critical properties of miscellaneous gases.

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:

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

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

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

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

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.

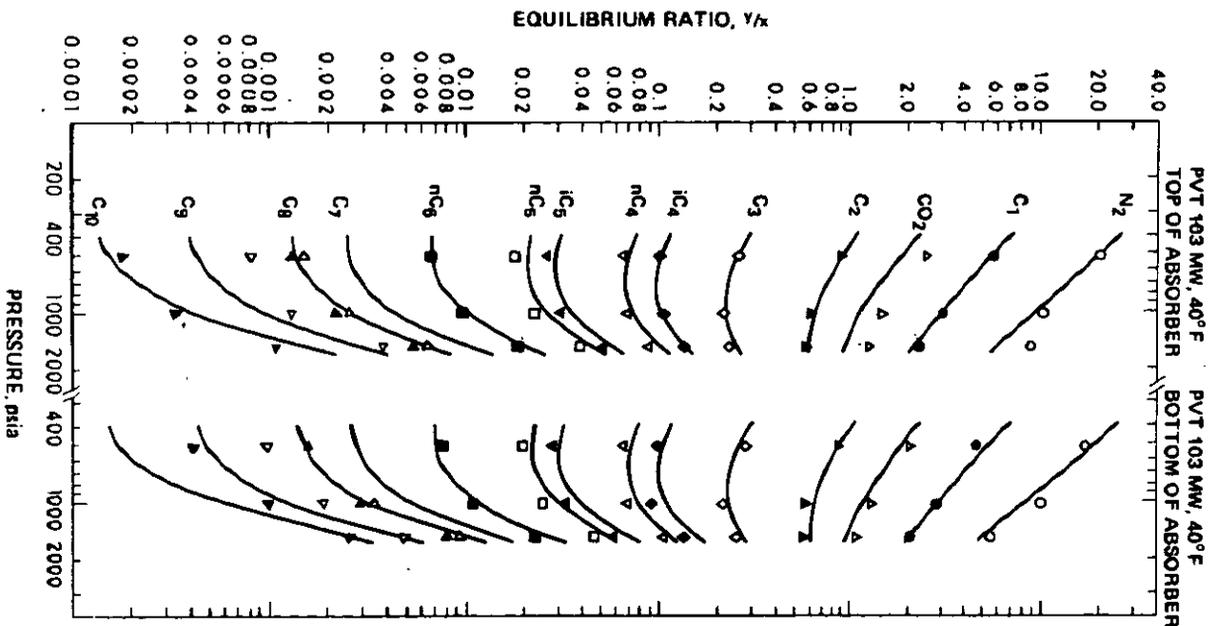
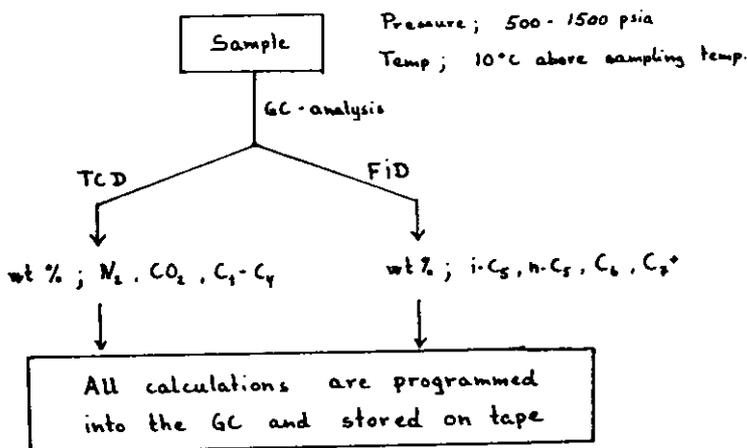


Figure 4. Experimental and calculated equilibrium ratios for 103 Mw absorber oil at 40 °F.

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 plus (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.



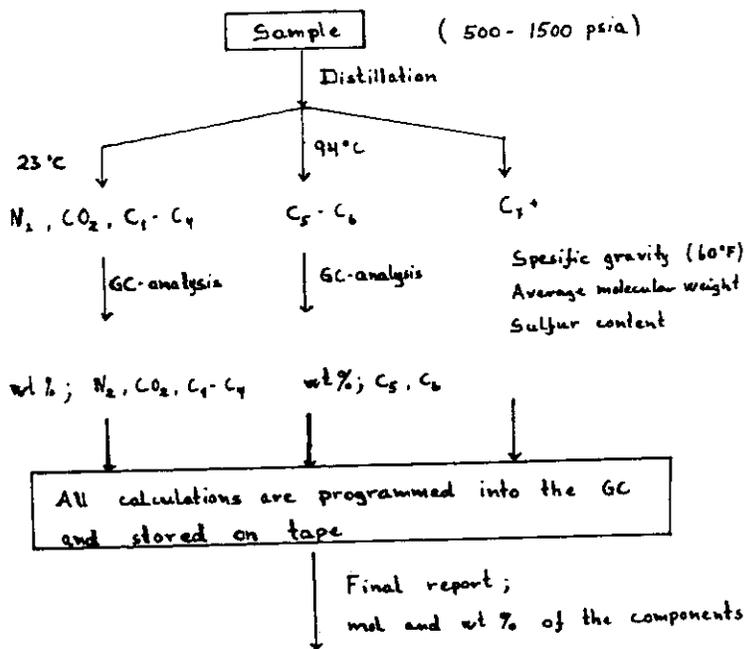
	NAMES	WEIGHT%	MOL%
1	NITROGEN	0.95	0.44
2	CARBONDIOXID	0.73	3.19
3	METHANE	61.65	96.22
4	ETHANE	12.86	9.93
5	PROPANE	9.12	4.32
6	I-BUTANE	1.35	0.42
7	N-BUTANE	2.31	0.51
8	I-PENTANE	1.00	0.25
9	N-PENTANE	1.74	0.35
10	HEPTANES	1.17	0.28
11	HEPTANES PLUS	0.36	0.06
	CRITICAL TEMPERATURE (DEG. C)		-33.9
	CRITICAL PRESSURE (PSIA)		676.2
	REAL GAS SPECIFIC GRAVITY		0.723
	AVERAGE MOLECULAR WEIGHT		20.22
	IDEAL GAS HEATING VALUE (CAL. BTU)		1951.2
	COMPRESSIBILITY FACTOR		0.927

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: 10:57 MAY 26, 1991

	NAMES	WEIGHT%	MOL%
1	NITROGEN	0.01	0.04
2	CARBON DIOXIDE	0.00	0.00
3	METHANE	0.00	20.84
4	ETHANE	1.75	5.97
5	PROPANE	2.10	5.80
6	I-BUTANE	1.79	4.48
7	N-PENTANE	3.11	3.77
8	I-PENTANE	1.50	2.15
9	N-HEPTANE	1.91	2.74
10	HEPTANES	6.16	7.41
11	HEPTANES PLUS	28.00	40.70

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.

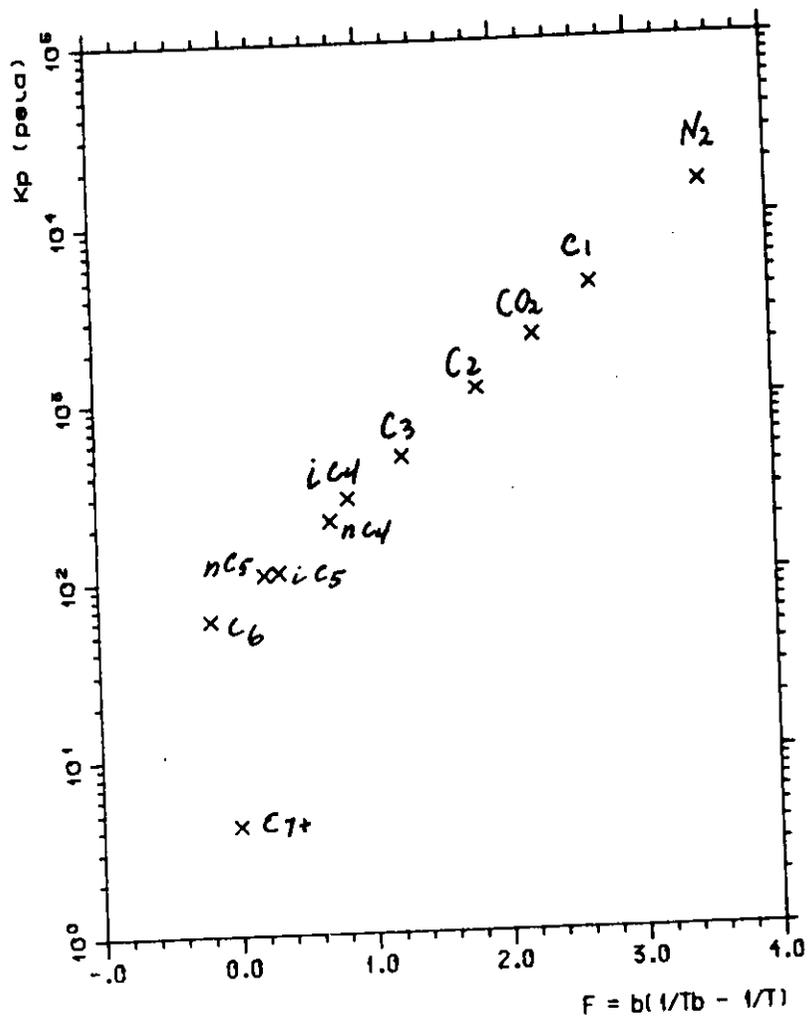


Figure 7. $\log K_p$ plotted as a function of F .

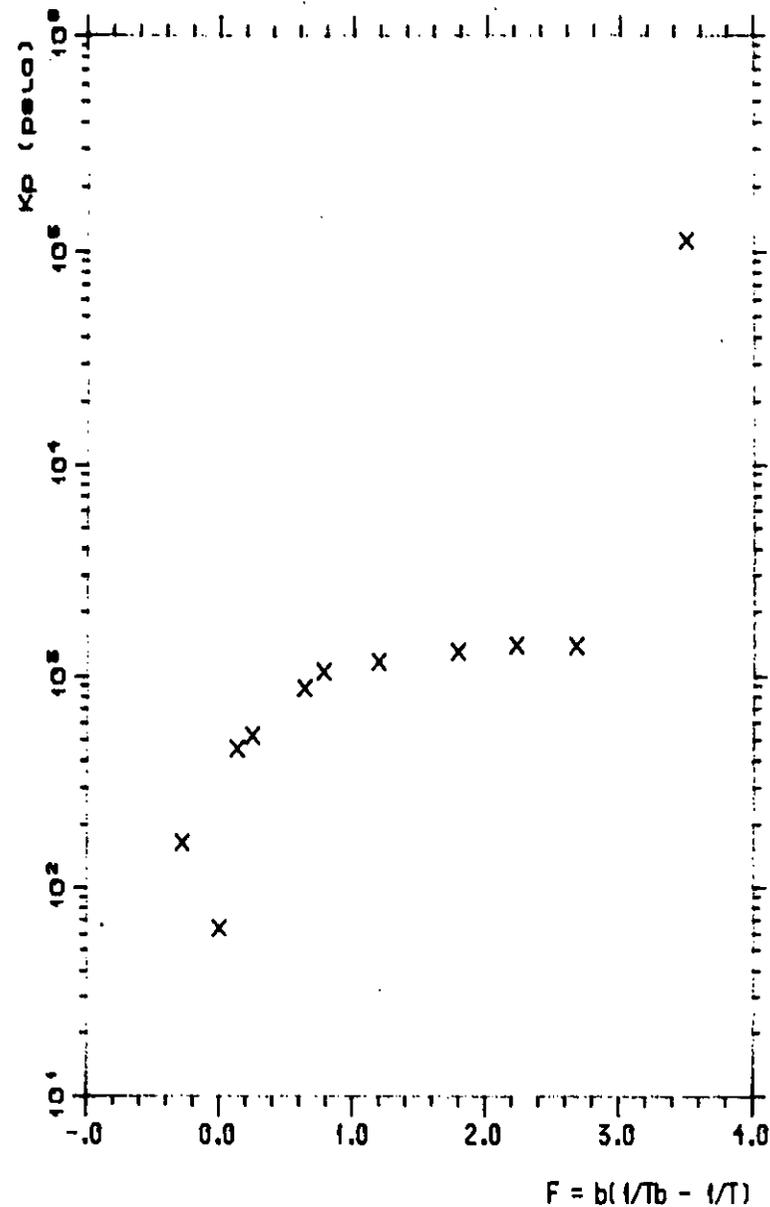


Figure 8. $\log K_p$ plotted as a function of F . The curvature of the plot indicates erroneous results.

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.

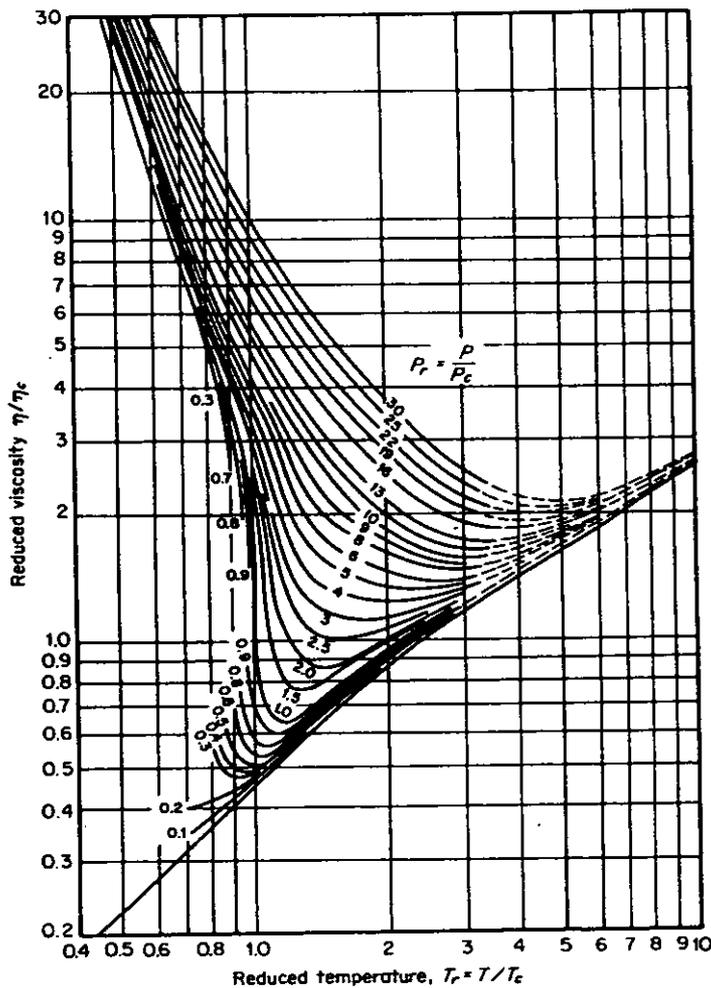


Figure 9. Generalized reduced gas viscosities.

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.

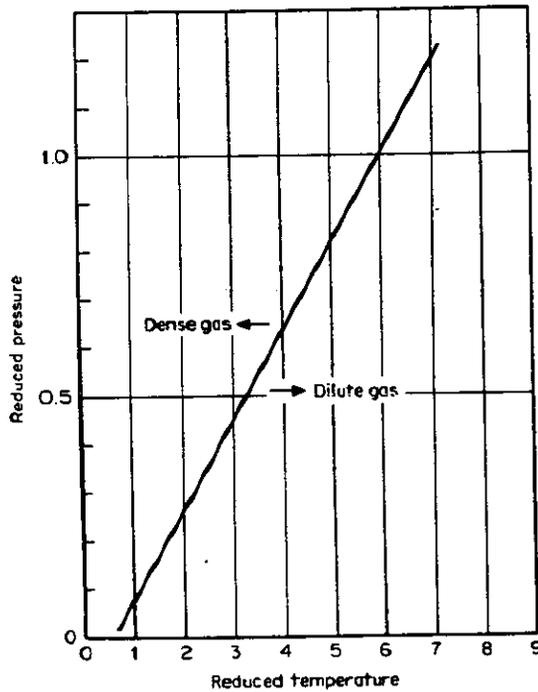


Figure 10. Ranges of reduced pressure and temperature for separating dilute and dense gases.

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.

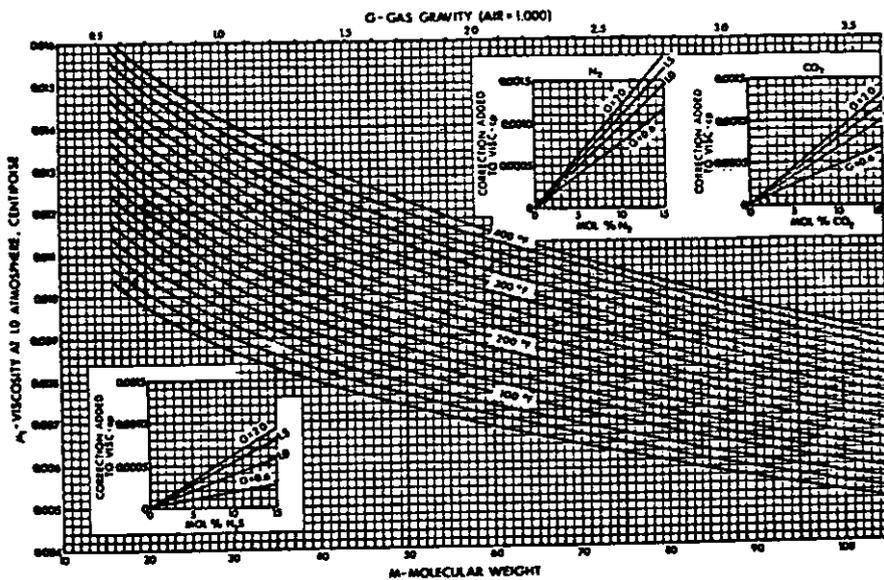


Figure 11. Viscosity of paraffin hydrocarbon gases at 1 atm.

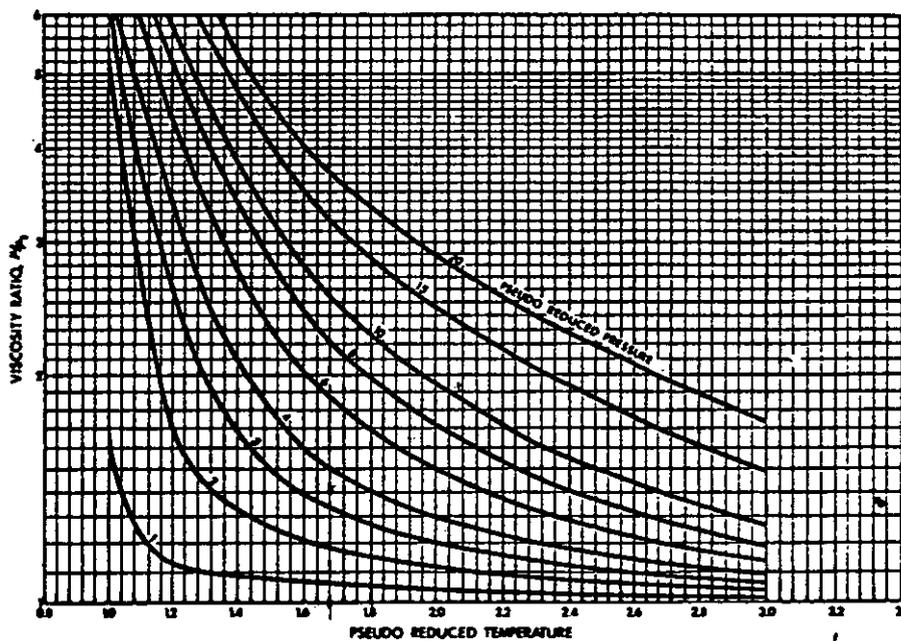


Figure 12. Viscosity ratio vs. pseudoreduced temperature

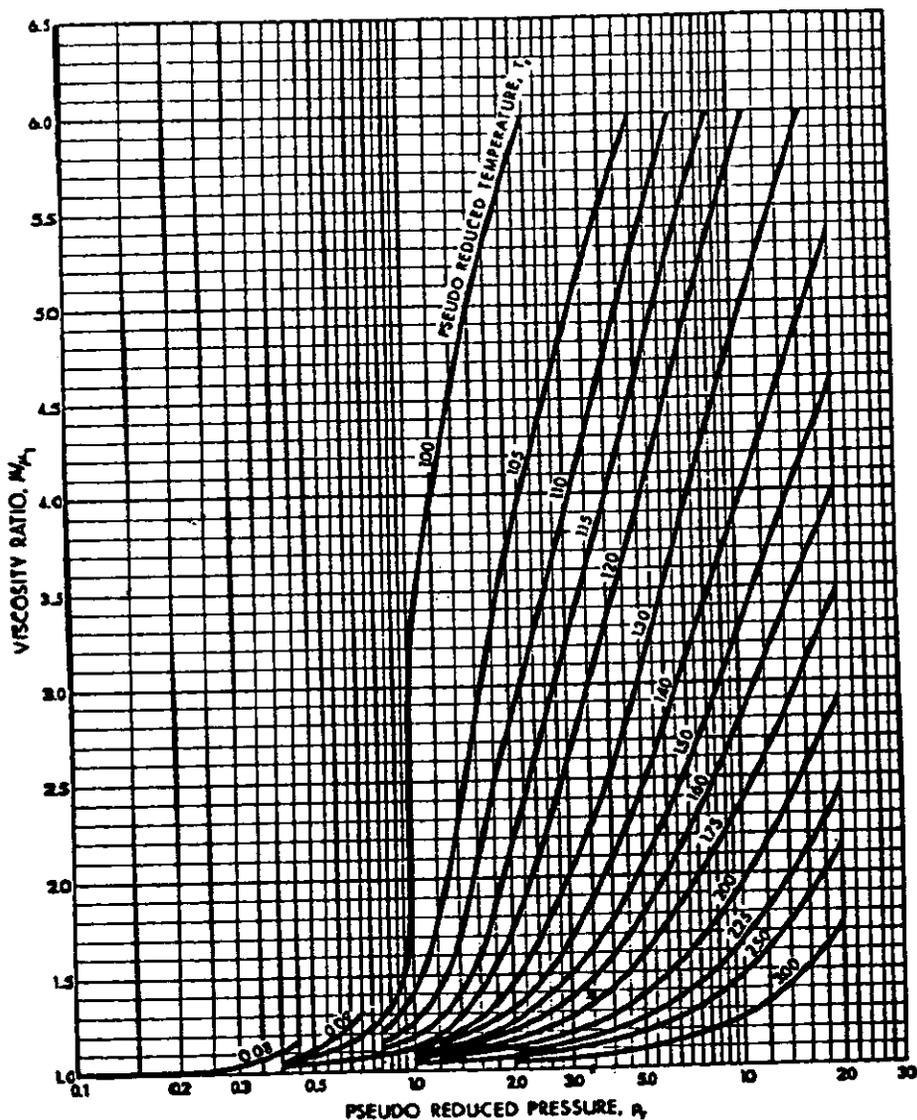


Figure 13. Viscosity ratio vs. pseudoreduced pressure.

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.

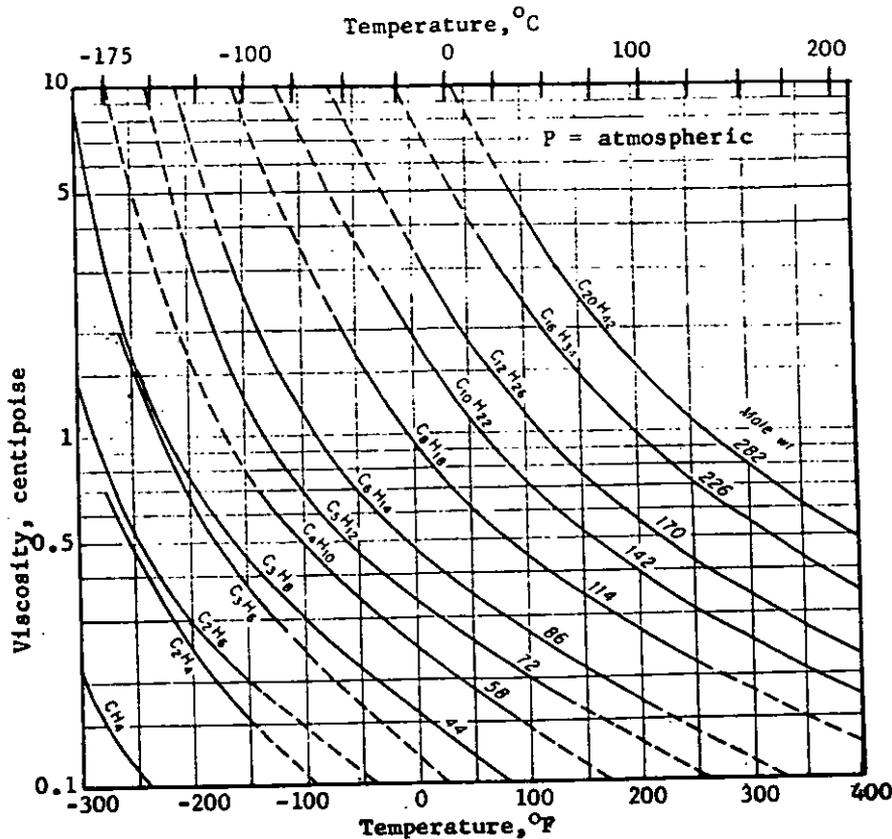


Figure 14. Viscosity of pure hydrocarbon liquids.

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:

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

μ 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.

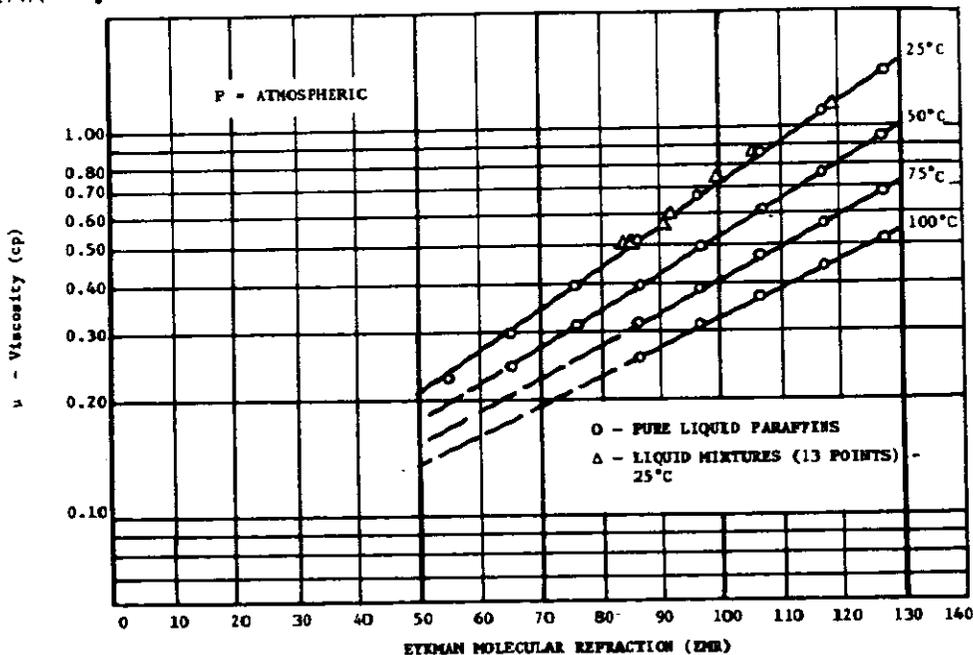


Figure 15. Viscosity vs. Eykman Molecular Refraction (EMR).

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:

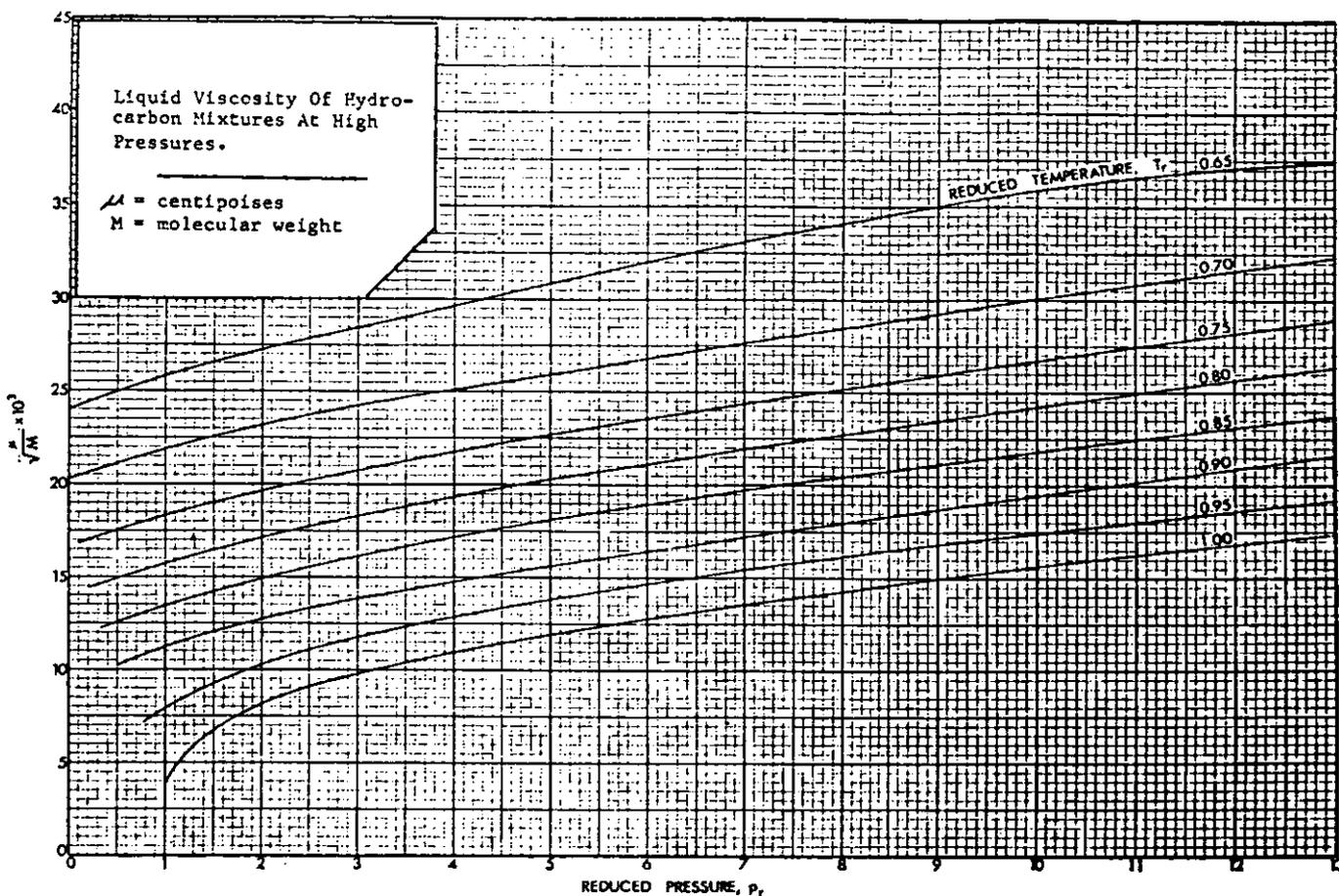


Figure 16. Liquid viscosity of hydrocarbon mixtures at high pressures.

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

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, ft³/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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MEASUREMENT OF GAS AND LIQUIDS

June 7-10, 1982

Rogaland Regional College

Stavanger

Operation, calibration and maintenance of
accuracy of orifice metering systems for
gas measurement - experience from the St. Fergus
gas terminal

Lecturer: Dr. P.L. Wilcox
Metering Engineer,
Total Oil Marine, Aberdeen,
Scotland

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**OPERATION, CALIBRATION AND MAINTENANCE OF ACCURACY OF ORIFICE METERING
SYSTEMS FOR GAS MEASUREMENT - EXPERIENCE FROM THE ST FERGIUS
GAS TERMINAL**

Dr P L WILCOX

Metering Engineer, Total Oil Marine, Aberdeen, Scotland

SUMMARY

The Frigg gas field was discovered in July 1971 and gas began flowing to St Fergus shore terminal in September 1977. Approximately one third of Britain's gas is supplied by Frigg, which underlines the importance of the gas custody transfer metering system at St Fergus. The operation and methods of calibrating the metering system is described in this paper, together with the metering experience gained over the past five years.

1. INTRODUCTION

Gas from the Frigg field is transported 365 km to a shore terminal through two 0.81 m diameter pipelines. The shore terminal is at St Fergus, Scotland, and it is here that the gas is treated and metered before hand over to the British Gas Corporation.

At an intermediate platform, MCP01, situated 170 km from shore, the Frigg pipeline takes in additional gas from Occidental's Piper field and will soon - end of 1982 - also take gas from Texaco's Tartan field.

A Norwegian association and a UK association share the field ownership in the approximate ratio 60 : 40 respectively, since the Frigg field was found to straddle the dividing line of the Anglo-Norwegian shelf (Blocks UK 10/1 and NW 25/1). The members of the Norwegian association are Elf Aquitaine Norge (42%), Norsk Hydro (33%), Total Marine Norsk (20%) and Statoil (5%). The members of the UK association are Elf UK (40%), Total Oil Marine (33%), Aquitaine UK (22%) and BP (1%). Elf are the operators on the field and Total Oil Marine are operators of the pipeline, the intermediate platform and the shore terminal.

The present maximum capacity of the two lines is approximately 67×10^6 sm^3/day , with an additional average daily quantity of 200 tonnes of condensate extracted at St Fergus. In late 1983 compression will be installed on MCP01 for one gas line, and then the maximum capacity will be boosted to approximately 75×10^6 sm^3/day .

Metering of the gas is carried out at three locations : Frigg for the Frigg field gas production (metering done by Elf); MCP01 for Piper gas and later for Tartan gas (metering done by Total Oil Marine); St Fergus gas terminal for the combined gas handed over to BGC. In this paper we shall be concentrating on the gas custody transfer metering system at St Fergus which is the responsibility of Total Oil Marine.

The accuracy of the metering is of great importance, because flow and energy measurements form the basis for contracts. Such contracts include those between buyer and seller of the gas, between partners of the field, between companies who enter into agreements to share facilities such as

the pipeline. In addition, the British and Norwegian governments are interested in the amounts of gas metered for their Royalty payments. Finally, measurements are required for the proper control of production, transportation, or treatment, where in these cases the measurements are process parameters producing reactions from operators or on control loops.

The metering at St Fergus gas terminal is open for inspection at any time by UK Department of Energy and Norwegian Petroleum Directorate inspectors, and Total Oil Marine is pleased that it has built up a good relationship with these inspectors by maintaining the accuracy of the metering system to the best of its ability.

2. ST FERGUS GAS CUSTODY TRANSFER METERING SYSTEM

At St Fergus the gas is treated to meet the specification set by BGC. This treatment involves : separation of entrained liquid; chilling and then removing the resulting condensate; reheating; and then metering.

The metering system was designed as two identical metering stations to ISO R541 recommendations for measurement of flow using orifice plates with flange taps. Each metering station comprises six 61 cm diameter metering tubes with normally four tubes on line, one tube on standby, and one tube being off line for calibration purposes. Every week one tube per station is calibrated, which means that all tubes are checked once every six weeks.

The volume flow rate, Q_v , in standard m^3 /hour is calculated according to ISO R541 using :

$$Q_v = \frac{0.039986}{1.22495} \cdot \frac{\alpha \epsilon d^2 \sqrt{\Delta p \rho}}{s} \quad (1)$$

where α = flow coefficient (ISO R541, section 6.6.1.1)

ϵ = expansibility factor (ISO R541, section 6.6.1.3)

d = diameter of orifice, mm

Δp = differential pressure across orifice plate, mbar

ρ = operating density upstream of orifice plate, kg/m^3

s = relative density

Hence, the volume flowed is obtained by multiplying Q_v by the time in hours.

At the present time, since we are using analogue flow computers, the constants and are only calculated at one point and so the equation for volume flow rate actually used becomes :

$$Q_v = \frac{K \sqrt{\Delta p \cdot \rho}}{s} \quad (2)$$

where K is a constant which is only changed when the orifice diameter is changed. For each metering tube, differential pressure is measured by two transmitters. The low range transmitter and high range transmitter cover 0 to 62.5 mbar, and 62.5 mbar to 250 mbar respectively.

The corresponding energy flow rate, E , in megajoules/hour (MJ/hour) is obtained from :

$$E = Q_v \times CV, \quad (3)$$

where CV is the calorific value of the gas in megajoules/sm^3 . Then the energy supplied by the gas is obtained by multiplying E by the time in hours.

3. CALIBRATION CHECKS MADE EVERY SIX WEEKS

The following checks are carried out at six-weekly intervals on an individual metering tube and associated instrumentation :

3.1 Complete system "as found" when tube taken off line.

- 3.2 Operating density cell.
- 3.3 High and low differential pressure transmitters.
- 3.4 Flow Computer
- 3.5 Signal conditioners and alarms
- 3.6 Orifice plates
- 3.7 Local recorder
- 3.8 Complete system "as left" at end of calibrations
- 3.9 Operational check out

A complete system check is made when the tube is taken off line to ascertain the "as found" state, prior to commencing re-calibration.

The operating density cells on each tube are calibrated with high purity nitrogen. These density cells are installed in pockets in the metering tubes downstream of the orifice plates. Positioning the cells in the pockets ensures that the gas in the density cell is at the same temperature as the gas in the meter tube, and calibration is carried out "in situ".

The high and low differential pressure transmitters are again calibrated using the high purity nitrogen, at a nominal line pressure of 42 barg. They are calibrated at five points on the instruments' ranges 0, 25%, 50%, 75% and 100% of span on both rising and falling applied inputs.

The flow computer is checked by applying simulated values of differential pressure and density and comparing the observed output with the calculated value.

The signal conditioner from the relative density analyser to the flow computer is checked, together with all the alarms to the metering tube under calibration.

The orifice plate is physically removed from the orifice plate carrier and checked by eye for cleanliness, condition of the orifice edge, and the condition of the "O" ring on the plate.

The local recorder needed for standby metering measures upstream static pressure, differential pressure and temperature. The static pressure and differential pressure pens are calibrated at 0, 25%, 50% and 75% of span (75 bar) and 0, 12.5%, 25%, 37.5% and 50% of span (250 mbar). Temperature is only checked at one point, the gas temperature.

With the complete system having been checked and recalibrated where necessary, a check is made on the complete system "as left" at the end of calibration.

This is followed by an operational check out which compares the measured time for metering 20,000 sm³ of gas and hence the measured flow rate, against a calculated flow rate.

4. CALIBRATION CHECKS MADE EVERY WEEK

For each metering station there are two relative density meters and two calorimeters. At any one time there are always one relative density analyser and one calorimeter on line, whilst the second instruments are on stand-by or are available for calibration. The calibration of the calorimeters is checked at one point, weekly, with high purity methane. The calibration of the relative density analysers is also checked weekly with the high purity methane and a mixture of methane and nitrogen.

5. MONTHLY CHECKS COMMON TO ALL TUBES

Once monthly on each station a check is made by comparing for one hour the flow obtained from the sum of the individual tube totalisers with the volume obtained on the station totaliser. In addition, a transmission check is made on the calorific value output at source and the actual value being inputted into the computer.

6. METERING IMPROVEMENTS OVER THE PAST 3 YEARS

The main changes made to the metering in the past 3 years are :

6.1 Change of high differential pressure transmitters

6.2 Change from operating density analysers out of the metering tubes to density analysers in tube pockets.

The original high differential pressure transmitters manufactured by Westinghouse kept drifting, and after evaluating 3 other types of transmitter, Rosemount transmitters were chosen as replacements. The Rosemount transmitters have been found to be extremely reliable, with little or no drifting taking place over the six week calibration interval.

The original operating density analysers were outside of the metering tubes and due to temperature discrepancies occurring between the gas inside the metering tube, and the gas inside the analyser, these discrepancies caused errors in the measured value of operating density. To overcome the problem, two possible solutions were investigated. One solution was to place the analyser in a thermostatically controlled enclosure which maintained the enclosure at the same temperature as the gas in the metering tubes. The second solution was to change to analysers in pockets in the metering tubes. These pockets were placed downstream of the orifice plates. After a detailed comparison of the two possible solutions analysers in pockets were chosen as the preferred solution, and all the metering tubes were converted to take the new density analysers.

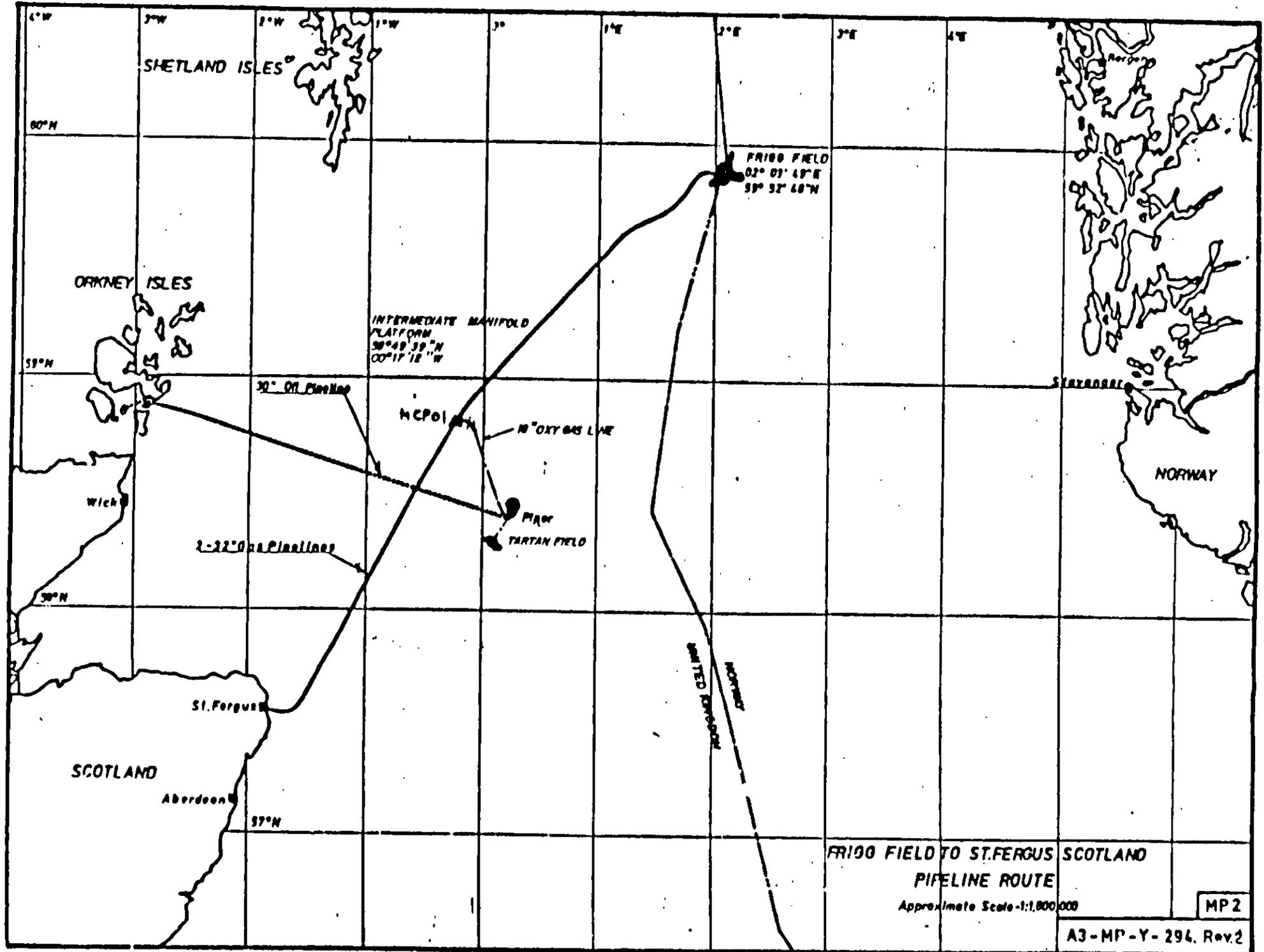
7. FUTURE PLANS

At this moment, the performance of Solartron relative density analysers are being evaluated with the intention of using them in place of the ageing Spanner-Pollux analysers.

To increase the present capacity of the metering, Total Oil Marine is in the process of uprating the differential pressure transmitters to 500 mbar. This uprating is expected to be completed by Autumn 1982.

It is also proposed to change the metering standard from ISO R541 to ISO 5167 at the same time as the change is made on MCP01 and Frigg. This change will take place towards the end of 1982.

During the summer of 1983 it is proposed to change from the existing analogue flow computers to digital flow computers. The main reasons for the change are the increased computation accuracy to be gained, and increasing difficulty in obtaining spares for, and in maintaining, analogue flow computers.



PAUL WIRECOX

FRIGG FIELD TO ST. FERGUS SCOTLAND
 PIPELINE ROUTE
 Approximate Scale - 1:1,000,000

MP 2

A3-MP-Y-294, Rev.2

PARTNERS IN FRIGG

60%

40%

<u>%</u>	<u>NORWEGIAN ASSOC.</u>
42	ELF AQHI. NORGE
33	NORSK HYDRO
5	STATOIL
20	TOTAL MARINE NORSE

<u>U.K. ASSOC.</u>	<u>%</u>
AQUITAINE U.K.	22
ELF U.K.	44
TOTAL OIL MARINE	33
B.P.	1

CONTRACTUAL YEAR 1980-81

AVERAGE AMOUNT OF GAS $\approx 50 \times 10^6 \text{ m}^3/\text{day}$,

AND ≈ 200 TONS OF CONDENSATE / day.

MAXIMUM CAPACITY OF LINE

NOW (2 LINES) $\approx 67 \times 10^6 \text{ m}^3/\text{day}$.

LATER -

CAPACITY OF ONE PIPELINE TO BE BOOSTED TO $42 \times 10^6 \text{ m}^3/\text{day}$ BY COMPRESSION ON MCP01

PARTIES INTERESTED IN THE METERING

FIELD DEVELOPERS

BRITISH AND NORWEGIAN GOVTS.

B.G.C.

OCCIDENTAL + TEXACO

WHY ?

BECAUSE ALL CONTRACTS REQUIRE TO KNOW
GAS QUANTITIES INVOLVED

- BUYER AND SELLER
- ROYALTIES
- OPERATING COSTS
- TRANSPORTATION COSTS

Additionally : REQUIRE METERING FOR THE PROPER
CONTROL OF : PRODUCTION
TRANSPORTATION
TREATMENT.

METERING AT 3 LOCATIONS

1. FRIGG - FRIGG GAS PRODUCTION

METERED BY ELF AQUITAINE NORCE, THE FIELD OPERATOR

2. MCPOI - PIPER (OXY) + TANTAN (TEXACO) GAS

METERED BY TOTAL OIL MARINE, THE PIPELINE OPERATOR

3. ST. FERGUS - GAS CUSTODY TRANSFER SYSTEM

METERED BY TOTAL OIL MARINE, THE TERMINAL OPERATOR

THE CONDENSATE IS ALSO MEASURED HERE.

ALLOCATION

GAS METERED AT ST FERGUS, MCPDI, AND FRIGG IS (OR WILL BE) A MIXTURE OF FLOWS FROM THE VARIOUS FIELDS

OWNERSHIP : GAS ANALYSIS AT SOURCE AND DELIVERY POINT.

COMPUTE SHARE OUT ACCORDING TO MATHEMATICAL RULES AGREED BETWEEN EACH SHIPPER OF GAS ENTERING FRIGG SYSTEM.

B.G.C. MODULATES NOMINATION OF GAS

ACCORDING TO : { WEATHER
HOLIDAYS
SUPPLY FAILURES ELSEWHERE

FRIGG FIELD TAKES CARE OF THE VARIATIONS "PACKING" THE LINE IF NEEDED.

THUS "BANKING" THE GAS FOR TIMES OF PEAK DEMAND, OR SHORTAGE OF PRODUCTION.

ST FERGUS GAS CUSTODY . METERING SYSTEM

THE GAS IS FIRST TREATED TO MEET B.G.C. 'S' SPECIFICATION :

- SEPARATION OF ENTRAINED LIQUID
- CHILLING AND REMOVAL OF RESULTING CONDENSATE
- REHEATING
- METERING

2 IDENTICAL METERING STATIONS TO ISO 1541, USING ORIFICE PLATES WITH FLANGE TAPS

IN AUTUMN 1982 INTEND CHANGING TO ISO 5167 .

EACH METERING STATION COMPRISES 6 METERING TUBES OF 2 FT. (0.61m) O.D.

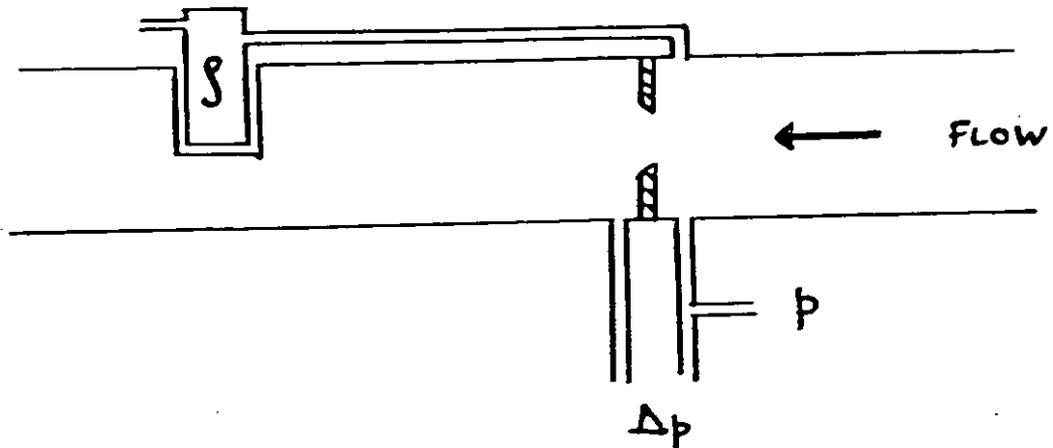
GENERALLY HAVE :

- 4 TUBES IN OPERATION
- 1 TUBE ON STANDBY
- 1 TUBE DEPRESSURISED FOR CHECKING AND CALIBRATING .

ONE TUBE CHECKED PER WEEK \Rightarrow ALL TUBES ARE CHECKED ONCE EVERY SIX WEEKS .

TWO METERING TECHNICIANS TO EACH STATION

BASIC METERING SYSTEM



RD AND CV ARE MEASURED ON A GAS SAMPLE LINE.

$$\text{VOLUME FLOWED (sm}^3\text{)} = \frac{0.039986}{1.22495} \cdot \frac{\alpha \epsilon d^2 \sqrt{\Delta p \cdot \rho}}{S} \times \text{time}$$

- WHERE
- α = FLOW COEFFT. (RS41, section 6.6.1.1)
 - ϵ = EXPANSIBILITY FACTOR (RS41, 6.6.1.3)
 - d = ORIFICE DIAMETER, mm.
 - ρ = OPERATING DENSITY, Kg/m^3
 - Δp = DIFFERENTIAL PRESSURE, mbar
 - S = RELATIVE DENSITY.

SINCE AT PRESENT USING ANALOGUE FLOW COMPUTERS, CONSTANTS α AND ϵ ARE ONLY CALCULATED AT ONE POINT.

THEN :

$$\text{VOLUME FLOWED} = \frac{K \sqrt{\Delta p \cdot \rho}}{S} \times \text{time} \quad (\text{sm}^3)$$

HERE K IS A CONSTANT WHICH IS ONLY CHANGED WHEN THE ORIFICE DIAMETER IS CHANGED.

CORRESPONDING ENERGY FLOWED :

$$E = \frac{K \sqrt{\Delta p \cdot \rho}}{S} \times \text{time} \times CV \quad (\text{MJ})$$

HERE CV IS THE GROSS CALORIFIC VALUE IN MJ/sm^3 .

CALIBRATION CHECKS EVERY SIX WEEKS

1. COMPLETE SYSTEM "AS FOUND" WHEN TUBE TAKEN OFF LINE .
2. OPERATING DENSITY CELL
3. HIGH AND LOW D.P. TRANSMITTERS
4. FLOW COMPUTER
5. SIGNAL CONDITIONER AND ALARMS
6. ORIFICE PLATES
7. LOCAL RECORDER
8. COMPLETE SYSTEM "AS LEFT" AT END OF CALIBRATION
9. OPERATIONAL CHECK OUT

1. COMPLETE SYSTEM "AS FOUND"

8. COMPLETE SYSTEM "AS LEFT"

A SIMULATED RANGE OF PROCESS INPUTS OF OPERATING DENSITY AND DIFFERENTIAL PRESSURE ARE APPLIED TO THE FLOW COMPUTER, USING HIGH PURITY NITROGEN.

THE OBSERVED OUTPUT IS COMPARED WITH THE CALCULATED OUTPUT OBTAINED FROM A COMPUTER PROGRAMME CALLED METCAL

DURING THESE TESTS THE RELATIVE DENSITY IS MEASURED ON THE GAS SAMPLE LINE.

RANGE OF APPLIED INPUTS :

<u>DIFF. PRESS.</u>	<u>OPER. DENSITY</u>
100% HI - 250 mbar	$\approx 46 \text{ Kg/m}^3$
75% HI - 187.5 mbar	$\approx 46 \text{ Kg/m}^3$
75% NI - 187.5 mbar	$\approx 33.8 \text{ Kg/m}^3$
50% LO - 31.25 mbar	$\approx 25 \text{ Kg/m}^3$

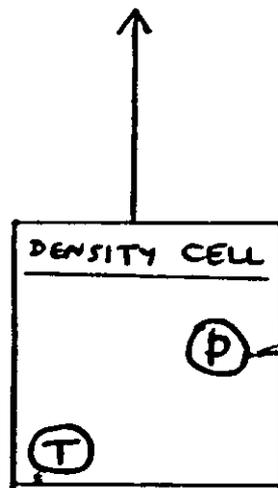
2. OPERATING DENSITY CELL

CALIBRATION TEST CERTIFICATE ISSUED BY FACTORY USED AS A BASIS FOR THE VERIFICATION TEST ON THE TRANSDUCER.

OBSERVED FREQUENCY OUTPUT
— SOLARTRON 2602 TIMER/COUNTER

ATMOS. PRESSURE

— FORTIN BAROMETER



GAUGE PRESSURE OF WHITE SPOT N_2 — BUDENBERG DEADWEIGHT TESTER TYPE 10405/246.

TEMPERATURE OF WHITE SPOT N_2

— SOLARTRON TYPE 100 RESISTANCE

THERMOMETER WITH WHEATSTONE BRIDGE

FOR MEASURING RESISTANCE, TYPE LEADS AND

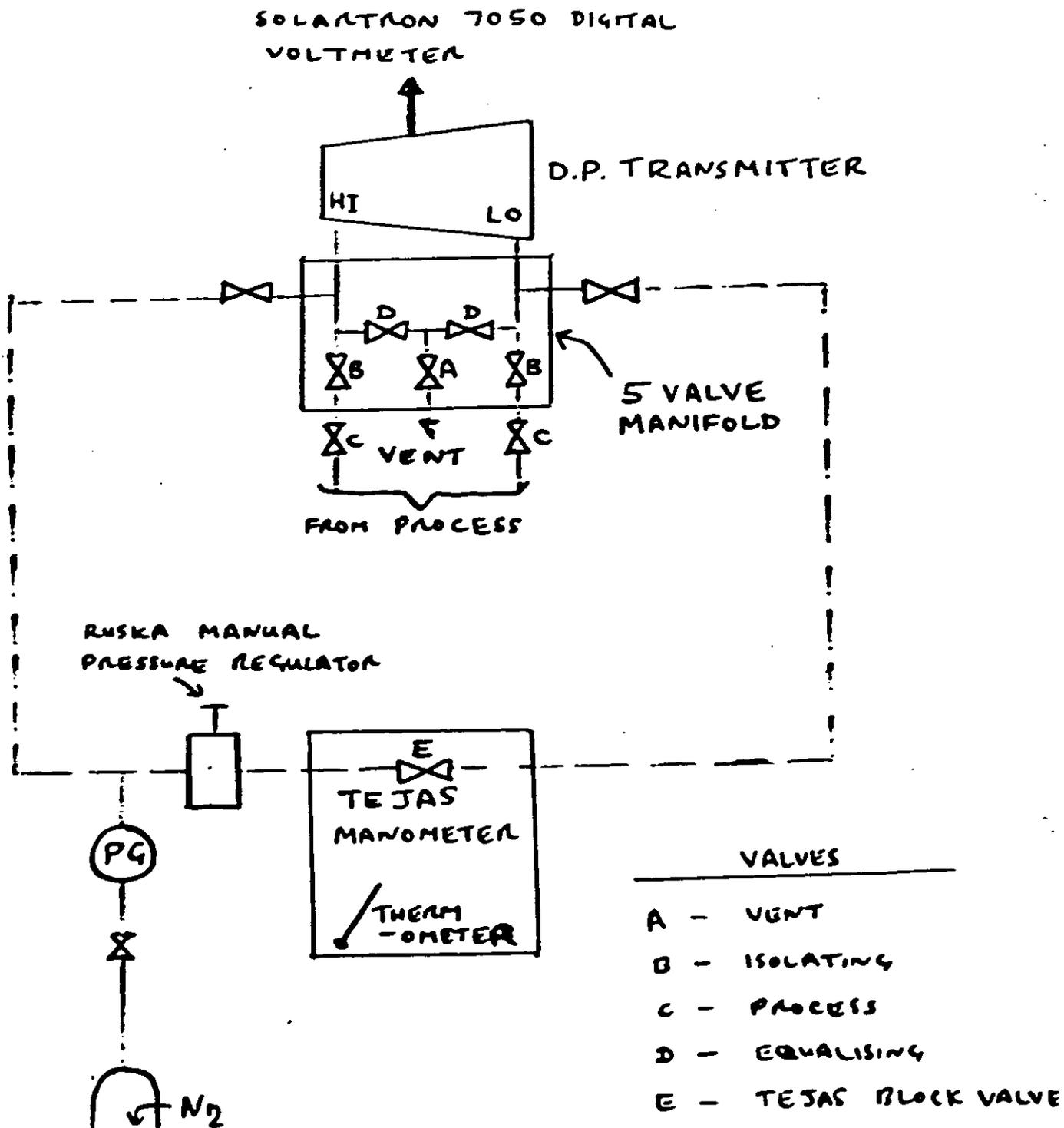
NOTHROP MODEL 19867/13.

USING A COMPUTER PROGRAM WHICH HAS A STORED TABLE OF DENSITY VALUES OF WHITE SPOT N_2 FOR VARYING p AND T , THE CALCULATED N_2 DENSITY (AND HENCE CALCULATED N_2 TRANSDUCER FREQUENCY AND PERIODIC TIME) IS OBTAINED

3. HIGH AND LOW D.P. TRANSMITTERS

CHECK CARRIED OUT AT 42 BAR G. PRESSURE ON BOTH HIGH AND LOW D.P. TRANSMITTERS AT

0, 25, 50, 75, AND 100% OF INSTRUMENT RANGES ON BOTH RISING AND FALLING DIFFERENTIAL PRESSURES THE ACTUAL D.P. IS SET BY THE TEJAS.



4. FLOW COMPUTER

CHECK CONDUCTED IN TWO PARTS

PART ONE

APPLYING SIMULATED INPUTS OF D.P. FOR TWO VALUES OF OPERATING DENSITY WITH RD FIXED.

CHECK OBSERVED FLOW RATE AGAINST CALCULATED FLOW RATE BY COUNTING TIME FOR 100 PULSES.

PART TWO

AT SIMULATED FIXED VALUES OF DP AND ρ , VARYING RELATIVE DENSITY INPUT.

CHECK OBSERVED FLOW RATE AGAINST CALCULATED FLOW RATE BY MEASURING TIME FOR 100 PULSES

NOTE : TOLERANCES FOR THESE TESTS WERE OBTAINED FROM THE ORIGINAL MANUFACTURERS CERTIFICATION TESTS ON THE FLOW COMPUTER.

WEEKLY CALIBRATION CHECKS

1. CALORIMETERS

COLD BALANCE TEST

HIGH PURITY METHANE

Calorimeters completely stripped once in 3 months.

2. RELATIVE DENSITY ANALYSERS.

HIGH PURITY METHANE — 0.555

MIXTURE METHANE / N₂ — 0.584

Spanner - Pollux tendency to drift.

MONTHLY CALIBRATION CHECKS

1. FOR EACH STATION, A ONE HOUR CHECK IS MADE THAT THE FLOW OBTAINED FROM THE SUM OF THE INDIVIDUAL TUBE TOTALISERS

= VOLUME ON STATION TOTALISERS

2. TRANSMISSION CHECK ON THE CV OUTPUT AT SOURCE = ACTUAL VALUE BEING INPUTTED TO COMPUTER

STATISTICAL ANALYSIS OF THE CALIBRATION RECORDS

FROM THE 6-WEEKLY CALIBRATION SHEETS CARRIED OUT A STATISTICAL ANALYSIS IT GIVES A GOOD INDICATION OF THE ACCURACY AND DRIFT OF THE SYSTEM

KNOWING THE APPLIED INPUT VALUE AND MEASURING THE INSTRUMENT VALUE READ IT IS POSSIBLE TO OBTAIN AN OBSERVATION OF THE SPOT DEVIATION OF THE INSTRUMENT.

IT CAN BE EXPRESSED AS :

$$\frac{(\text{READ} - \text{APPLIED})}{\text{APPLIED}} \times 100\%$$

FROM THESE RESULTS CAN THEN FIND :

OBSERVED MEAN DEVIATION - μ

OBSERVED STANDARD DEVIATION - σ

$$\text{UNCERTAINTY} = 2\sigma$$

		HIGH DP CELL	OPERAT ^G DENSITY CELL	RELATIVE DENSITY CELL	FLOW COMPARER	CALOR- IMETER	INDX. TUBE VOL.
AS	μ	-0.061	-0.025	-0.003	+0.016	+0.039	-0.056
FOUND	UNCERT. (2 σ)	1.062	0.500	0.238	0.116	0.338	0.562
AS	μ	-0.057	-0.016	-0.035	+0.015	-0.002	-0.042
LEFT	UNCERT. (2 σ)	0.494	0.146	0.140	+0.012	0.052	0.336

OUT OF DATE.

METERING IMPROVEMENTS OVER THE PAST

3 YEARS

1. CHANGE OF HIGH D.P. TRANSMITTERS

ORIGINAL WESTINGHOUSE DRIFTING CONSIDERABLY

3 ALTERNATIVES TRIED, AND FINALLY

ROSEMOUNTS CHOSEN.

ROSEMOUNTS FOUND TO BE EXTREMELY RELIABLE

- LITTLE OR NO DRIFTING OVER 6 WEEKLY INTERVALS.

2. CHANGE TO DENSITY CELLS IN POCKETS

FOUND THAT WITH THE ORIGINAL OPERATING DENSITY CELLS OUTSIDE THE METERING TUBES, TEMP. VARIATIONS BETWEEN GAS IN ANALYSER AND GAS IN TUBES CAUSE ERROR IN ρ .

SO INVESTIGATED TWO SCHEMES

1. DENSITY CELLS IN POCKETS IN TUBES
2. DENSITY CELLS IN THERMOSTATICALLY CONTROLLED ENCLOSURES OUTSIDE METERING TUBES.

3. CHANGE TO SOLANTION RELATIVE DENSITY ANALYSERS

PRELIMINARY TESTS HAVE SHOWN THAT THE SOLANTION R.D. ANALYSERS ARE MORE STABLE THAN THE EXISTING SPANNA POLUX ANALYSERS.

WE HAVE BOUGHT 4 WITH THE INTENTION OF USING THEM AT ST. FULGUS.

HOWEVER THERE IS A PROBLEM,

CALIBRATION MUST BE DONE AT 7 bar pressure.

COMPRESSIBILITY MUST BE TAKEN INTO ACCOUNT.

WE ARE AT PRESENT LOOKING AT THIS PROBLEM.

4. UPGRADING METERING TO 500 mbar

TO INCREASE METERING CAPACITY

GIVES AN INCREASE THEORETICALLY FROM A

MAXIMUM OF $\approx 9.6 \times 10^6 \text{ sm}^3/\text{day}$

TO $\approx 13.6 \times 10^6 \text{ sm}^3/\text{day}.$

MEASURING IMPROVEMENTS THAT DIDN'T WORK

1. USING ONE DESGRANGES & HUOT D.P. TRANSMITTER TO MONITOR THE PERFORMANCE OF SIX ROSEMOUNT TRANSMITTERS
2. COMPARISON TESTS BETWEEN A D & H AND A ROSEMOUNT D.P. TRANSMITTER

FUTURE PLANS

1. CHANGE TO ISO 5167 - LATE 1982
2. ANALOGUE FLOW COMPUTERS TO DIGITAL FLOW COMPUTERS - SUMMER 1983.

Calibration of Gasmeters

by

H. Bellinga

ERRATA

- page 2 line 6 : lightning must lighting
- page 2 line 10: manufactured gas was
- page 12 line 16: meters of 4000 m³/h
- page 14 line 2: power must be prover

2 juni 1982

Doc. 1038L/0056L

[Main Index](#)

TURBINE METERS FOR GAS.

by

H. Bellinga

1. Introduction

The history of the turbine meter for gases goes back to the beginning of this century. At that time the size of the wet drum type gas meters had increased to impractical dimensions and the search for other flow measuring techniques started.

One of the techniques was to measure the velocity of the medium by means of a propeller.

In these early turbine meters the shaft was always vertical in order to decrease bearing load thus minimising friction and wear (fig 1)

Because of the lack of proper materials and bearings these meters were regarded as unreliable and were only used for industrial measurements. However, at that time it is already recognized that these meters were highly sensitive to changes in the flowrate had a good repeatability and a one to ten rangeability. That was in the mid thirties.

Not until the technology permitted the manufacture of bearings, comparable with the modern bearings the development of the turbine meter as we know it today started.

2. Metrological position.

Prior to dealing with the properties of the turbine meter itself a comparison will be made between the two systems in use for large scale gas flow measurement.

The uncertainty of a well maintained system consisting of an orificeplate, a dp cell, a densitometer and a flow computer, operating with high pressure gas at 70% of its rated capacity is estimated to be 0,7%.

The uncertainty of a system consisting of a turbine meter, calibrated with high pressure gas a densitometer and a flow computer, operating under comparable conditions amounts to 0,45%. Application of the most accurate secondary instrumentation would bring down these figures. For the orifice plate system to 0,6% and for the turbine meter system to 0,35%.

The price for this improvement is the cost of an automatic differential pressure gauge (Dfl 180.000) in the case of the orificeplate.

In the turbine meter system a flow computer with linearisation for the error curve of the turbine meter has to be installed at an extra cost of approximately Dfl 8000.

The rangeability of the orificeplate system is one to three with the penalty of increased uncertainty at low flowrates. For the turbine meter system the rangeability is one to ten with a constant figure for the uncertainty over the whole range. For the more sophisticated systems for the orifice plate the penalty of increased uncertainty is removed while the rangeability of the turbine meter system increases to one to twenty and in some cases even to one to forty.

Although the orifice plate is more simple and more robust than a turbine meter it requires more secondary instrumentation. Especially when high accuracy is required the complexity of this secondary instrumentation is such that the simplicity of the primary element is compensated.

In general the upstream length for a turbine meter is shorter than for an orifice plate, thus offering the opportunity to build more compact installations.

3. Operating principle.

The axial flow turbine meter is a device that measures the velocity of the medium, flowing through a passage of known area. This passage is formed by the hub of the turbine wheel and the wall of the meter housing. The blades on the turbine wheel are positioned under an angle with the direction of the flow.

The ideal turbine meter has infinitely thin helical blades and is not subject to any kind of friction.

Under the conditions mentioned above it would be possible to calculate the relationship between throughput and speed of the rotor from the dimensions of the meter. That relationship is linear can be represented as:

$$\frac{c \cdot n}{\phi} = \frac{f}{\phi} = a$$

Where C is a constant determined by the gear train, n the rotor speed, ϕ the actual flowrate, f the indicated flowrate and a the coefficient depending on meter geometry.

In the real meter however the blades have a certain thickness, friction occurs obstructions in the flowpath are present.

The result of this is that the relationship between rotor speed and throughput is not any more linear.

It can be represented as:

$$\frac{c \cdot n}{\phi} = \frac{f}{\phi} = a_1 + \frac{a_2}{\phi} + \frac{a_3}{\phi^2} \quad (1)$$

The values of a_1 , a_2 , a_3 are determined by curve fitting of the calibration data. The value of a_1 in general differs from the value a.

4. Construction

In general axial flow gas turbine meters consist of a pressure resistant spoolpiece in which the measuring element is inserted, so the static pressure does not influence the dimensions of the measuring element. At the moment the products of most gas turbine meter manufacturers present a strong similarity in aerodynamic design.

(1) Upstream of the measuring element is the flow conditioning zone which consist of a nose cone provided with flow guide vanes at its upstream end. The annular passage between the nose cone and the wall of the spoolpiece in most cases has a constant cross sectional area over a length of about eight times its width.

As the majority of the gas turbine meters are used for custody transfer they have to fulfill a number of requirements. A good survey of these requirements is given in the EEC directives (2). One of these requirements is that the meter should have a mechanical index.

That mechanical index is driven by the turbine wheel by means of a gear train. Part of this gear train is a magnetic coupling that brings the rotation of the drive shaft outside the pressure resistant housing. In addition to the mechanical index most turbine meters have one or two electronic sensors as a second means of read out. In the case of application of a flow computer this is a must. This can be a sensor direct to the turbine wheel or to a disc on the mainshaft or a sensor to a disc on a shaft on the atmospheric side of the magnetic coupling somewhere in the gear train.

A combination of the two offers the possibility of a continuous check of the integrity of the electronic output of the meter.

As the robustness of meters is increasing there is a tendency to rely on a combination of sensor in the gear train solely.

On the gas turbine meters there is one or sometimes two pressure tappings in the same plane, marked Pr, where the relevant pressure must be measured.

The position of this tapping is the result of careful investigation during the design and test phase of the meter. In the case that for the computation of the mass flow rate through the meter the pressure, measured at another tapping would be used an unknown error is introduced.

As the expansion over a turbine meter is low the point where gas temperature is measured is of little importance; some larger meters do have a thermometer pocket. For the smaller meters it is recommended to measure the temperature as close to the outlet flange as possible.

Another important feature of the construction of a turbine meter is the choice of the bearings of the turbine shaft. This is a compromise between low friction and durability.

Initially the turbine meters were used to measure gas at low pressure. That means driving torque is low, so the bearing friction has to be low to achieve good rangeability. However bearing load was also low. The calibration of the meters was performed with air under atmospheric conditions. Later on with the increase of the pressure of the gas to be measured the bearing load increased. The calibration with atmospheric air however remained the criterion for acceptance and judgement of the quality, especially the rangeability, of the meters. At the moment the equilibrium between durability under high pressure and rangeability with atmospheric air has been achieved, provided the gas conditions are not too adverse. An additional benefit of this development is that the rangeability at high pressure is extremely large. Unfortunately only very few users do need that.

With the appearance of the calibration of gas meters under operating conditions the situation is changing. As soon as the high pressure calibration has the same official status as the low pressure air calibration the task for the meter designers will be to find the new equilibrium between durability and rangeability both at high pressure. It will enable the manufacturers to build meters that will be able to withstand even more adverse conditions than the present meters already do.

5. Characteristics

Deviating from the representation given in chapter 3 in practice the behaviour of the gasmeter is represented by the error in the reading at various flowrates. The reason therefore is that the same way of representation is used in the legal prescriptions and certificates.

The error is defined as:

$$Y = \frac{f - \vartheta}{\vartheta} \cdot 100\%$$

where f is the indicated flowrate and ϑ the actual flowrate.

The results of a calibration are presented as a table containing the measured data and the following polynomial which represents the best fit curve of the measured data. (fig. 3).

$$Y = A_0 + A_1 X^m + A_2 X^n + A_3^o$$

Y = error of the meter in %

$$X = \frac{\vartheta}{\vartheta_{\max}}$$

$$m = -0,2$$

$$n = -0,33$$

$$o = -2,0$$

The exponents m , n and o always have the values mentioned above. The values A_0 , A_1 , A_2 , A_3 are characteristic for the individual meter.

Much can and has been said about the behaviour of turbine meters under various operating conditions. In the following those properties that are of importance for the practical application will be discussed.

- 1) Turbine flowmeters always repeat very well, in the order of 0,1 per cent ($2\sigma = 0,1\%$) or better as long as the properties of the medium do not change too much. The lowest flow rate for which this statement is valid depends on the static pressure and decreases with increasing pressure.
- 2) For the majority of the modern meters it is possible that the calibration curve with atmospheric air and with gas at elevated pressure are both within the legal tolerance limits (fig. 4). As the shift of the error curve under influence of the density is individual a calibration under conditions approaching the operating conditions is

necessary to get the full profit from the good repeatability. In this case it may be necessary to adjust the meter in such a way that the air curve lays outside the tolerance limits. (fig. 5).

- 3) With increasing density of the gas the linearity at low flowrates improves resulting in an increased rangeability (3). For larger meters a rangeability of one to fourty is no exception. (fig. 6). For the smaller meters this figure is one to thirty. (fig. 7)
- 4) When operated with gas at high pressure (40 bar and higher) bearing friction only influences the error curve at very low flowrates.
- 6) Installation and operating conditions.

The general statement can be made that disturbances in the gasflow upstream of a turbine meter should be avoided. Anything other than a straight pipe must regarded as a source for disturbances. Disturbing elements can be divided into two groupes. High level disturbances and low level disturbances. The first catagory is generated by throttling elements with over critical expansion. The second catagory is generated by piping configurations as bends, Tee's and headers in which the velocities are of the same order as the entrance velocity of the meter.

In the case that high level disturbances are present the application of a modified Sprengle straightening (fig. 8) vane installed as indicated in fig 9 will make sure that the error of the meter will not deviate by more than 0,2% from the error in an undisturbed flow (4). The high pressure drop of this type of flow straightener can be accepted because of the presence of the throting element with the overcritical expansion.

When low level disturbances occur the application of a flow straightener with a lower pressure drop will be sufficient. The tube bundle straightening vane is the most wellknown type (fig. 10). To give an impression of the influence of one of the most severe low level disturbances, two bends not in one plane, the following. This disturbance ten pipe diameters upstream of a turbine meter causes a shift in error between 0,1% and 0,8%. These figures have been measured in a test with 5 meters of different construction.

- 9 -

The more vanes the internal flowstraightener of the meters has, the lower is the influence of the disturbance.

More difficult to quantify is the influence of flow pulsations so only some qualitative remarks will be made.

Pulsating flow will not influence the indication of the meter as long as its frequency is low in respect to the time constant of the turbine wheel. When the frequency is such that there is an influence, that influence is smaller as the amplitude is smaller with respect to the average flow rate. More detailed and quantified information on this subject is given by Dijstelbergen. (5).

Recent developments

One of the most recent innovations in turbine meter construction is the auto adjust meter manufactured by Rockwell. In this meter downstream of the main turbine wheel a second free running turbine wheel with only a very small blade angle is installed. Because of the friction of the meter the flow downstream of the main rotor is slightly rotating, proportional to the friction of the main rotor. This rotation will influence the rotation of the second turbine wheel (sensor rotor) proportional to the slip of the main rotor. So the slip of the main rotor is measured and is corrected for electronically.

In this meter errors caused by changes in the mechanical friction are corrected.

Changes of the error caused by an increased inlet velocity at the turbine wheel because of blockage of the flowpath will not or only partly be detected. A more detailed description is presented by the manufacturer (6).

Another design which mechanically very much resembles the Auto Adjust meter is a prototype of a direct massflow meter.

It consists of a normal turbine wheel which is coupled to a (constant torque brake.) Downstream of the turbine wheel is a free running sensor wheel with straight vanes that measures the swirl of the gasflow downstream of the turbine wheel. Both rotors have an electronic readout. It can be derived that the mass flow through the meter is:

$$q_m = K \cdot \frac{\Omega}{\omega}$$

in which $q_m = \text{kg/sec.}$

$K = \text{a constant.}$

$\Omega = \text{speed of the turbine wheel.}$

$\omega = \text{speed of the sensor wheel.}$

INFERIOR TO STD. TYPE
METER

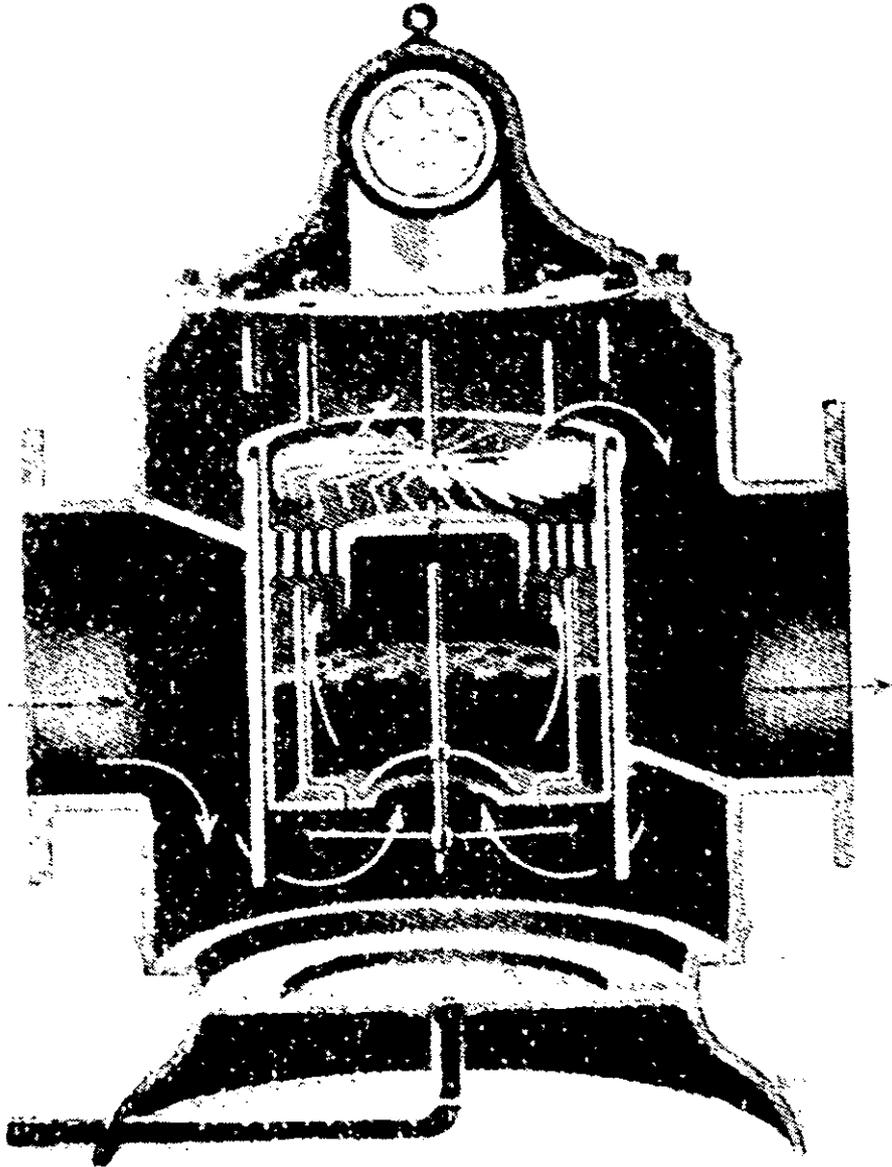
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in Developments in Flow measurement - 1.
R.W.W. Scott (ed.).
published by Applied Science Publishers Ltd.
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les COMPTEURS de VOLUME de GAZ.
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- 2 b Recommendation Internationale No 32: COMPTEURS de VOLUME de GAZ
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COMPTEURS de VOLUME de GAZ a TURBINE.
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1964 Gas und Wasserfach. 105. Jahrg Heft 43.
seite 1192 - 1200 . 23 oktober 1964.
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straighteners with turbine flowmeters for gas. Paper BFL 242. Pre-
sented at the IMEKO VII Congress. London May 10 - 11, 1976.

5) Dijstelbergen H.H., Dynamic response of turbine flowmeters.
Instrument Review 241 - 4. June 1966.

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Gas Line Vol 6 number 2 1980.
publ: Municipal and Utility Division
Rockwell International
Pittsburgh USA.

FIG. 1



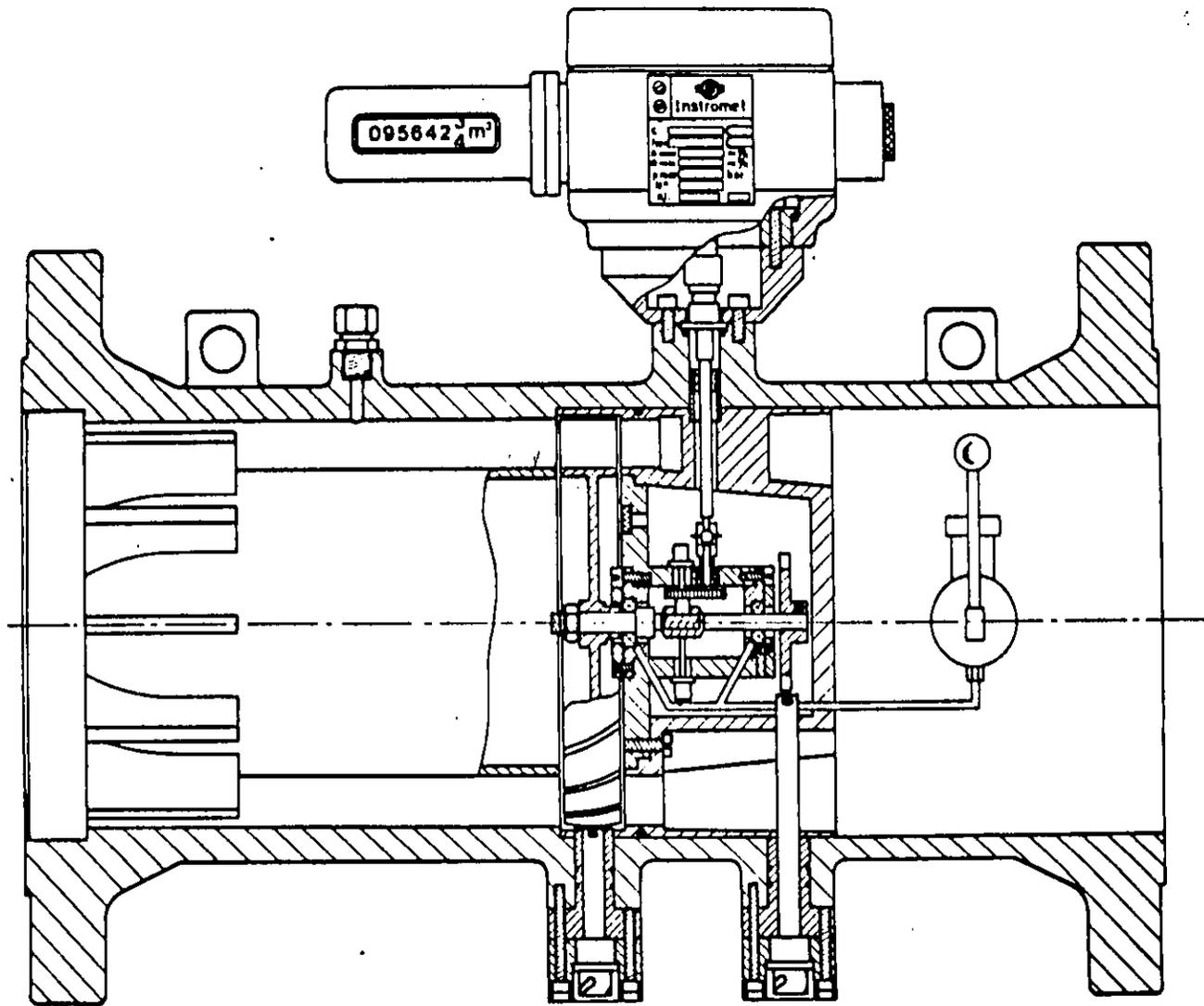


FIG. 2

Certificaat

29110131

Soort meter	: Turbinegasmeter	justeerwielen
Fabriekaat	: Instromet	
Nummer + jaartal	: 27452 - 1975	
G-waarde, type ϕ :	G 16000, SE-RI-D 600 ϕ	telwerk 50 meter 66
Q _{max} /Q _{min}	: 25000 / 1300 m ³ /h	
p _{max}	: 80 bar	

debiet m ³ /h	fout Z 62 bar	fout Z bar
20400	+0,1	
20160	+0,2	
15640	+0,3	
13200	+0,4	
9870	+0,4	
8120	+0,4	
6040	+0,4	
4040	+0,3	
2600	+0,2	
1270	+0,2	

polynoom

$$Y = A_0 + A_1 X^m + A_2 X^n + A_3 X^o$$

waarin:

Y = fout van de meter in Z

$$X = \frac{Q}{Q_{\max}} \text{ in m}^3/\text{h}$$

$$m = -0,2$$

$$n = -0,33$$

$$o = -2,0$$

$$A_0 = -9,3271$$

$$A_1 = 19,7645$$

$$A_2 = -10,3709$$

$$A_3 = 0,0034$$

$$\text{fout} = \frac{\text{aanwijzing meter} - \text{doorgestroomd volume}}{\text{doorgestroomd volume}} \times 100 \text{ Z} \quad (Y = \frac{f - \phi}{\phi} \cdot 100\%)$$

druk bij P_r is bij de bepaling van de fout maatgevend gesteld

testmedium: gas 58 kg/m³

bijzonderheden: plaats Westerbork, datum 30 maart 1981

datum onderzoek: 30 maart 1981

eindstand telwerk: 00019157 x 10 m³

aantal ijkmerken: 11 stuks

opschrift:

vol.massa gas 0,5 - 65 kg/m³

1 m³ = 90,0124 imp.



J.P.A. van Feulen
Afd. Hoeveelheidsmeting Gassen

Fig 3

(3)

FIG. 5

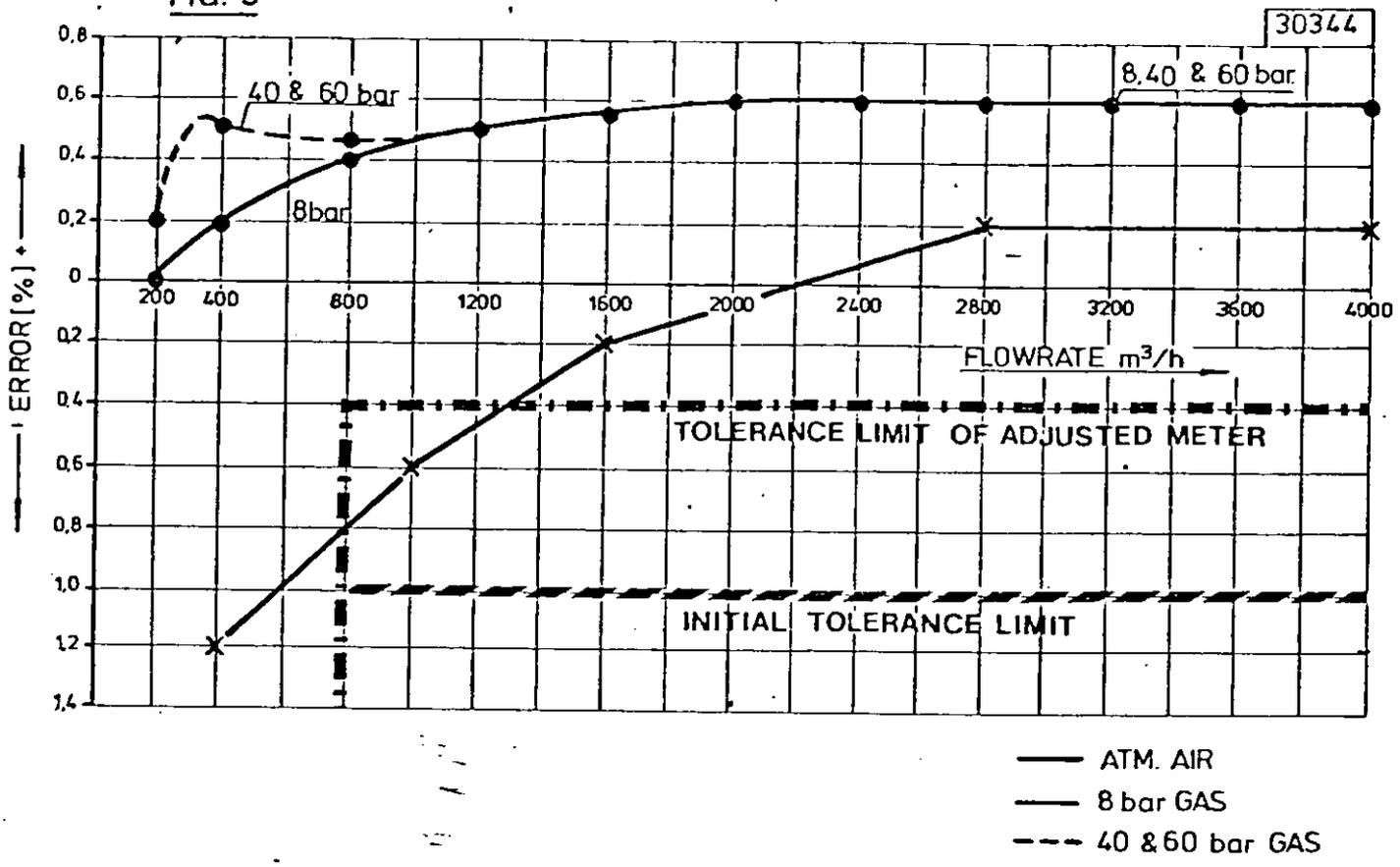
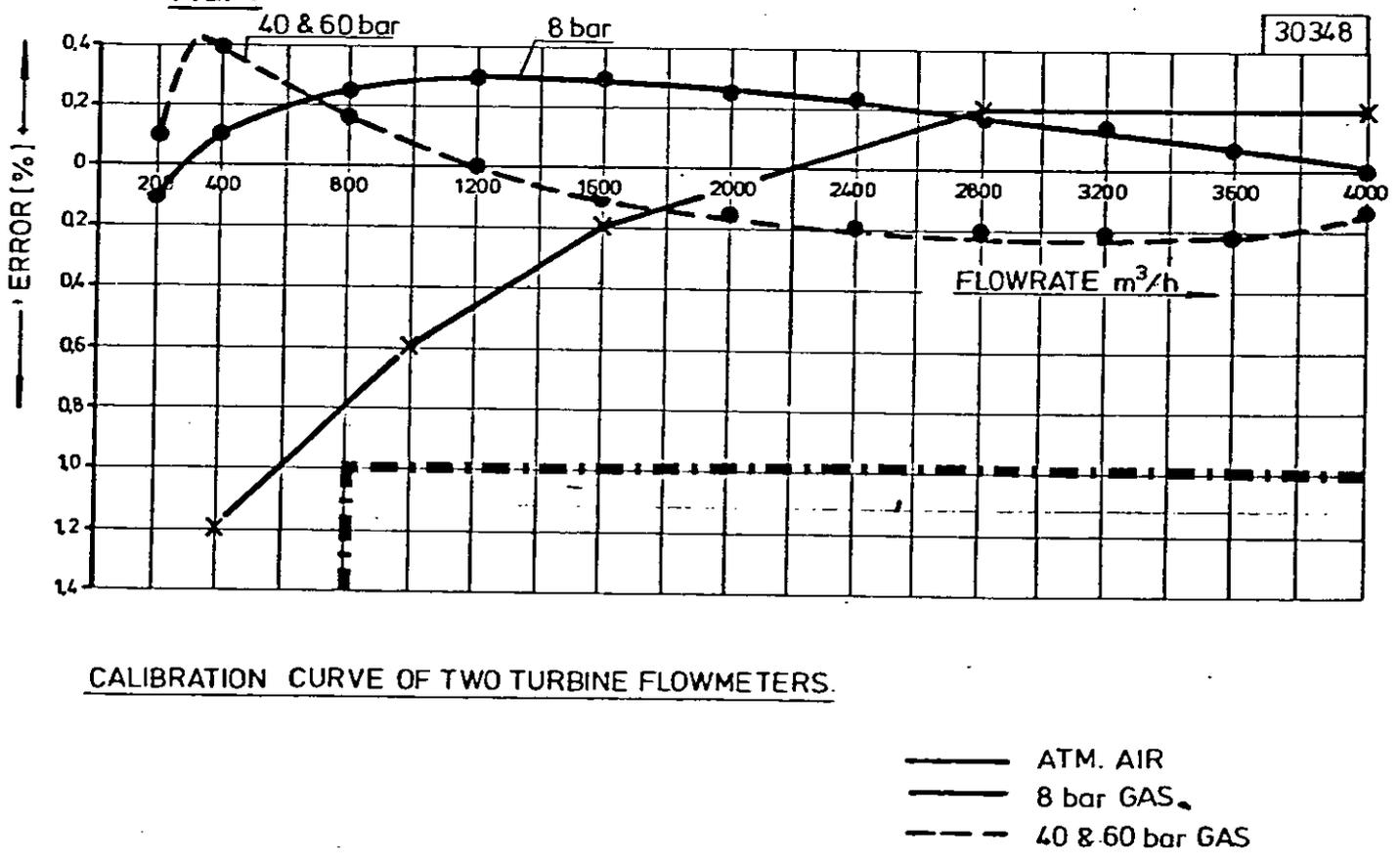
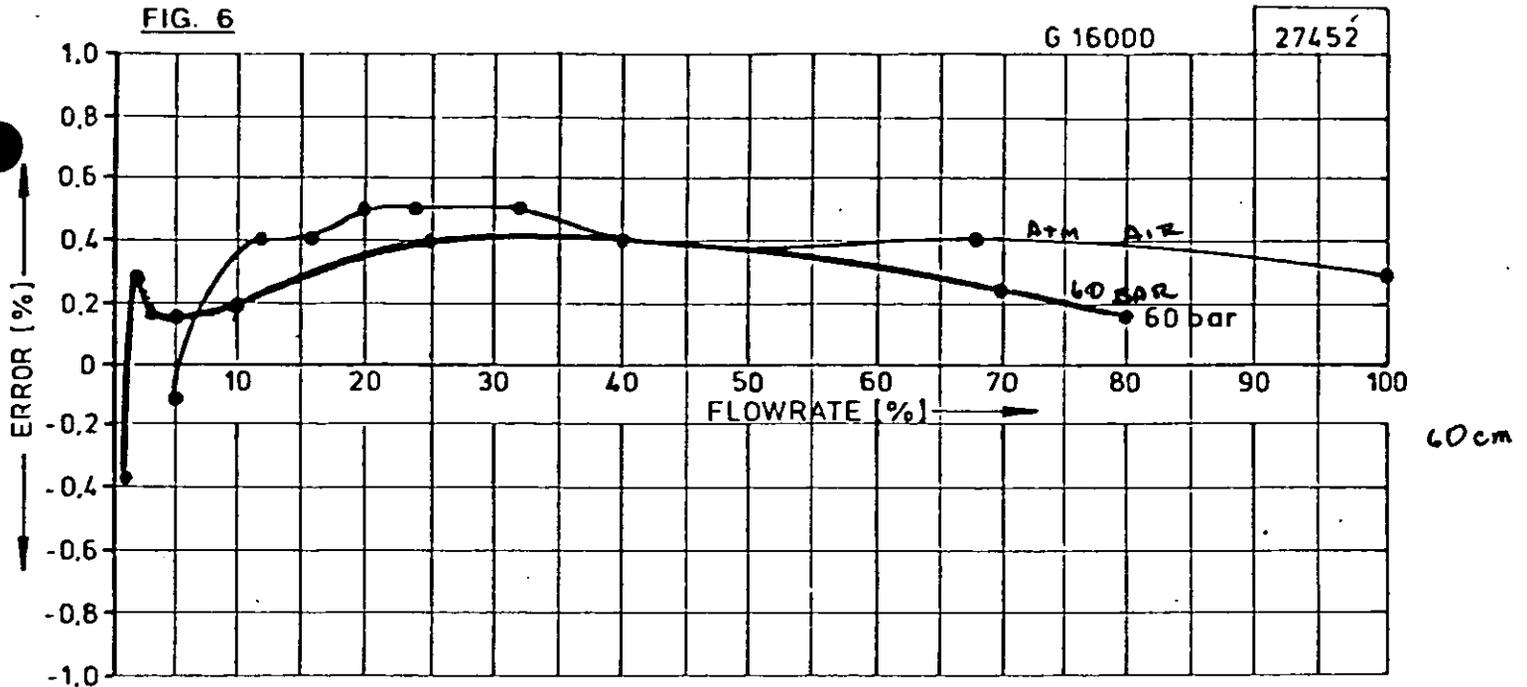


FIG. 4



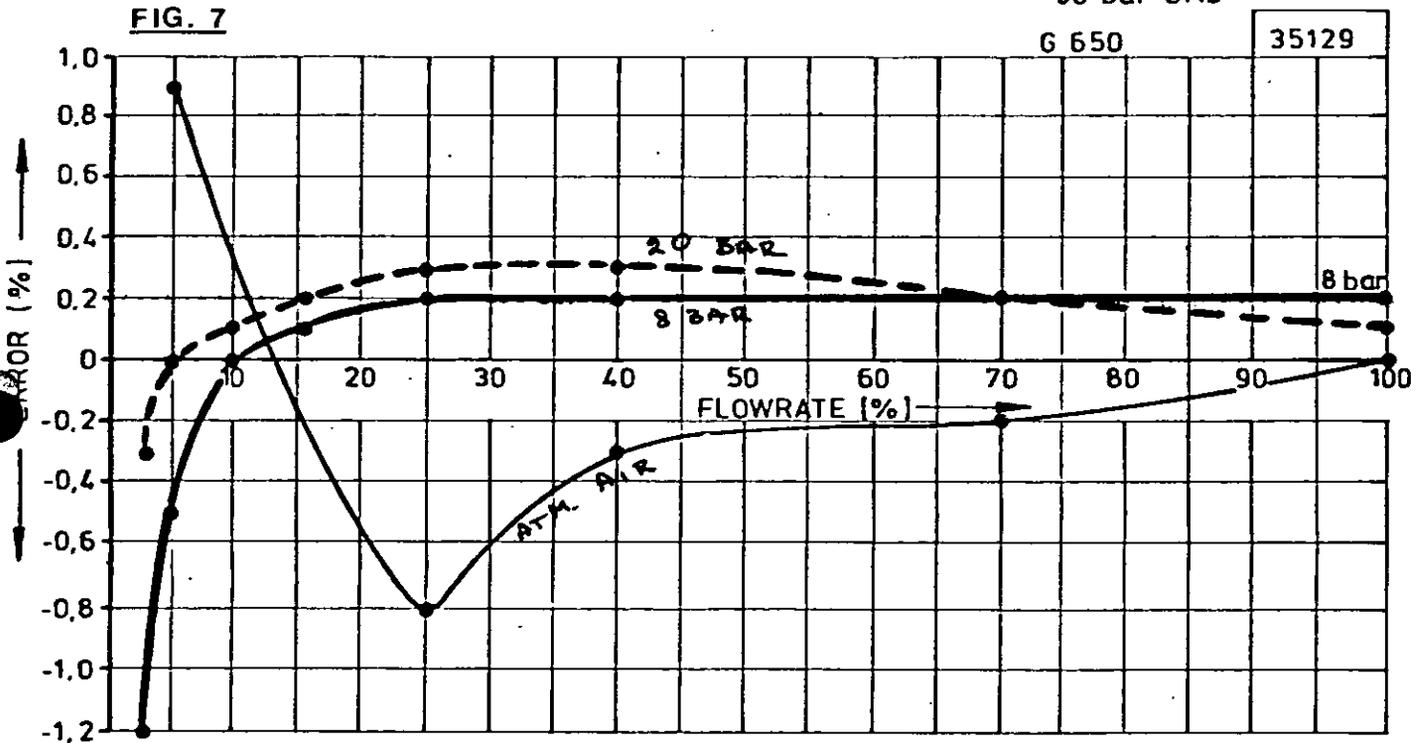
CALIBRATION CURVE OF TWO TURBINE FLOWMETERS.

FIG. 6



- ATM. AIR
- - - 20 bar GAS
- 8 bar GAS
- 60 bar GAS

FIG. 7



CALIBRATION CURVE OF TWO TURBINE FLOWMETERS

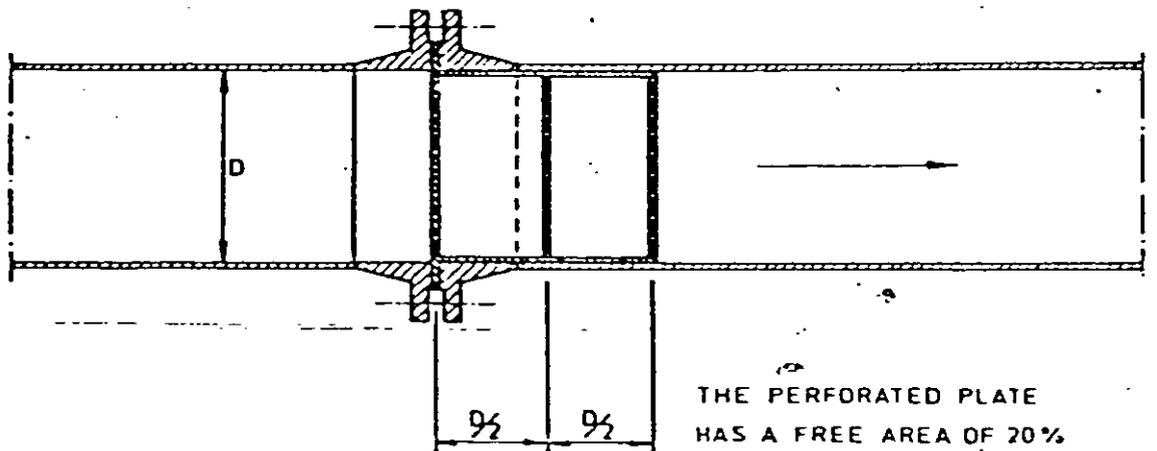
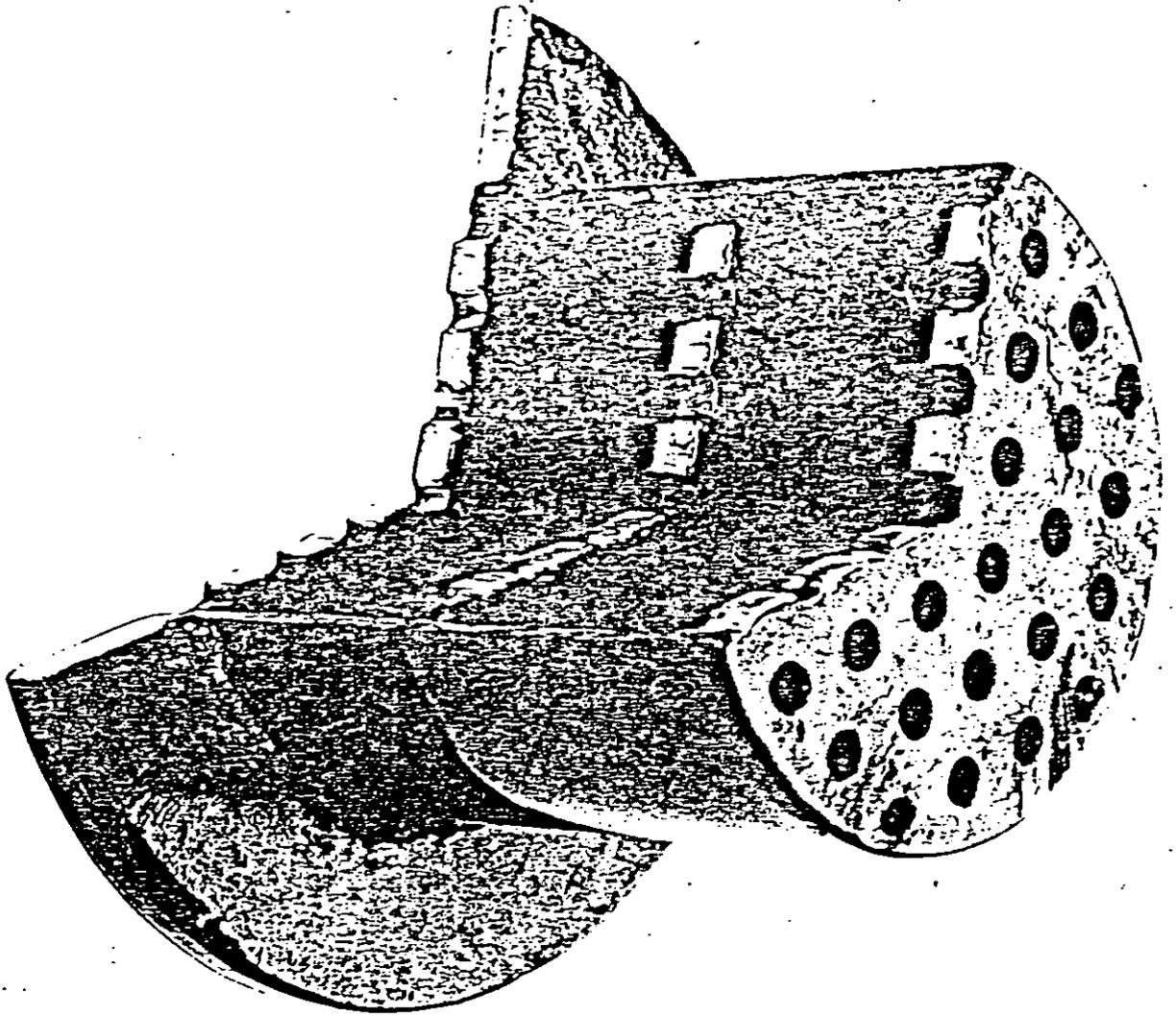
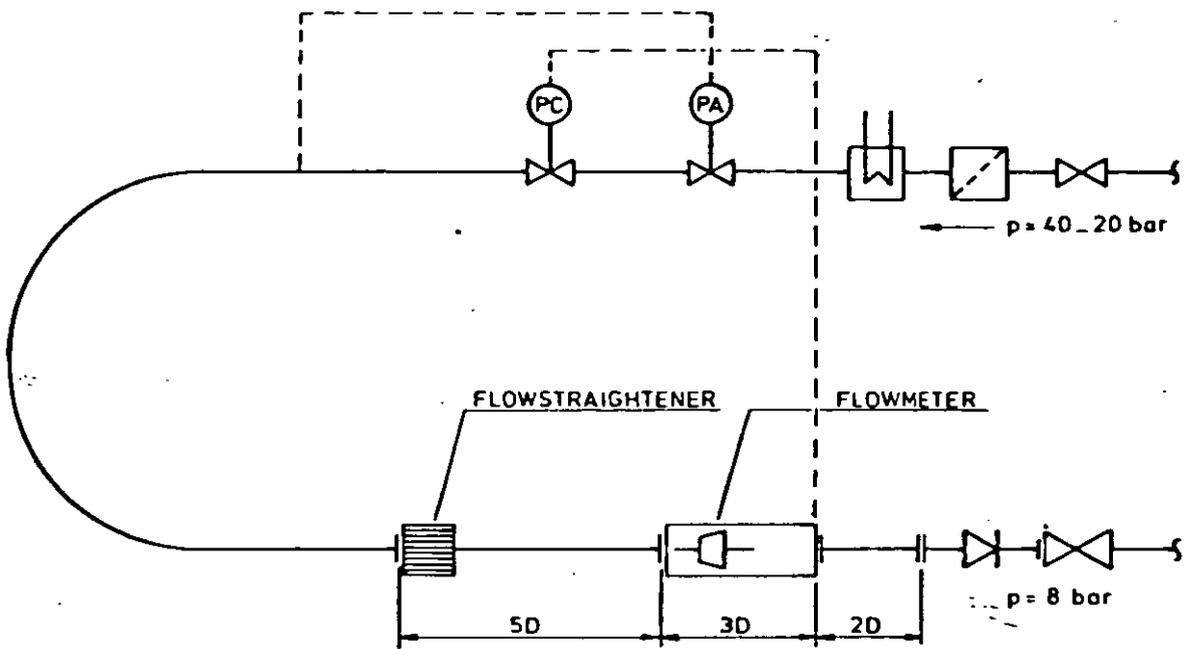


FIG. 8



GAS PRESSURE REGULATING AND METERING UNIT

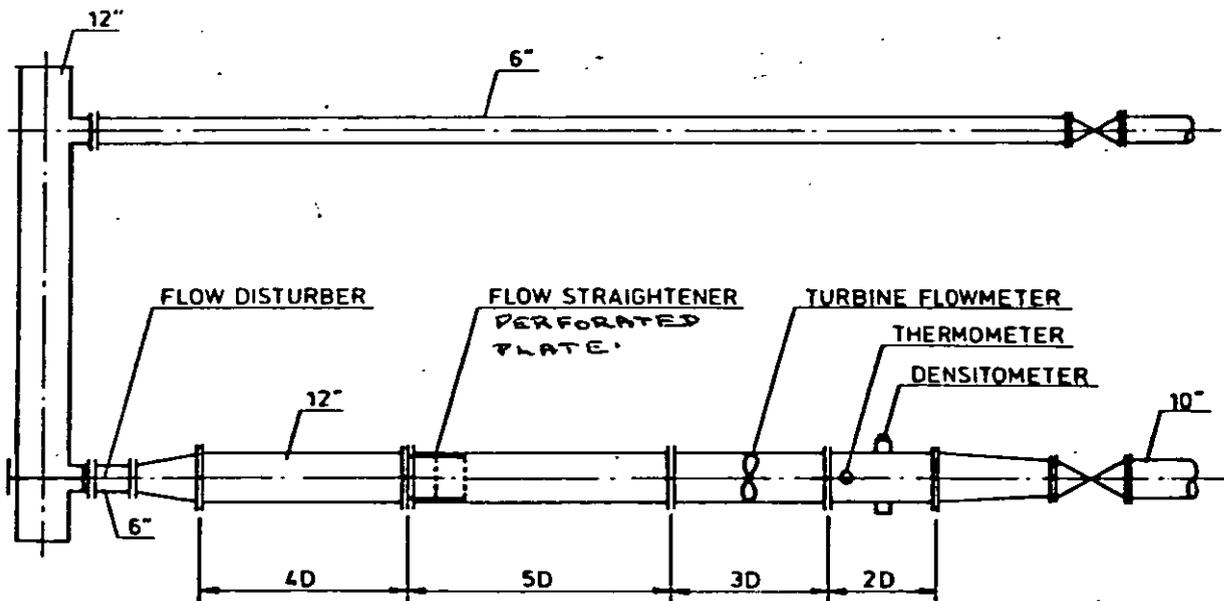


FIG. 9

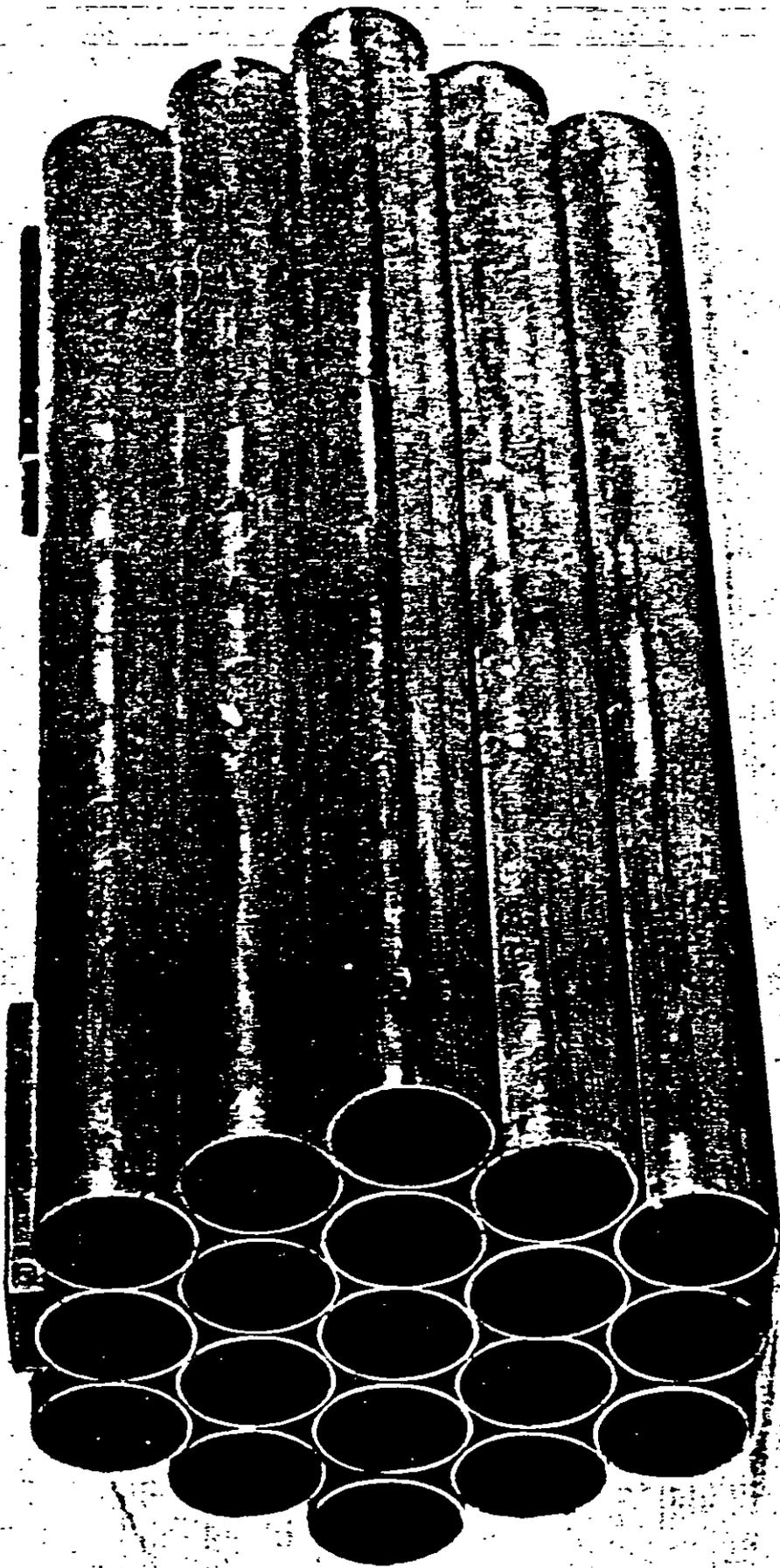
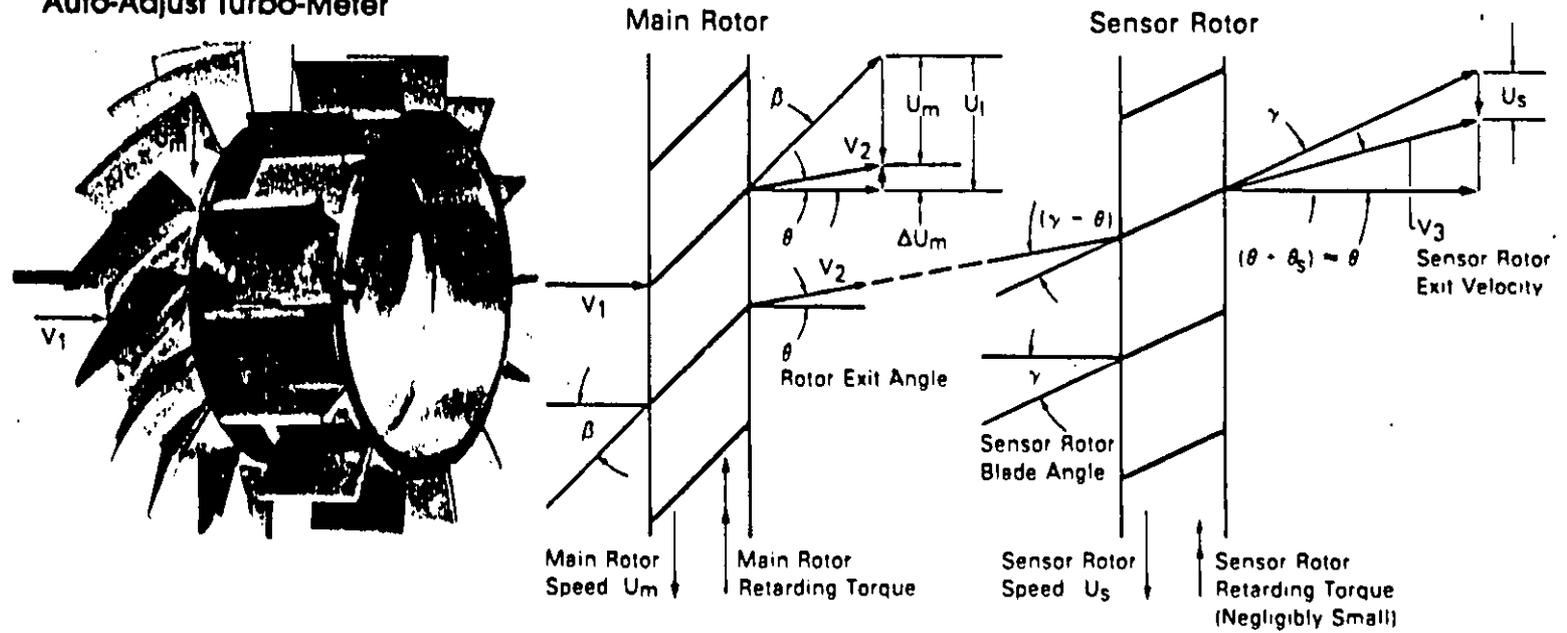


FIG. 10

Auto-Adjust Turbo-Meter

FIG. 11



Defined Auto-Adjust Turbo-Meter Accuracy $\propto \frac{U_m - U_s}{U_1} = \left(\frac{U_m}{U_1} \right) - \left(\frac{U_s}{U_1} \right)$

$$= \left(1 - \frac{\tan \theta}{\tan \beta} \right) - \left(\frac{\tan \gamma}{\tan \beta} \cdot \frac{\tan \theta}{\tan \beta} \right)$$

$$= \left(1 - \frac{\tan \gamma}{\tan \beta} \right) = \text{Constant}$$

FIG. 12

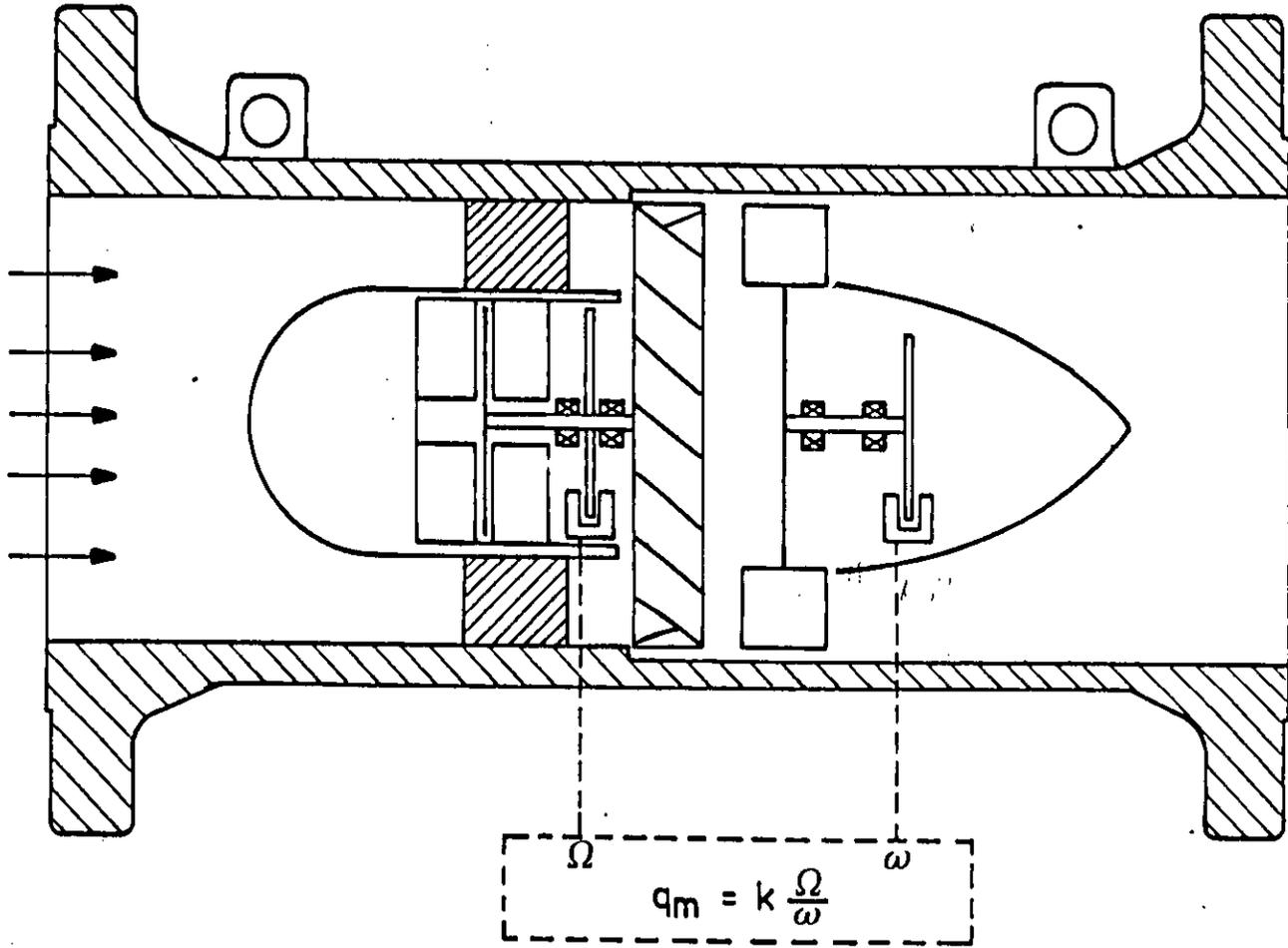


FIG. 12

NORSKE SIVILINGENIØRERS FORENING

MEASUREMENT OF GAS AND LIQUIDS

June 7-10, 1982

Rogaland Regional College

Stavanger

CALIBRATION OF GAS METERS

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Introduction

Right from the start of the gas industry the necessity to measure the product that was sold was appreciated. The way this measurement was accomplished strongly depended on the use that was made of the gas and of its origin.

In the early days, i.e. early 19th century the gas was manufactured gas, almost exclusively used for lighting so the customer was charged for the number of lights and the time he was allowed by contract to burn them. As the application of gas became more diverse the development of gas metering equipment started, finally resulting in the gas meters as we know them today. As manufactured was an expensive product there was a strong incentive to aim at high accuracy. The manufactured gas history mainly took place in Western Europe.

In the United States, however, the start of the gas industry was different. The gas to be handled was natural gas that was won associated with oil which was the target of the drilling activities. Initially it was considered as a waste product and the quantity was quite large. The result of this was that the applications were quite different from the applications of manufactured gas. As the quantities of natural gas were larger its price was lower.

In view of the circumstances mentioned above it is understandable that simplicity and robustness of the measuring system was appreciated more than accuracy. Under this climate the pressure differential flow measuring system was developed resulting in the orifice plate metering system as we know it today.

With the increase of the price of energy the demand for accuracy became the same for all metering systems.

However different the various measuring systems are in construction they have one thing in common. The need for calibration. The methods of calibration are just as different as the constructions.

In the case of the orifice plate the calibration is carried out by carefully measuring the dimensions of the orifice plate and the adjacent tubing and checking if the whole fulfils the requirements of the relevant standard. It is remarkable that there are more than one, quite deviating standards for the same measuring device.

This indicates that development is still taking place.

The other gas meters are of a more complex construction than the orifice plate. The result of this is that it is not possible to predict their behaviour from their dimensions.

The consequence is that this group of meters, and among them the turbine meter, has to be calibrated by comparing their output signal with the output signal of some standard metering device of which the characteristic is known.

According to most of the international and national regulations this calibration had to take place with atmospheric air. The reason for that is the fact that air under atmospheric conditions is a universal medium which easily could be agreed upon at the completion of international directives.

However, it is concluded from experiments that correct registration of atmospheric air is not always a guarantee for good measuring capabilities of gas at high pressure (1).

For this reason several calibration facilities operating with a medium of higher density have been built or are under construction in a number of countries.

In the Netherlands the Metrology Act has been amended to give the results of high pressure calibration facilities a legal status.

In the late seventies an intercomparison measurement campaign has been carried out between the British, French and Dutch calibration facilities (2).

Preparations are being made to give all these installations a common basis, acceptable for the Community Bureau of Reference of the Commission of European Communities.

As soon as this has been accomplished the high pressure calibration can be introduced in the international directives.

In the following a number of high pressure test facilities traceable to primary units or a primary standard for gas measurement will be described.

2. Installations

2.1 France

Of the basic installations in Europe the installation of Gaz de France in Alfortville is the oldest one and for that reason it will be dealt with first.

The GDF has 3 test facilities, namely a primary one (fig. 1) and two secondary ones, for measuring the flow of natural gas at high pressures (3).

Only the installations designated as the secondary test facilities, however, are suitable for the calibration of normal commercial high-pressure gasmeters at higher flowrates. Although different in size the secondary installations do have the same layout and operating principle (fig. 2). Here critical flow nozzles serve as standards for the flow measurement, their coefficients C_D having been determined with the aid of the volumetric method in the primary test facility, fig. 1. The GDF estimate that the measuring uncertainty on the primary test facility is of the order of + 0.25 per cent (4).

In the primary test facility (fig. 1) first of all the gas passes through a filter and subsequently it is controlled by an adjustable pressure-regulating valve to give the desired pressure for the nozzle to be tested. During the measuring phase the flow fills the measuring vessel which has a volume of approximately 2 m^3 . The density of the gas in the vessel in the initial and final state of each test is calculated by means of the pressures and temperatures measured after a period of stabilization and related to the density measured at a location ahead of the nozzle. The instrument, which works on the basis of the principle of the vibrating cylinder and is calibrated with methane, serves as a densitometer.

The density is checked by being calculated from the measured composition of gas as well as from the indication of a measuring instrument for relative density which can be made available.

The volume of the vessel at atmospheric pressure was determined by measuring its content by filling it with water with the aid of a calibration bottle. At the test conditions the increase in volume because of the higher gas pressure is taken into account analytically. In order to achieve a better and quicker stabilization of temperature, the nozzles, measuring vessel, control devices and the inherent piping are placed in a water bath which is temperature controlled.

The measurement is then carried out in the following manner : with the aid of the pressure regulator the desired mass flow through the nozzle which is to be calibrated is set with valves R_1 and R_2 open and a period allowed to obtain stabilization of the flow conditions.

The following measurements are taken :

- a the pressure p_1 ahead of the nozzle,
- b the density ρ_1 ahead of the nozzle,
- c the temperature T_1 ahead of the nozzle,
- d the initial pressure p_i in the vessel, and
- e the initial temperature T_i in the vessel.

Both of the last named measurements serve to determine the initial density ρ_i and thus the mass of the gas contained between the valves R_1 and R_2 inclusive of the vessel at the beginning of the measurement. The second phase of the measurement begins with the closing of valve R_2 . At the same time an electronic timer is started. The gas then flows into the vessel.

The values, p , T and ρ at the inlet of the nozzle have to remain constant and so the filling of the vessel has to be finished before the pressure crosses the threshold given for the maintenance of the critical pressure ratio.

- 6 -

With this design of nozzle this amounts to approximately 90 per cent of the pressure ahead of the nozzle. The valve R_1 is then closed and at the same time the electronic timer is stopped. The stabilization of the temperature of the gas in the volume shut off between the valves R_1 and R_2 has now to take place before the final values p_f , ρ_f and T_f can be measured for this volume. From these values the discharge coefficient of the nozzles is calculated.

The primary test facility is not used (and not suited) for the calibration of meters. The secondary test facilities therefore serve this purpose. They are also fed from the high-pressure natural gas pipeline. In the smallest one (fig. 2) seven sonic nozzles with flowrates, 1.5, 5, 10, 20, 40, 100 and 200 m³/h relative to the standard condition ($p_n = 1.01325$ bar and $T_n = 0^\circ\text{C}$) are available. In the framework of these flowrates other values can also be obtained by connecting the nozzles in parallel.

The coefficients of the standard nozzles were determined using the primary test facility. The nozzles, with straight inlet lengths between 8 and 20D, are installed in the secondary test facility so that the smallest length is applicable to the nozzle with the smallest throat.

The test facility is in the open air. In order to guarantee adequate temperature conditions those parts of the installation which are essential for the test are supplied with thermal insulation. This secondary test facility allows the testing of nozzles and gasmeters up to pressures of 41 bar and flowrates up to 2.6 kg/s at, for all practical purposes, an unlimited duration of the test.

Recently in Alfortville another secondary facility of larger capacity has been built. The standards used are of the same origin as those in the first installation.

The layout of the installation is quite the same as the smaller installation. It is also in the open air which requires thermal insulation. The maximum capacity is $60.000 \text{ m}_s^3/\text{h}$ at pressures up to 50 bar.

In Poitiers at the premises of the Centre d'Etudes Aérodynamiques et Thermiques a calibration facility is installed (5). In this case sonic nozzles, calibrated at the primary installation in Alfortville are used as standards.

The fluid is compressed air. Maximum flowrate is $150.000 \text{ m}_s^3/\text{h}$ at a pressure of 50 bar. At maximum flowrate the measuring time is limited to 120 seconds (fig. 3).

2.2 United Kingdom

In the UK two facilities do exist in which gasmeters can be calibrated with gas or air at elevated pressure. Of these installations one is directly coupled with a primary installation, the other is a secondary installation derived from the primary installation.

The test facility of NEL in Glasgow (fig. 4) is operated with air (6).

First of all a compressor feeds the air into a 12 m^3 storage volume by way of a plant which dries the air and virtually eliminates oil vapours and dust. From the container a loop system with a volume of 6 m^3 is filled with high-pressure air at a pressure of up to 82 bar. When the temperature conditions in the ring conduit are stabilized the valve X is opened. The air - the pressure of which is kept at the desired value by the adjustable pressure controller - flows first of all through a sonic nozzle which serves as a standard and which adjusts to a mass flowrate corresponding to the inlet pressure and the temperature.

From the nozzle the compressed air flows through a switching device either into a high-pressure spherical vessel (diameter 1.5 m) which can be weighed on a scale or into a testline where the flowmeter which is to be calibrated is installed.

Downstream of the testline the air flows through a silencer unit into the open air; various valves make possible an adjustment of the pressure in the testline to the desired value.

Thus the NEL facility is a primary test facility based on the gravimetric method and is combined with a secondary test facility. When the valve X is opened to start the measuring process, the loop system is connected at the same time with the storage container. Thus the air of known temperature which flows from the loop to the critical flow nozzles is replaced by air from the storage container. It is true that this air from the storage container cools off as a result of the decrease of pressure, but the temperature of the air in the testline is not influenced by this so long as there is still warmed air between the inflowing air from the storage container and the outlet of the loop.

Before a gravimetric test can be started the flow conditions through the standard sonic nozzle have to be stabilized. As soon as the readings of pressure at the critical flow nozzle indicate that the conditions are stabilized, the diverter can be switched. An electronic timer is started at the same time by this switching. The spherical vessel that had been weighed previously when it was empty, is now being filled for a given diversion period. At the end of this period the diverter is switched back to its starting position again and the timer is stopped automatically. The vessel can then be disconnected from the filling line and lowered onto a weighbridge scale. From the difference of the weighing in the initial and final state and the time interval of the switched period the mass flowrate can be obtained. From the readings of the pressure and temperature ahead of the nozzle registered during the measuring period the nozzle coefficient can then be calculated.

The direct gravimetric calibration of a flowmeter installed in the testline on the other side of the diverter is not possible. Either a secondary calibration test can be made measuring the mass flowrate with one of the standard sonic nozzles installed upstream of the diverter or an indirect connection with the gravimetric rig can be made.

Calibrations can be carried out at the National Engineering Laboratory at pressures up to 70 bar and at flowrates up to 5 kg/s. The accuracy obtained in the primary measurement as estimated by NEL is approximately ± 0.1 per cent at the 95 per cent statistical confidence levels. The overall uncertainty of a calibration of a flowmeter is estimated to be better than ± 0.3 per cent.

At present British Gas is constructing a calibration facility in Bishop Auckland. This installation is installed in the connection between two gas transmission systems.

In this installation turbinometers will be used as standard meters which are calibrated with sonic nozzles that have been calibrated at NEL. The medium will be natural gas and the maximum flow rate will be $8,5 \times 10^5 \text{ m}^3/\text{st/h}$ at pressures ranging from 35 to 70 bar.

2.3 The Netherlands

In the Netherlands there are 5 installations available to carry out gasmeter calibration with a medium of high density. In four cases this is natural gas, in one case it is ethylene.

The test installation in Groningen (fig. 5) has been set up in the laboratory of Gasunie.

The gas after having passed through the test installation flows into the piping system of the local distribution company.

The temperature of the gas in the installation equals the room temperature. Temperature drop due to reducing pressure is compensated by heat exchangers.

Part A is called "the primary high pressure standard installation". It consists of 10 rotary meters of the CVM-type with a capacity of $400 \text{ m}^3/\text{h}$ each. The CVM meters are calibrated individually with natural gas at atmospheric conditions by means of the 3.5 m^3 bell prover of the Service of Weights and Measures.

A CVM meter (nr. 11), tested with the bell prover as well, and installed at position D is calibrated at a gauge pressure of 8 bar.

The gas after having passed through this meter at high pressure is measured at low pressure by means of the 10 CVM meters arranged in parallel. In order to prevent great differences in temperature a heat exchanger is placed between the meter installed at D and the meters working at low pressure.

Subsequently each of the 10 CVM meters in the installation is calibrated separately at 8 bar gauge pressure with the calibrated meter no. 11. Then meter nr. 1 and nr. 11 were made to change places and the same test cycle is carried out with meter nr. 1.

By carrying out this series of tests with each CVM meter of the installation, sufficient results have been obtained from which accurate error curves could be deduced.

The "primary standard installation" is mainly used for the calibration of meters used as reference meters in other installations. Meters installed at position D can be tested at pressures up to 40 bar. The maximum operating gauge pressure of the CVM meters is 8 bar. This means that the maximum capacity of the installation approximately amounts to $40.000 \text{ m}_{\text{st}}^3/\text{h}$ ($\text{m}_{\text{st}}^3 = \text{m}^3$ at base conditions).

Part B is the installation with which the normal verifications and calibrations are carried out. As shown in fig. 4 the standard meters, being a CVM meter of $400 \text{ m}^3/\text{h}$ and turbine meters of $650 - 1.600$ and $4.000 \text{ m}^3/\text{h}$, can be calibrated directly with the CVM standard installation.

Meters to be tested are installed at position C. If necessary a heat exchanger is installed between the meter under test and the standard meters.

The test installation in Bergum is at the moment undergoing a total reconstruction. The reason to do that is three fold.

First : the installation is modified such that the meter under test and the standard meter always operate at approximately the same pressure.

This has the advantage that during the calibrations the knowledge of the gas composition, i.e. compressibility, only has a secondary influence on the results.

Second : a set of small standard meters is installed so also small meters can be calibrated.

In the new configuration the capacity of the installation will range from $90 \text{ m}^3/\text{h}$ to $200.000 \text{ m}^3/\text{h}$ at pressures ranging from 9 bar abs to 51 bar abs.

Third : by modifying the pressure in flow control system the influence of pressure changes in the outlet piping is minimised or completely avoided.

The gas that has been used in the installation is supplied to a power station.

Schematically the installation is shown in fig. 6.

The gas enters the installation at a pressure of approximately 60 bar.

After having passed a filter the gas is flowing through a combination of a heat exchanger and a bypass in order to compensate for the temperature drop during the eventual pressure reduction.

In the case the pressure during the calibration is below 15 bar gauge pressure the pressure ratio in the outlet control valve is subcritical.

In that case the outlet valve is switched in the pressure control mode and control the pressure in the installation.

The pressure controller at the inlet is set at such a pressure that the pressure drop in the flow control valve, which is set manually, is over critical, thus forming a strong "flow source".

If the pressure during the calibration is higher than 15 bar the pressure drop in the outlet valve is over critical and the outlet valve is switched in the flow control mode and is operated manually.

In this case the flow control valve at the inlet is fully opened and the pressure in the installation is controlled by the pressure controller at the inlet.

In the case the gasflow through the meter to be calibrated is so small that the temperature and pressure controller would be operating outside their range a bypass can be opened to bring the control system within its operating range.

-12 -

The standard meters have been calibrated with the CVM standard meters in Groningen.

In addition to this another three meters are calibrated at Groningen and are used at Bergum as "transfer reference meters"

Periodically the standard meters are checked by comparing them with the "transfer reference meters".

In Westerbork, as part of the Bernoulli laboratory, a meter test installation is situated.

At full operating conditions the capacity of this installation amounts to $2.5 \times 10^6 \text{ m}_{\text{st}}^3/\text{h}$.

The installation is constructed in a bypass around a valve in a main transmission pipe line and operates at line conditions, i.e. a pressure of approximately 60 bar and a temperature of about 7°C.

The gas after having passed through the test installation returns to the same transmission line (fig.7).

The standard meters, being 10 turbine gas meters to 4.000 m^3/h each, and the meter under test are operating at about the same pressure. The standard meters are installed in a building, the meter under test and a part of the tubing at the inlet is located under a pentroof.

The standard meters have been calibrated at Bergum over approximately 60% of their range and over their full range at Westerbork by installing the "transfer reference meters" at the position of the meter to tested. So all three Gasunie calibration facilities are metrologically connected and traceable to the standard of the Metrological Service.

At Utrecht, the Instromet test installation is situated in series with a measurement and control station delivering gas to a big gas distribution company. The standard meters are derived from the CVM-standard installation at Groningen.

The standard meters and the meters to be tested operate at a gauge pressure of 8 bar.

The facility is suitable for testing meters having a capacity of $100 \text{ m}^3/\text{h}$ and higher. The maximum flow rate at which tests can be carried out depends on quantity of gas being delivered to the gas company. In winter time a maximum capacity of $6\,500 \text{ m}^3/\text{h}$ can be achieved.

At the SHELL refinery on the Netherlands a piston prover (8) is constructed for the calibration and legal verification of gas meters to be used for measuring supercritical ethylene (pressure 60 - 100 bar at ambient temperatures).

The ethylene behaves as a gaseous fluid with densities of about $180 - 400 \text{ kg/m}^3$. The prover system is shown in FIG. 8. It consists of a honed pipe (diameter 300 mm), a sealed piston (aluminium), detector switches (1,2,3,4,5), control valves (B, C, D) and a two-position fourway valve (A). The measuring section between the switches 2 and 5 has a volume of 1 m^3 . To avoid leakage between meter and prover, the meter to be tested is placed between the prover and the fourway valve. So the prover can be used only in one direction.

When the piston is near switch 1, fourway valve A is in the position shown in FIG. 8, valve B is open, valve C is closed. Ethylene is flowing through the prover and the meter under test. When the conditions of pressure and temperature are stable valve C is opened, valve B is closed. Then valve C is closed slowly. The increasing differential pressure over C will launch the piston. Valve C is fully closed before the piston passes the pipe in which C is installed. Switch 2 starts the counting of the pulses generated by the meter under test. Switches 3, 4 or 5 are used for stopping the counter. When the piston reaches the end of the pipe, valve D is opened by the differential pressure across the piston.

By turning the fourway valve the direction of the flow in the power changes. Valve D, closes, valve B is partly opened. The piston returns to its start position. When the piston reaches switch 1, valve B is fully opened, the fourway valve is returned to its original position and the next test can be started.

The piston prover is calibrated by the liquid department of the Dutch Metrological Service. The calibration is carried out with water the volume of which is measured with a calibrated meter.

A check on the accuracy of the prover is carried out periodically by:

- comparing the error curve of a turbine gas meter measuring ethylene with its original curve (the meter is only used for this purpose);
- determining the leakage across the piston with nitrogen;
- determining the switching moment of the switches in the measuring section of the prover.

From calculations taking into account the sources which may contribute to an uncertainty in a measurement and from the statistical treatment of the great number of test results available, it turns out that the uncertainty in the results obtained with the high pressure gas calibration facilities amounts to no more than 0,3%. The uncertainty in the results obtained with the piston prover is less than 0,2%.

2.4 United States.

An interesting gas flow reference system is developed as an extension of a cryogenic flow metering reference system at the National Bureau of Standards Laboratories located at Boulder, Colorado (7). The process is shown schematically in figure 9. Basically, the process is a closed loop thermodynamic cycle. The process fluid, nitrogen, is circulated between temperature limits of 85 K and 300 K at pressures of 5 bar (abs) to 41 bar (abs) depending on the point of the cycle under consideration. Work is done on the system by centrifugal pumps operating at 85 K which increase the liquid nitrogen pressure to 41 bar from 5 bar. Heat energy enters the system through a steam heat exchanger which controls the gas temperature at the test section. Heat energy is extracted from the system by refrigeration provided by boiling liquid nitrogen in the subcooler, auxiliary liquid nitrogen introduced at the main heat exchanger and cooling at the water heat exchanger following the gas test section.

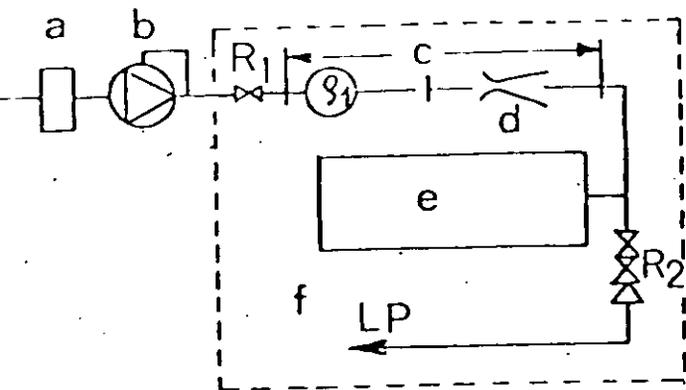
The low pressure cryogenic portion of a cycle is maintained at a pressure of 5 bar (abs) by means of helium gas introduced at the catch and weigh tank. This inert pressurant provides necessary over-pressure to inhibit boiling of the liquid nitrogen, allows liquid phase-gas phase separation for weighing of the liquid nitrogen and provides a controlled environment for the stable operation of the load cell and calibration weights. The interaction of the helium with the liquid nitrogen is negligible as the solubility of helium in liquid nitrogen is less than 0.1 percent. Pumping of the process fluid is accomplished in two steps. The boost pump increases the pressure about 2 bar to the suction of the pressure pump. The pressure pump in turn raises the process fluid pressure to 41 bar (abs). Both pumps are centrifugal types. The boost pump speed is variable while the pressure pump speed is fixed at about 8400 revolutions per minute. Mass flow rates are varied by operation of the expansion valve.

In preparation of a measurement the dump valve at the outlet of the weigh tank is held in the open position until thermodynamic and process equilibrium has been established at a chosen flow rate. This procedure allows liquid nitrogen under helium gas pressure to circulate through the catch to the pump suction. When a test draft is to be run, the weigh tank is closed and sealed and liquid nitrogen accumulates in the weight tank. The force resulting from the liquid accumulating in the weigh tank is measured by the load cell which in turn has been calibrated in reference to standard weights. Mass flow rate is determined by dividing the mass accumulated by the time lapsed since the dump valve was closed.

H.Bellinga/jt

Encl.: 1 up to 9

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HP = high pressure 60 bar
 LP = low pressure 1.1 bar
 a = filter
 b = pressure regulator
 c = test line
 d = venturi nozzle to be tested
 e = vessel
 f = water bassin, temperature regulated

FIG 1 PRIMARY TEST RIG OF THE GDF

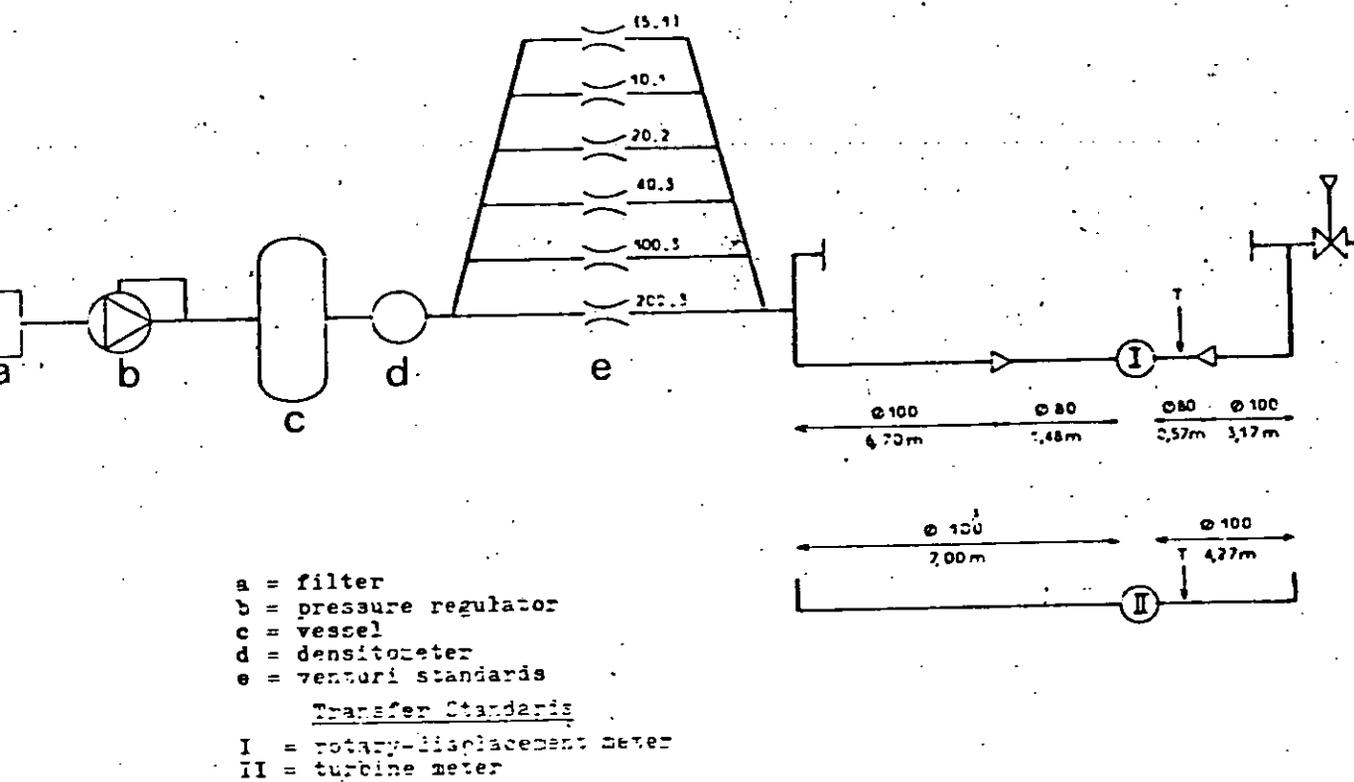


FIG 2 SECONDARY TEST RIG OF THE GDF

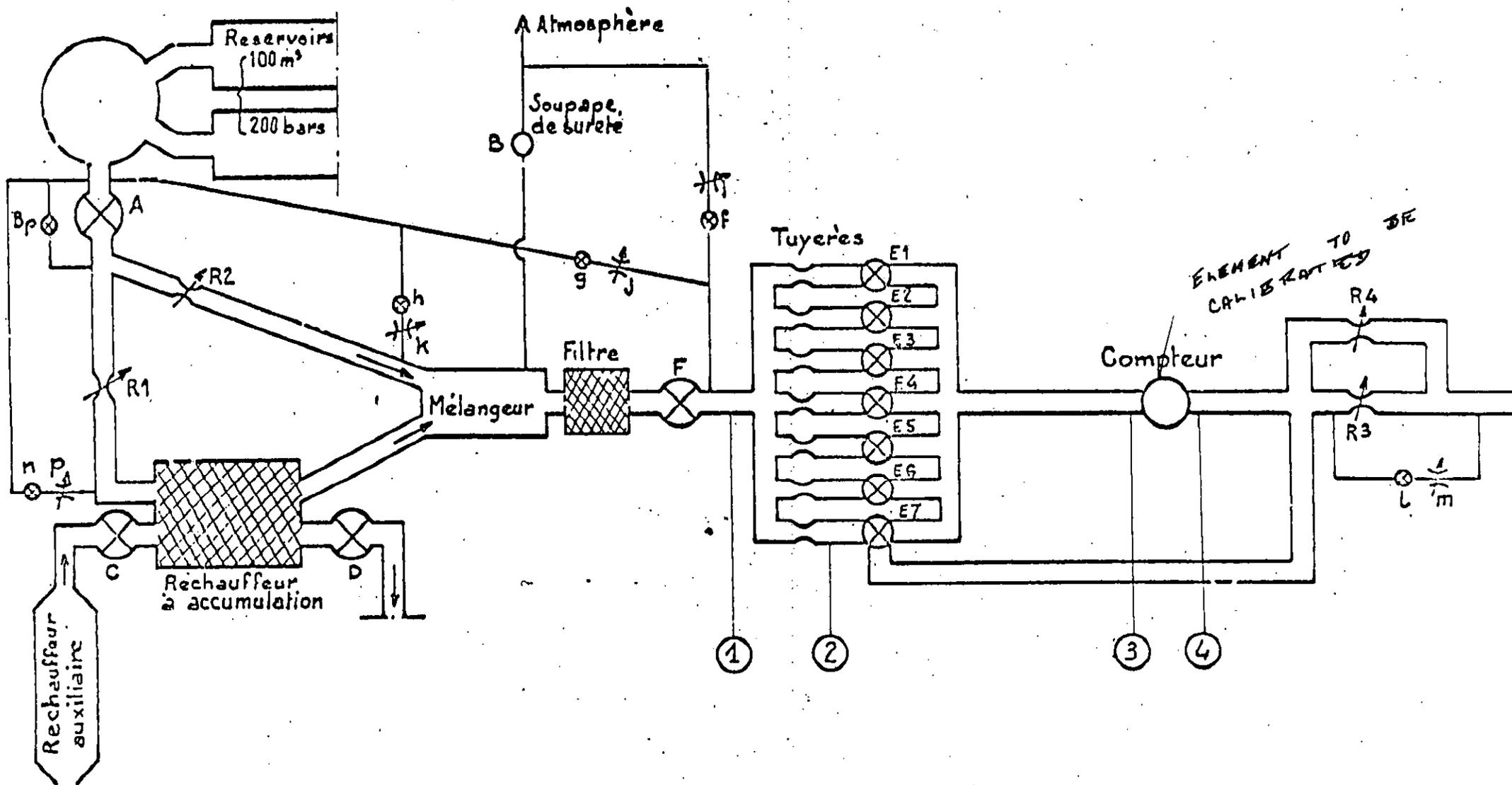


FIG. 3 FLOWSCHME OF TEST AND CALIBRATION RIG FOR FLOWMETERS.

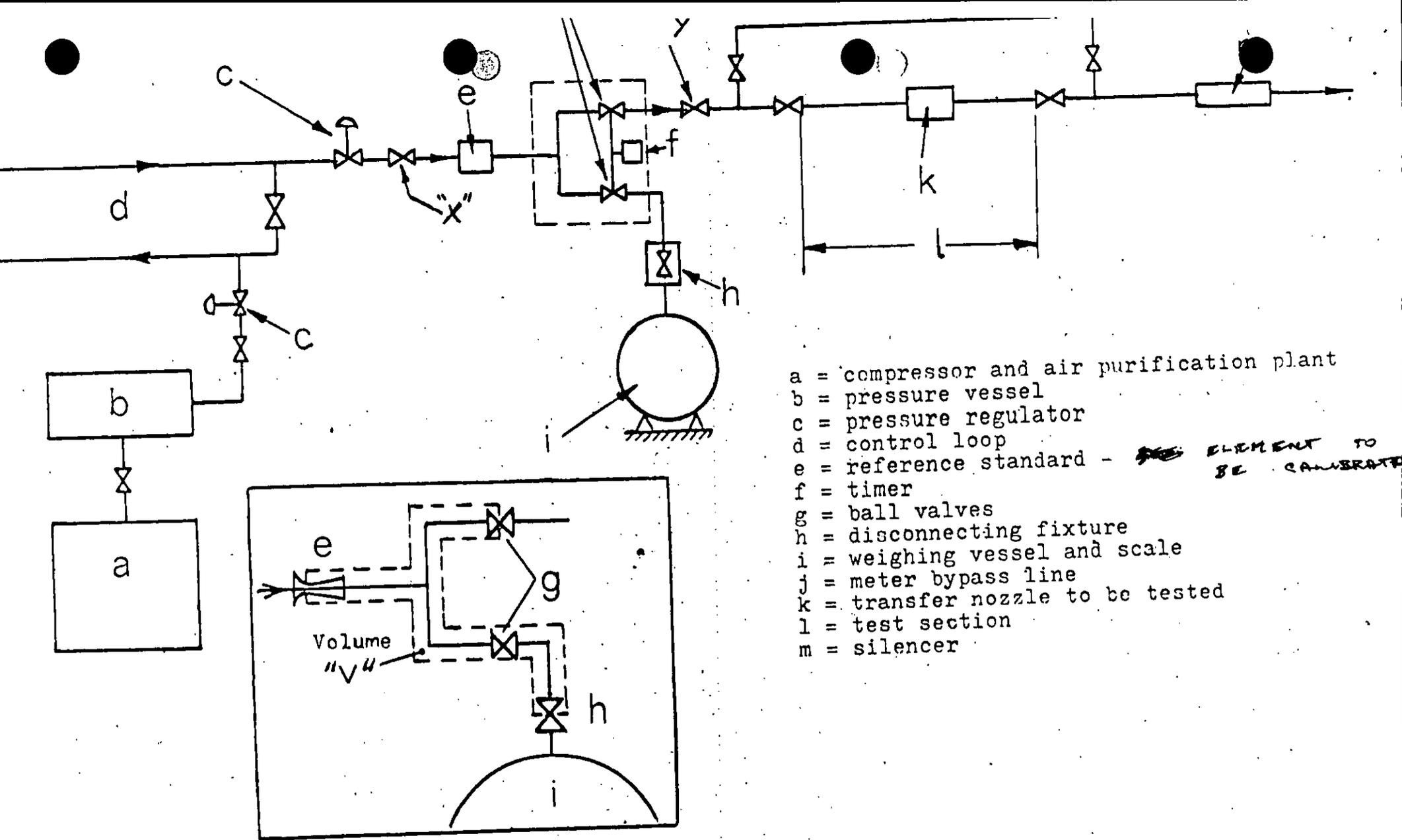


FIG 4 SIMPLIFIED LINE DIAGRAM OF THE GRAVIMETRIC SYSTEM AT NEL.

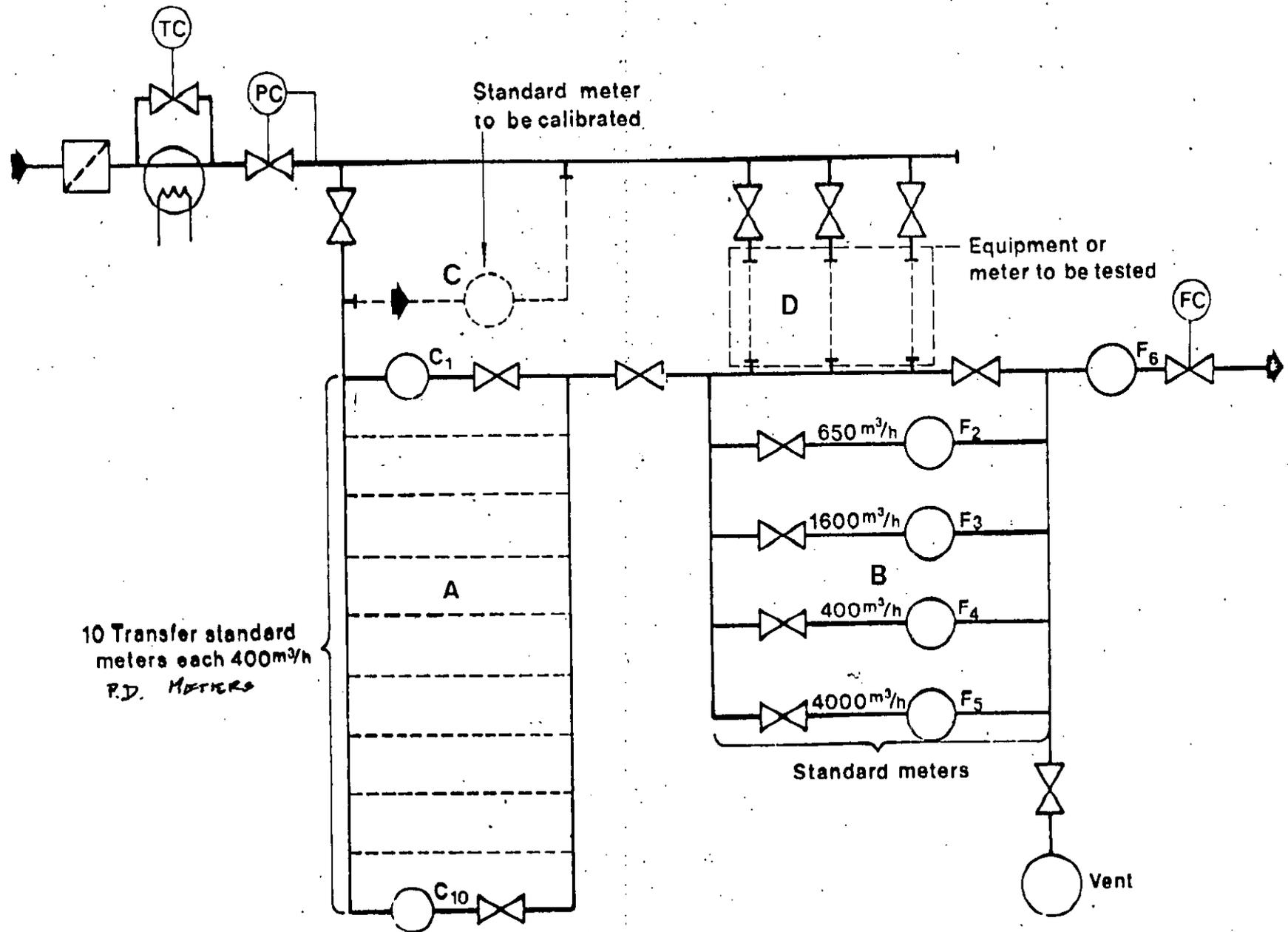


Figure 5; Research flow rig Groningen

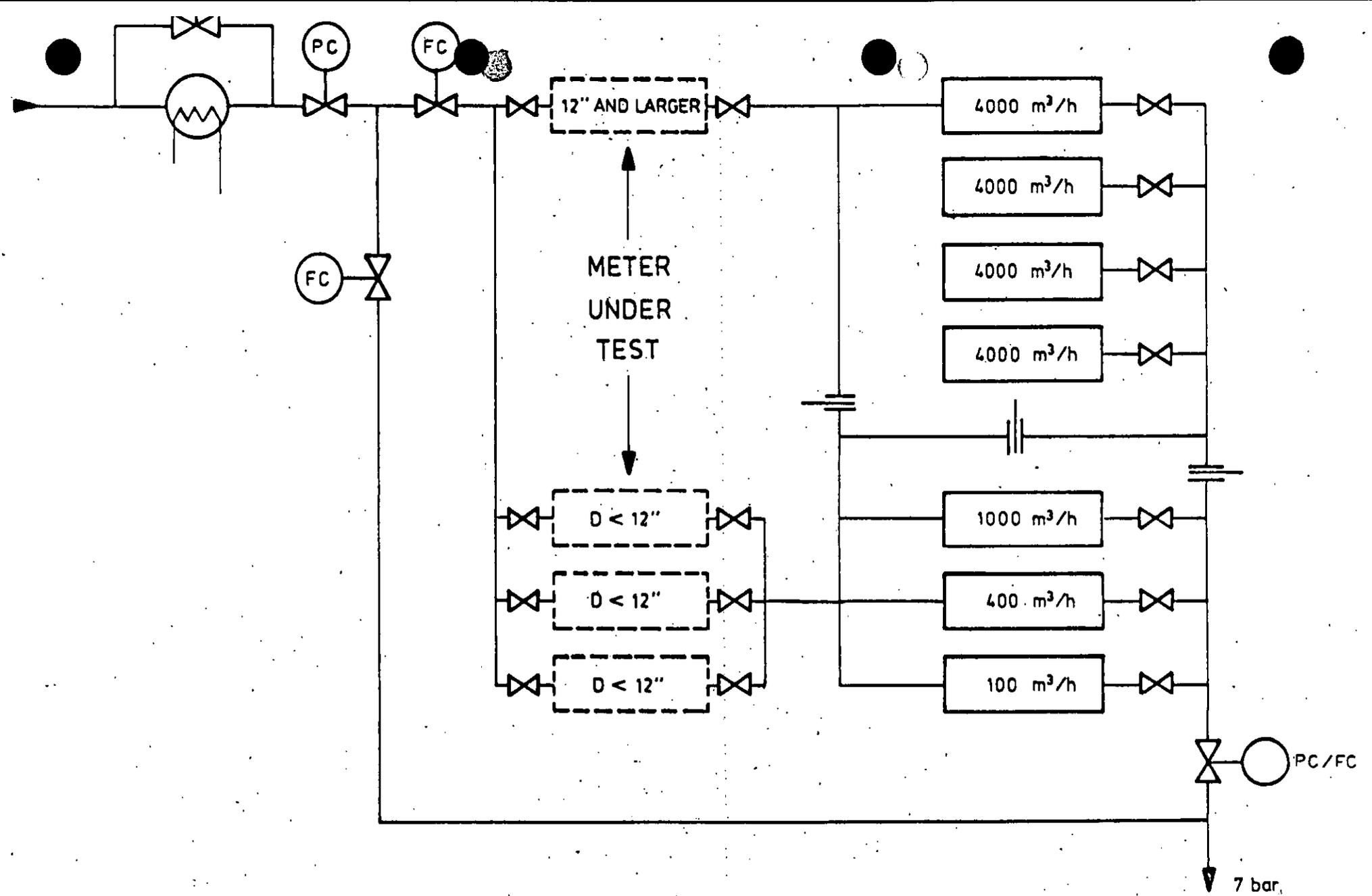


FIG. 6 CALIBRATION FACILITY BERGUM

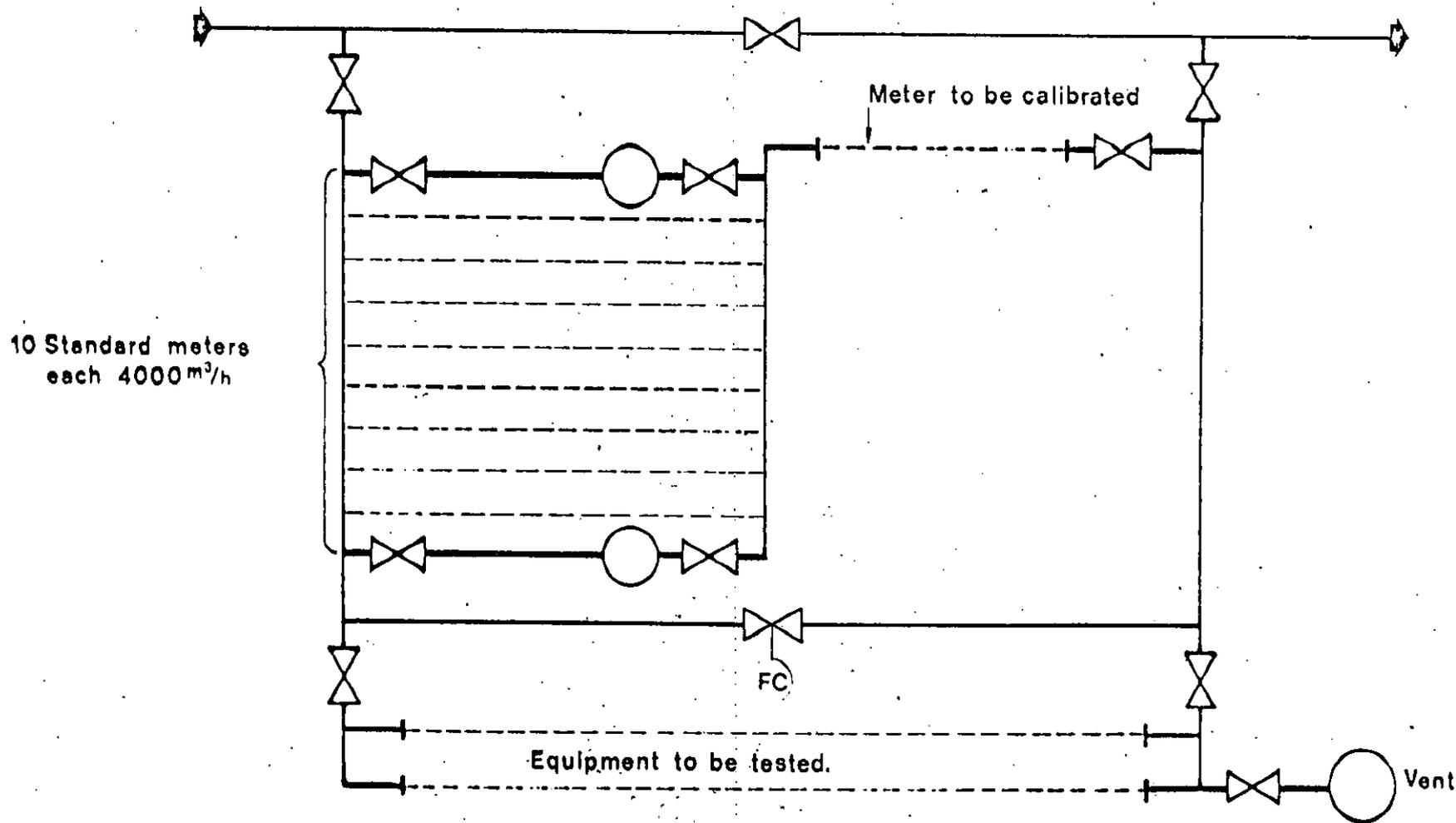


figure 7; High flowrate test facility "Westerbork"

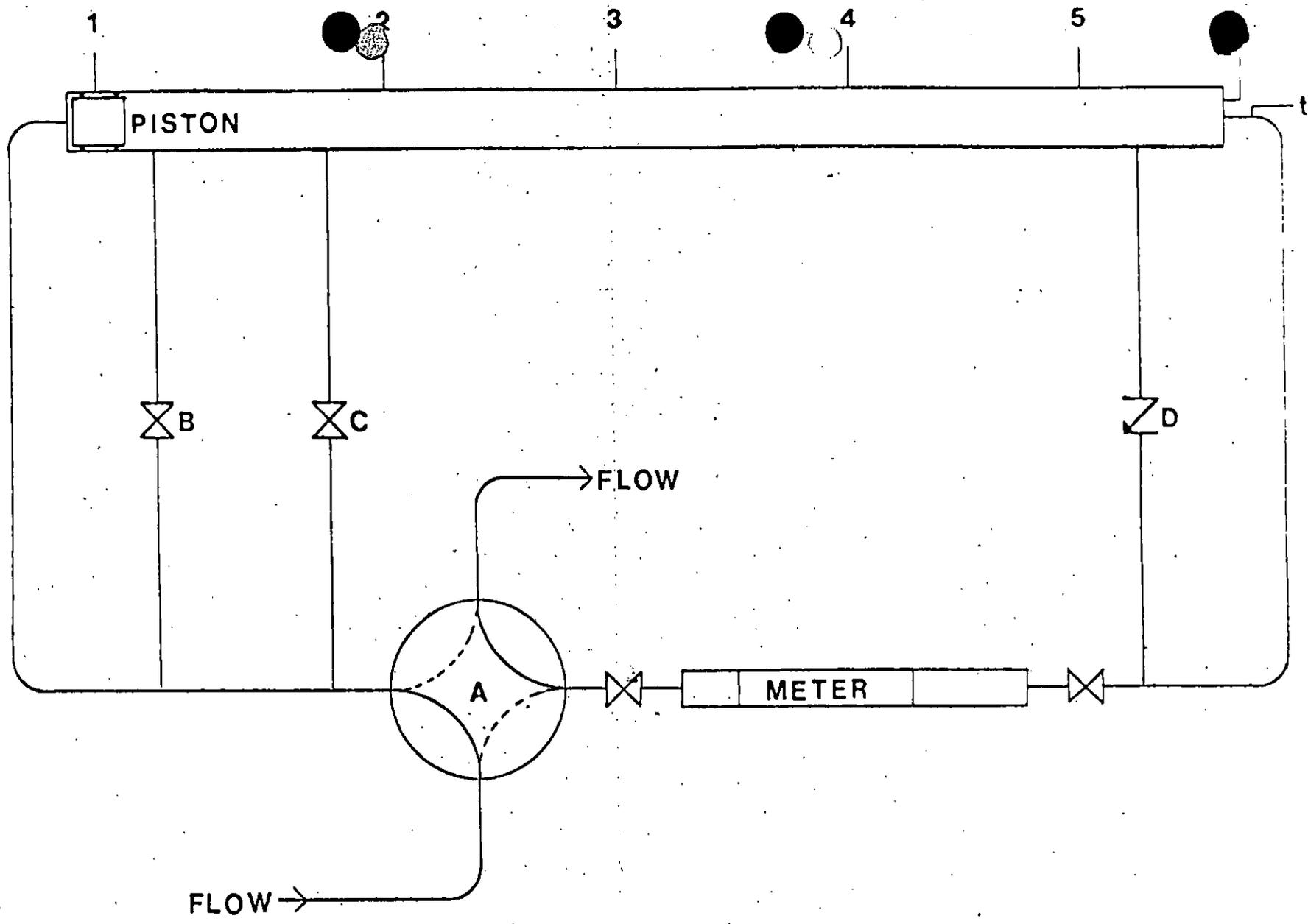
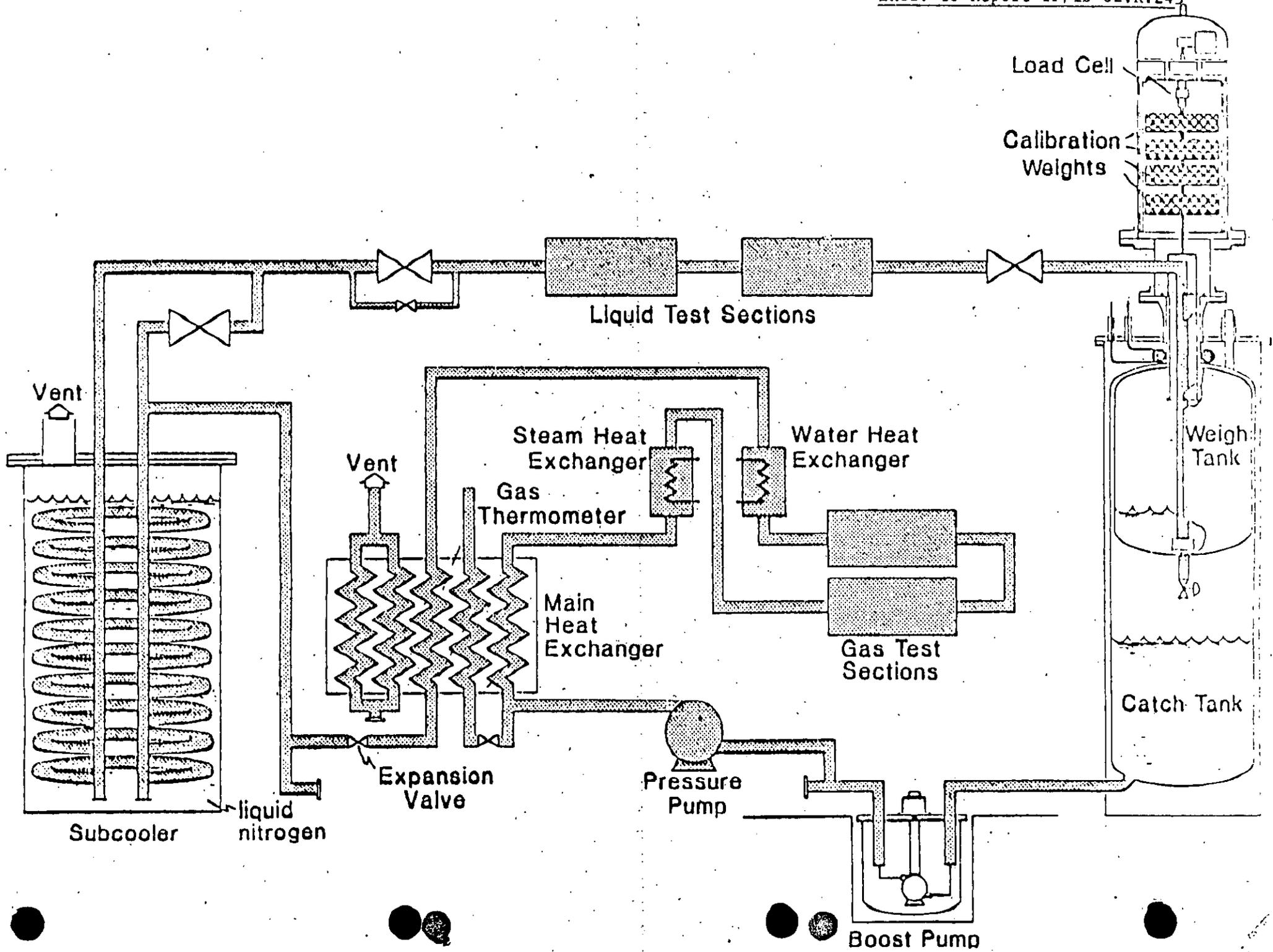


FIGURE 8



MEASUREMENT OF GAS AND LIQUIDS

June 7-10, 1982

Rogaland Regional College

Stavanger

TURBINE METERS

Lecturer: Peter A.M. Jellfs
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TURBINE METERS

INTRODUCTION

This paper reviews the design, performance and application of turbine meters for custody transfer measuring systems.

The details of the mechanical displacement provers are reviewed separately in another paper.

DESIGN

2.1 Theoretical Design Considerations

A simplified equation showing the relationship between the torque due to bearing friction and pick-up drag is given below:

$$T_{B+P} = n \int_{r_1}^{r_2} \frac{1}{2} \rho V_R^2 (C_L \sin \beta - C_D \cos \beta) l r \cdot dr$$

- Where:
- n is number of blades of rotor
 - r₁ is radius of root of blade
 - r₂ is radius at tip of blade
 - l is length of blade
 - ρ is density of fluid
 - C_L is coefficient of lift
 - β is relative velocity angle
 - C_D is coefficient of drag
 - r is radius
 - T_{B+P} is torque from bearing and pick-up drag

An ideal turbine meter requires a constant meter factor, so that the velocity of the blades must be directly proportional to the velocity of the flowing fluid. It is evident from figure 1 that this will only occur if the angle β is the same at all flowrates. If this were to be the situation then the above equation can be reduced to the form given below where A and B denote composite constants

$$\begin{matrix} \text{LIFT} & \text{DRAG} & \text{TORQUE} \\ AC_L & = & BC_D + T_{B+P}/V_R^2 \end{matrix}$$

This equation is only valid if each separate term within it remains the same at all flowrates. The first term satisfies this condition, as the coefficient of

lift does not vary. However, this is not fulfilled with the second term because C_D - the coefficient of drag varies with Reynolds number. Also the third term, T_{B+P} , the torque due to the combined effects of bearing friction and pick-up drag does not increase in proportion to the square of the relative velocity.

It is evident, therefore, that a meter cannot be designed to produce a constant K factor.

The art of the manufacturer is to produce a meter with a near constant K factor over a wide working range of flowrate.

The above equation assumes a flat velocity profile i.e. the velocity is the same at all points in the cross-section of the pipe whereas in practice the velocity profile in the turbulent range ($Re > 3000$) will be a parabola. This is another reason why it is not possible to exactly predict the behaviour and performance of a turbine meter.

.2 Performance

a. Characteristic Curves

All turbine meters have a characteristic curve (K factor versus flowrate) of a general form (see figure 2).

The best meters may achieve a variation in K factor of $\pm 0.25\%$ over a flow range of 10 to 100%. However, there will always be a sharp fall in K factor at the low flowrates because the coefficient of drag increases at low Reynolds numbers and because of the increased significance of bearing friction at low rotation speeds.

b. Linear Meters

In order to improve the linearity some manufacturers have recently designed turbine meters so that the torque due to the bearing friction and pick-up drag have been reduced.

One meter utilizes two rotating elements instead of one; and up-stream indicating turbine rotor which induces the signal and a down-stream slave turbine rotor. The shaft with the slave rotor attached rides on one set of bearings and the indicating rotor rides upon a separate set of bearings attached to the rotating shaft. The separate shaft arrangement ensures that the relative motion between the indicating rotor and its bearing remains at a near zero level irrespective of the velocity of the fluid passing through the meter. The pick-up drag normally due to inductance or variable reluctance when generating pulses is eliminated by using a sensing system based on high frequency radio-wave signals.

Another meter employs a rotor which rotates on a tungsten carbide shaft which in turn rotates in two tungsten carbide journals. This arrangement ensures that relative motion between the rotor and the journals is reduced.

One important advantage of these new types of meter is that they can withstand very high flowrates (sonic velocity) associated with vapour boil off conditions often encountered in LPG systems without bearing failure.

c. Viscosity, Density and Size of Meter (See figures 3 & 4)

The effects of viscosity becomes progressively greater as the size of meter becomes smaller. Also the linearity deteriorates as the meter becomes smaller. The effect of increasing viscosity is not only to change the K

factor but it also reduces the rangeability of the meter. (This is due to the increased bearing friction with the more viscous oils). As the density of the fluid is reduced the linearity of the meter deteriorates. This is mainly due to reduction in the fluid momentum available for overcoming the rotor torque.

APPLICATION

In order to achieve a satisfactory performance from a turbine meter a number of conditions must be fulfilled when designing the installation.

Swirl

Liquid swirl in a flowing liquid is mainly caused by pipe bends and fittings and can effect the K factor and repeatability. (Swirl is not constant).

In practice it is necessary to install a flow straightener (5 diameters) upstream of the meter and to ensure that there are no pipe fittings directly downstream of the meter.

Cavitation

In order to prevent cavitation in the meter it is essential to maintain a back pressure above the minimum specified by the manufacturer.

Air

In order to remove entrained air - particularly in viscous oils - it is necessary to ensure that the level of liquid in the tank supplying the meter is 2-3 metres above the suction or alternatively to keep the floating roof floating at all times.

REPEATABILITY

The turbine meter can be likened to a flywheel as its function is to dampen down the random or individual variations in velocity of the flowing fluid.

With meters such as the Ultrasonic or Vortex with no moving parts it is necessary to have sufficient volume throughput in order to obtain a good repeatability. Whereas turbine meter repeatability can be achieved with a relatively small throughput volume.

INTEGRATION OF THROUGHPUT

a. Variations in Flowrate

Experience has shown that there are often considerable variations in the flow conditions in offshore production metering and onshore ship loading terminals. While there are often dedicated pipe provers on hand to prove the individual meters, there are usually problems in carrying out this task.

For instance, the meter must be calibrated immediately after the flow conditions alter significantly or errors may be incurred in the integration of the throughput due to the incorrect meter calibration factor being applied.

5. Linearising K Factors

One method of overcoming the problem of meters sensitive to changes in flow conditions is to use a micro-computer programmed to read the flowrate, temperature and pressure signals in the line and to apply a varying K factor to the meter integration of throughput.

7. Calibration Procedure (See Fig.5)

A method of combining central proving with on-site proving is described below:

a. Laboratory Proving

The performance of the meter is usually established by initially proving at a central laboratory where the curves of K factor versus flowrate are obtained with a pipe prover using several oils of differing viscosities. At least three flowrates are required at each viscosity in order to determine a curve.

b. Meter Performance

The data obtained from the initial proving are plotted on a performance chart. A suitable equation, usually a polynomial derived by a least square method, is fitted to the proving data or alternatively a matrix for use with linear interpolation technique is derived. This information is programmed into a micro-computer which will read the flowrate (frequency), temperature and pressure as transmitted from a number of transducers installed directly in the metering runs. As there is a linear relationship between temperature and log viscosity for each type of crude oil it may be sufficiently accurate to measure temperature rather than viscosity directly in the line.

c. On-Site Measurement

The meters are then installed on-site and reproved. The micro-computer then automatically applies the appropriate K factor to the meter scaler for the flow conditions experienced during the throughput measurement integration.

d. Re-Proving

It is necessary to re-prove the meters at regular intervals so as to establish the long term scatter (see Fig.6) and to up-date the original performance curve. The mean K factor curve can then be established over a period of months and set into the micro-computer. This would shift the emphasis from changing K factor to monitoring K factor. (See Fig.7)

The long term drift of K factor with time due to bearing wear or rotor damage can also be monitored by analysing the moving average of 10 consecutive (period) K factors.

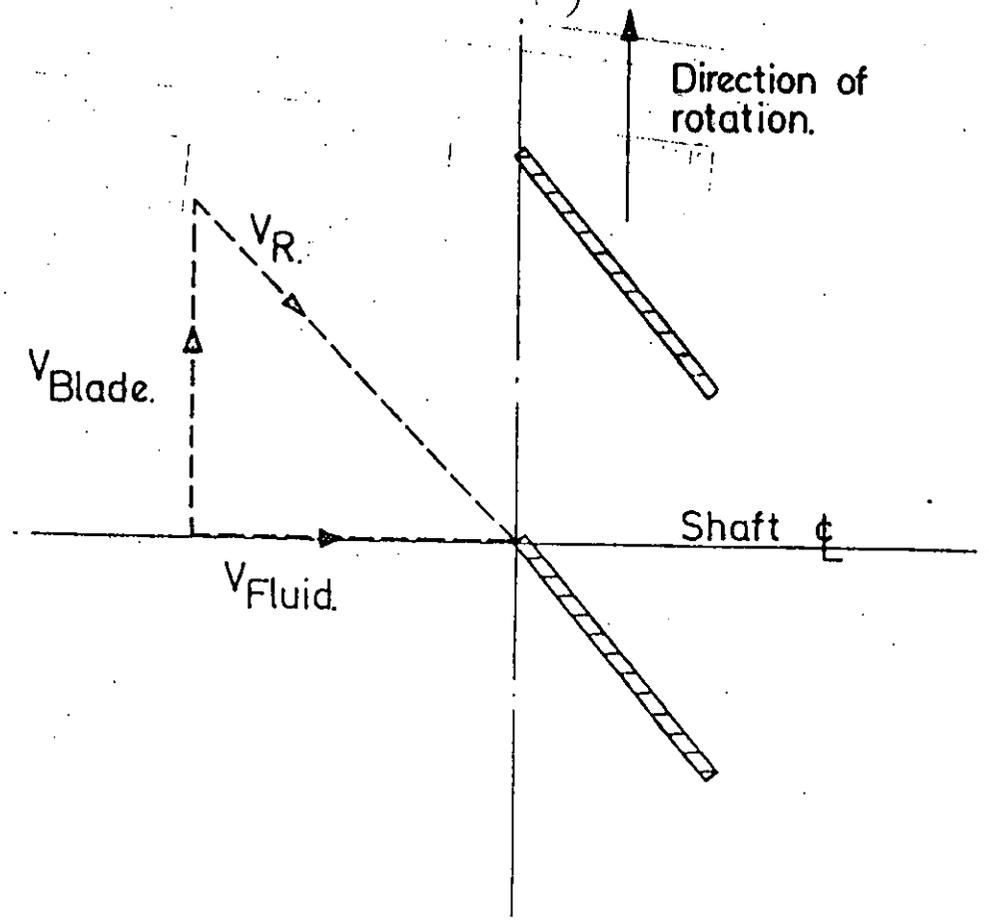


FIG. 1 INLET VELOCITY DIAGRAM.

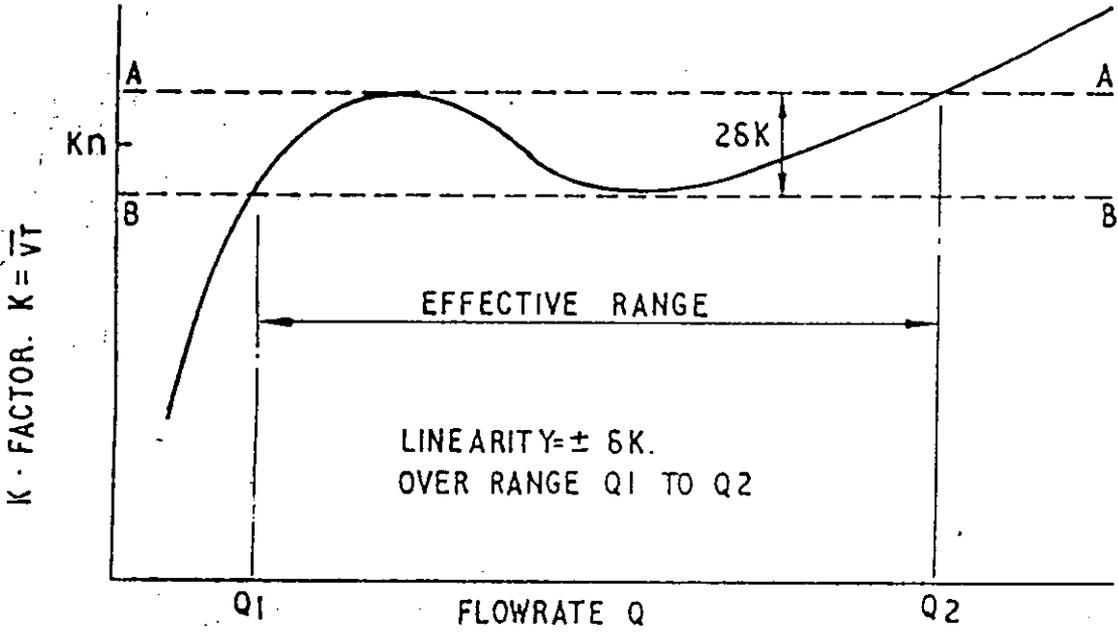
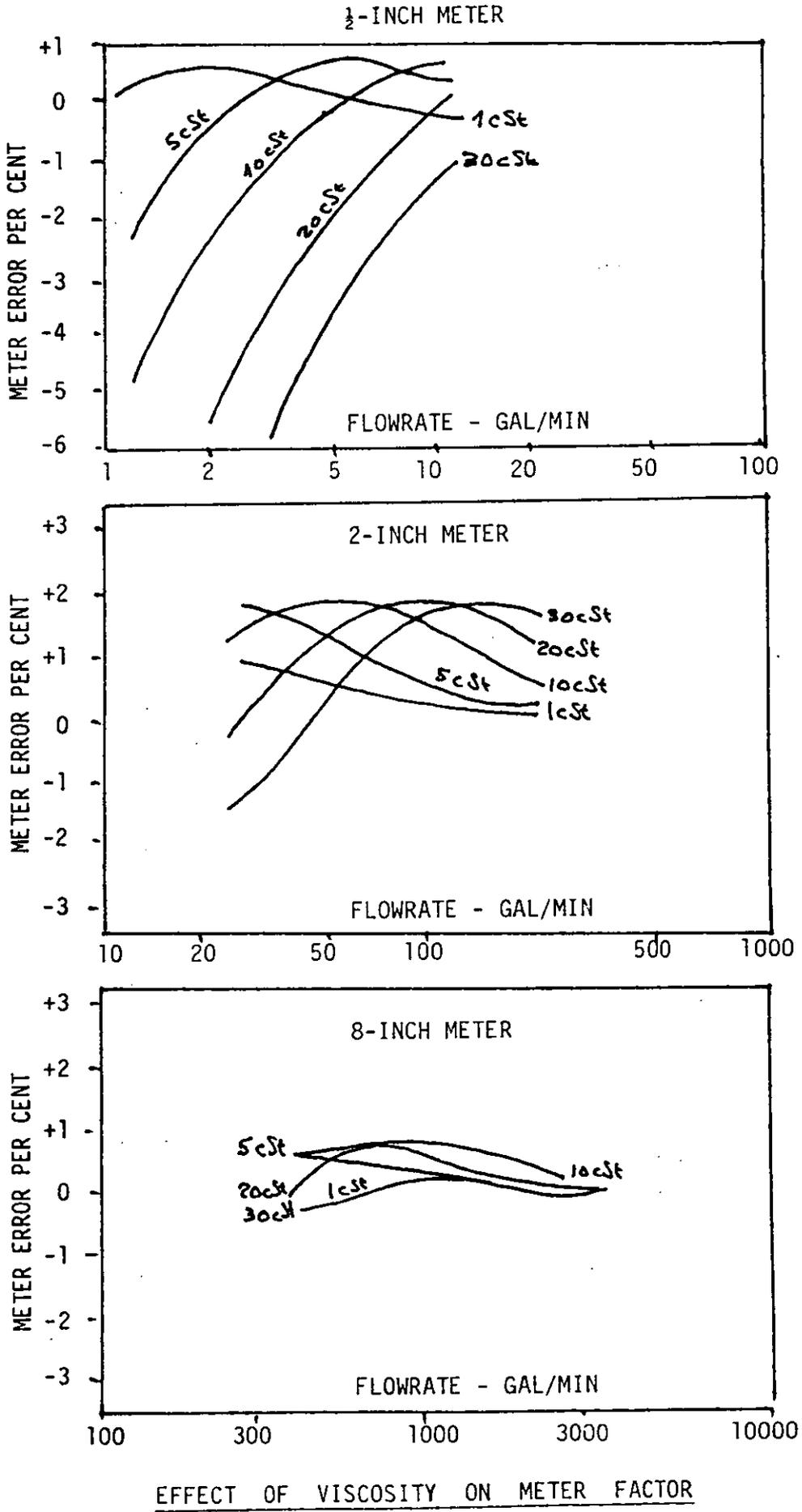


FIG. 2 TURBINE METER CHARACTERISTIC CURVE.



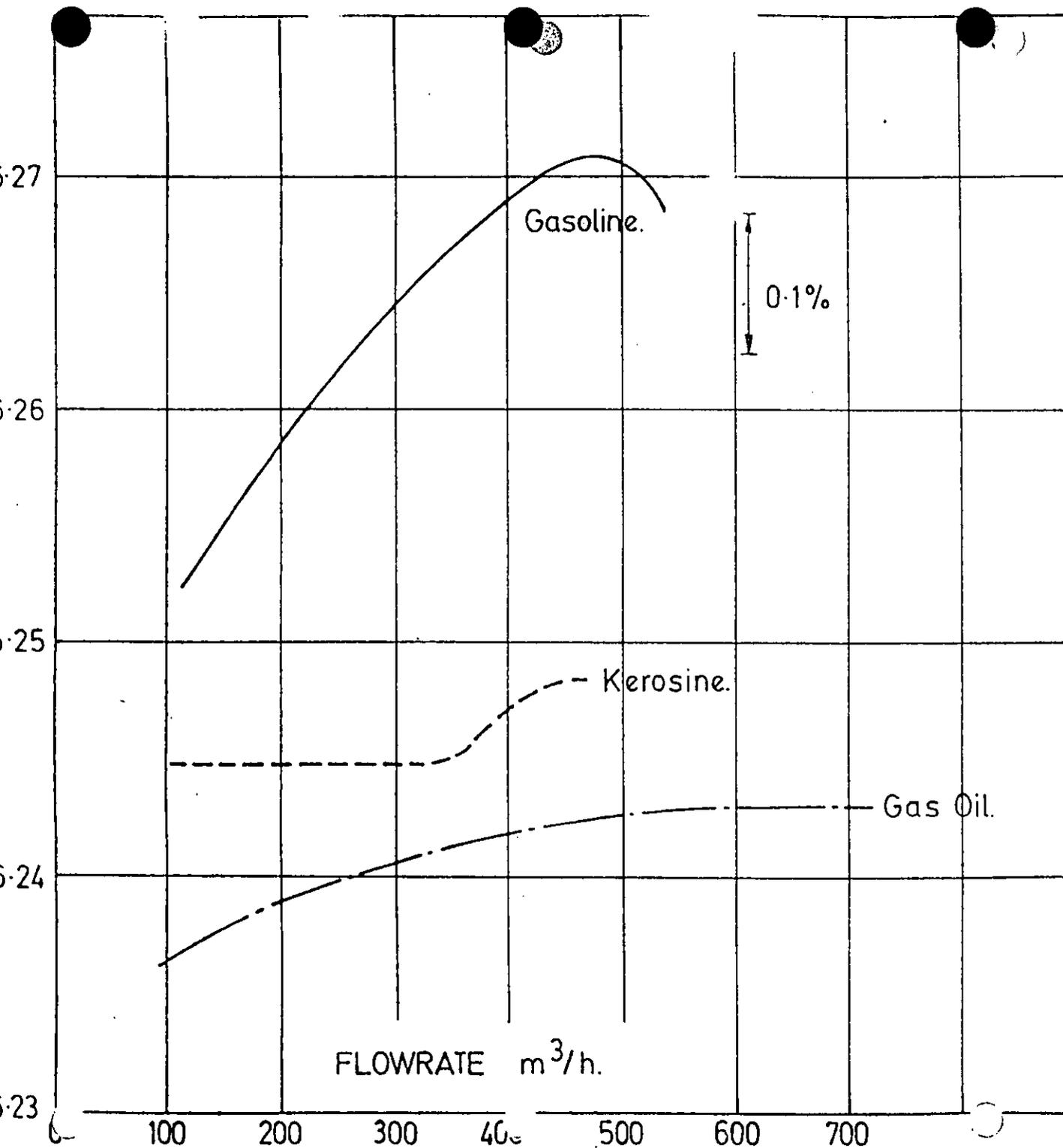
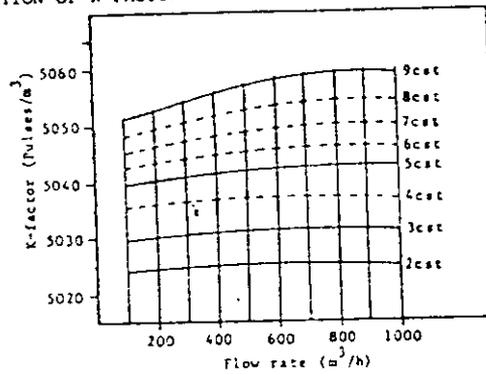
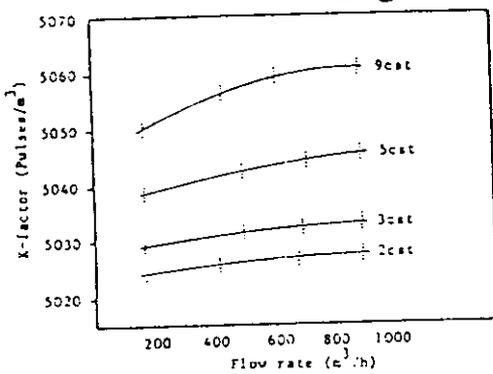


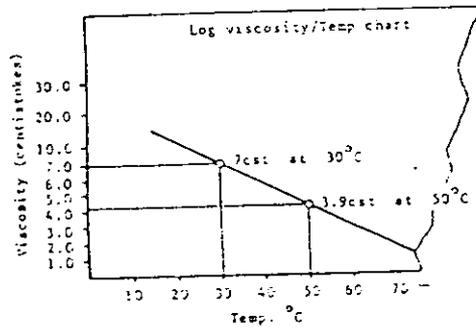
FIG. 4.
SOME TEST RESULTS
OBTAINED IN THE FIELD
WITH A 6-INCH TURBINE
METER & A PIPE PROVER.

FIG. 5 LINEARISATION OF K-FACTOR



STEP 1 - PERFORMANCE CURVES
(Obtained from flow laboratory on various viscosity oils)

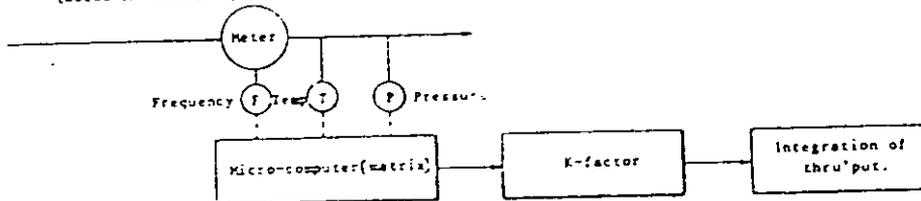
STEP 2. INTERPOLATION OF DATA FOR MATRIX (or best fit curve)



Flow rate (m³/h)	Matrix			5
	K-factor (Pulses/m³)			
	2cst	3cst	4cst	
	82°C	60°C	49°C	
100	5025.1	5029.6	5035.1	
200	5025.5	5030.1	5	
300	5026.1	5031.0		
400	5026.6			
500	5			

STEP 3 - CRUDE OIL VISCOSITY/TEMP. RELATIONSHIP
(Based on laboratory tests)

STEP 4 - CONSTRUCT MATRIX FOR MICRO-COMPUTER



STEP 5 - METER INSTALLED ON SITE

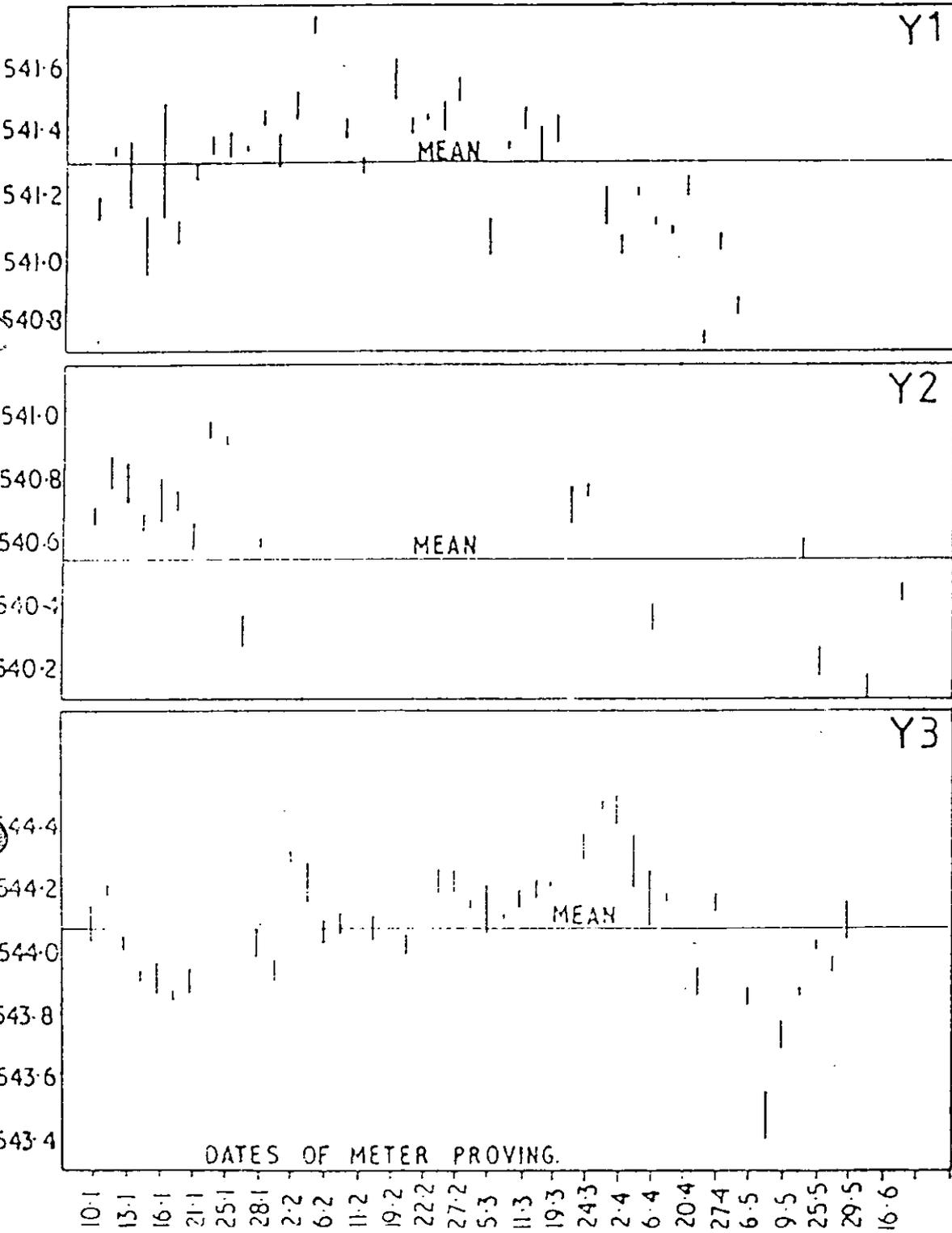
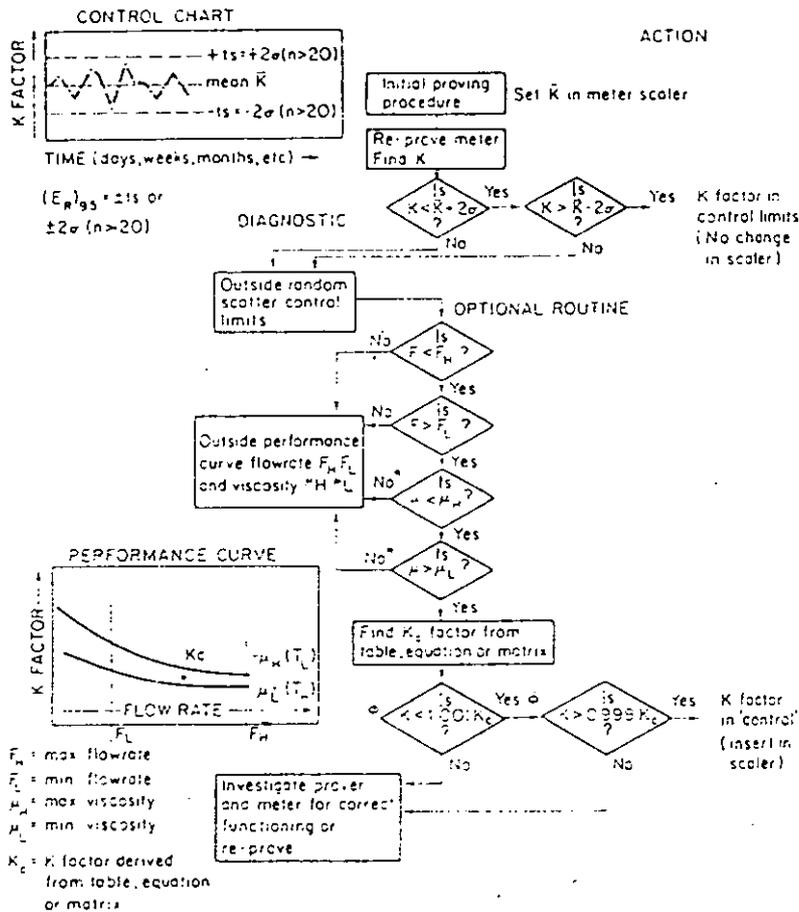


FIG. 6. CONTROL CHARTS FOR METERS Y1, Y2 AND Y3 SHOWING



NB * For one grade, temperature can be used instead of viscosity
 & This limit of ±0.10% can be varied

FIG. 7 Statistical control---Meter proving using control charts.

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"MEASUREMENT OF GAS AND LIQUIDS"
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7 - 10 JUNE, 1982

FLOW COMPUTERS AND TRANSMITTERS IN ORIFICE
METERING SYSTEMS FOR GAS

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ELF AQUITAINE NORGE A/S

FLOW COMPUTERS AND TRANSMITTERS IN ORIFICE METERING SYSTEMS FOR GAS

INTRODUCTION

In this lecture I have chosen to illustrate the use of flow computers and transmitters in orifice metering systems for gas by taking the new metering system on the Frigg field as an example. The basic layout of the system is shown Fig.1 and the architecture of the system is selected with both consideration to the general layout of platforms on the field and the wish to have the metering system available without degradation in metering accuracy during a number of fault conditions.

The general layout of the Frigg field is shown in Fig.2. The total number of platforms on the field is five. The gas from the wells platforms CDP1 and DP2 is transported to treatment platforms TP1 and TCP2 respectively via 26 inch pipelines. On production platforms TP1 and TCP2 the gas is treated to achieve the required water dewpoint specification where also the metering takes place before the gas enters the 32 inch pipeline for transportation to the St.Fergus terminal in Scotland.

This information is hopefully adequate as background information and I will now continue to talk about the details of the system.

Referring back to fig. 1 we can see that the transmitters which provide inputs to the flow computers are:

- Density
- Differential Pressure
- Static line Pressure
- Temperature.

In addition to a functional description/special installation requirements of the flow computers and the above sensors, I will also talk about calibration and maintenance aspects. Some of the comments given in this lecture represent ideas generated from my own personal experience and I do not of course consider these to be universal acceptable but rather input to a debate where the aim is to achieve metering system with maximum accuracy combined with minimum maintenance efforts.

FLOW COMPUTERS

The new flow computers on the Frigg field due to be delivered in August this year are made by the english firm Spectra-tek UK Ltd. Elf selecte these particular models due to the flexibility in adapting them to the our special requirements.

- a) 869 R Stream Measurement Microcomputer
- b) 869 V Central Control Microcomputer
- c) 869 V Database computer

The computers are based on a modular electronics system enabling them to be configured for their specified tasks by selecting the appropriate plant interface cards, carrying out the corresponding back-plane wiring and finally writing the application software. The heart in these flow computers is the central Processor unit incorporating the Motorola MC 6809 MPU chip.

From my point of view the integrity or the security of the system is very important when selecting one of the many flow computers available on the market.

In our case the control programmes are ROM based requiring no initialisation or commissioning procedure. This form of memory is highly secure and not prone to the obscure corruption which can, and often does, occur in core based minicomputers. In my view the exclusive use of ROM for the control programmes is essential to the security and reliability of the system.

One should be able to modify the calculation parameters held in RAM, but on-line changes to the control programme are highly undesirable. If necessary, changes in calculations can be implemented off-line and fully tested before being incorporated into the control program.

We shall now take a closer look at the Software Security features which are incorporataed in the machine.

- a) All running totals and other vital data are held simultaneously in three separate registers. During each programme, loop a check routine is performed to ensure that the data is identical in all three registers. If one register of the three should ever disagree with the other two, this is a triple register "partial failure", annunciated as an alarm. Upon alarm acceptance the computer attempts to correct the erroneous version. In the event of total triple register disruption, ie none of the three registers agree, then a triple register total failure is annunciated, which cannot be corrected automatically.

This feature means that in the absence of tri-register alarms, confidence in data security can be justifiably high, and should data corruption ever occur, then this is annunciated immediately.

- b) During each programme loop a check routine will be executed to verify the logical and arithmetic operation of the CPU itself.
- c) A creeping RAM monitor routine is used to verify the read/write operation of every location in the RAM memory. The routine is run as a background activity in which taking each location in sequence, the microprocessor stores the memory location data in a register, then writes and reads test patterns of data into the location before replacing the data and moving on to the next location. Any location failing the test raises an alarm, and the instrument indicates the identity of the faulty card.
- d) A ROM integrity check routine is incorporated based on the use of checksums in every ROM chip.
- e) An independent hardware timer (Watchdog) is provided to detect collapse of the system in terms of orderly programme execution.
- f) A power supply monitor routine is used to detect imminent supply failure enabling the machine to secure data and shut down in an orderly manner.
- g) All communication between microcomputers relies on rigorous interactive protocol to ensure high security.

Another very important feature of any metering system is availability. It is not acceptable to have the metering capacity reduced for long periods of time in order to perform maintenance and repairs.

Referring back to figure 1 it is seen that the metering concept is based upon the idea of decentralisation. The task of computing the flow rate, monitoring process conditions etc. is done as close as possible to the process elements themselves. Should an error situation occur, the influence on the metering capacity of the system is minimal.

As seen from the figure each meter run is equipped with a dedicated micro computer with complete autonomy with respect to its tasks.

Typical inputs to the machine are differential pressure, static pressure temperature and density. From these inputs volume and massflowrate are calculated on a continuous basis.

Two bidirectional serial data ports are provided on each of the stream micro computers to allow independent communication with the central control room machine CCM (located in the interface room on TP1), and with the CCM on TCP2 via data highway no.2. With this set up the information from the stream machine has alternative paths should either of the CCM fail.

During normal operation the micro computer located in the control room on QP (the data base computer DBM) controls the operation of the entire system.

From the keyboard on the DBM the control room operator addresses and initialises the stream machine directly. The operator can also ask for a display of data, working constants etc. presently used by the stream machines.

During normal operation the DBM is the master and communicates with CCM1 or CCM2 which again is responsible for down-loading the information to the stream machines.

Other tasks of the DBM are:

- To form a constantly updated database.
- To check the integrity of the CCM.
- To format and print the daily production log.

Should during an abnormal situation, the DBM cease to operate, its function of being a master will be taken over by the CCM.

Another important task of the CCM is to communicate with a dual cassette recorder containing all the data pertinent to the stream micro computer. All the data stored on tape can be displayed on the CCM's data screen. Should any of the data need to be changed this can be done using the keypad on the CCM. Dual cassette recording is provided so the original data is kept on the tape while the technician is editing the other.

The alarm and reporting function of the DBM is in this mode controlled by the CCM.

Since the CCM has 100% back up with respect to its tasks, normal production reporting will be provided in most situations because the changes that both CCM machines fail at the same time are considered to be minimal.

Another feature of the CCM is that data entered by the operator during for example a change of orifice are automatically compared to a pre-programmed list of permissible values stored in the CCM. Further transmission of data to the relevant SMM will only take place if the entered values agree with allowable orifice diameters and bore numbers.

Finally, the CCM checks the integrity of the stream machines by monitoring the SMM self check alarm flag in addition to the coherence of the transmitted data itself.

The back-up to the stream machine is considered to be made up by the large amount of meter tubes. The metering capacity on the Frigg field is approx 100 MMSCM while the max production is in the order of 65 MMSCM.

Therefore should a stream machine fail, immediate alarm is raised and the operator shut down that particular stream from the control room.

Finally, the integrity requirement to a fiscal metering system is very stringent. Therefore the parameters available for operator entry are limited to:

- Operating mode (Meter, Change orifice, Calibration or Initialisation).
- Orifice number and orifice bore diameter.

Other data may also be over written, but it is offered an extra level of security i.e security code must first be entered successfully before data can be entered in conjunction with the data entry key switch.

In this manner, system integrity is satisfied.

CALIBRATION / MAINTENANCE

Another great advantage with modern digital micro computers is that self checking facilities as described earlier are provided. Therefore, the traditional maintenance aspect as experienced with "old" analogue computers has dissappeared.

However, an element which should not be forgotten is the analogue to digital converter. Since a slight error in this unit will cause a systematic error in all parameters, particular attention should be paid to self checking facilities and regular calibration monitoring. The A/D converter to be used on the Frigg field has both automatic zero and span correction. During the auto-zero phase the total on-card offsets are measured, stored and subtraced. This effectively achieves a zero drift of 2 micro V/°C allowing a 50°C ambient shift before one count has been passed.

Span correction is achieved using a very high precision reference unit mounted in the micro computer and forming one of the scanned inputs of the system. The micro computer will scan this known voltage and store the converted digital number in its memory. The precise value of the chosen reference, obtained by commissioning measurement using a certified transfer standard voltmeter, can be stored in the computer memory using the key pad facilities on the computer front. Subsequent readings of all other channels will be ratioed by software to account for the error between the keypad entered value and the value measured by the ADC.

DENSITY MEASUREMENT

Principle of operation

The density transmitter which will be used on the Frigg field is manufactured by Solarton. The type 7811 has been chosen and is specially designed for high static pressure operation.

The operation principle of the transducer is shown in Fig.3. The transducer sensing element consists of a thin cylinder which is actuated so that it vibrates in a hoop mode at its natural frequency.

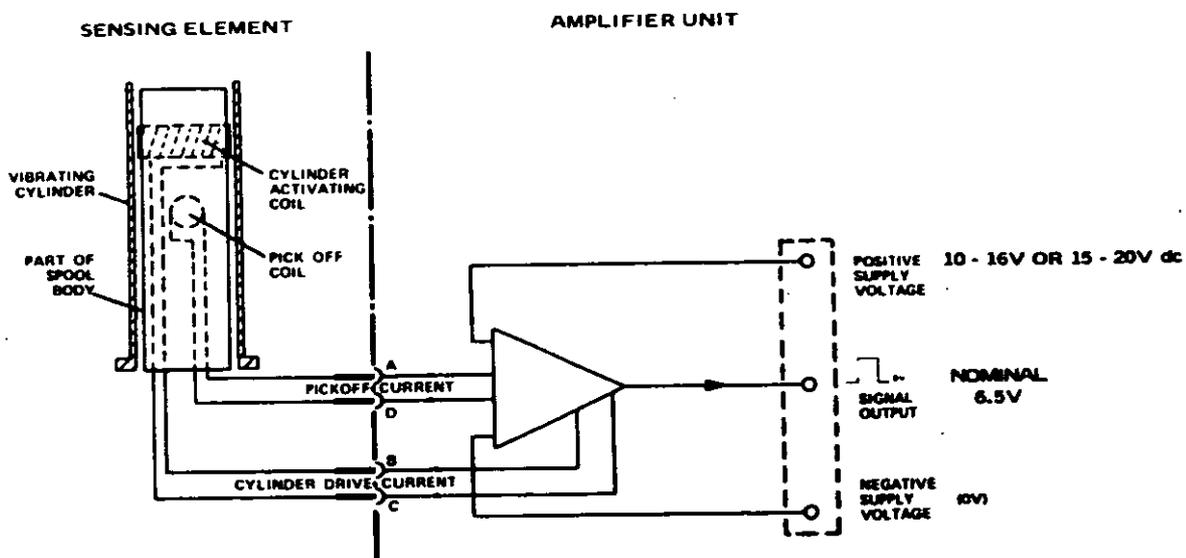


FIG. 3 Schematic Block Diagram of Transducer Circuit.

The gas is passed over the inner and outer surfaces of the cylinder and is thus in contact with the vibrating walls.

The mass of gas which vibrates with the cylinder depends upon the gas density and, since increasing the vibrating mass decreases the natural frequency of vibration, the gas density for any particular frequency of vibration can be determined from the formulae:

$$\rho = K_0 + K_1 T + K_2 T^2$$

$K_0, K_1, \rightarrow K_2$ CONST.

$\underbrace{\hspace{10em}}$
PERIODIC FREQUENCY

INSTALLATION OF DENSITYMETER

Ideally the density transducer should be located within the flowing gas adjacent to the reference volume metering plane. In ISO 5167 the metering plane is defined as the plane of the upstream pressure tapping point (ref section 3.4).

It is of course not permissible to install the density transducer in the pipe directly upstream of an orifice plate, because this would disturb the gas flow pattern at the orifice and cause errors in flow measurement.

So the task from an engineer's point of view will be to install the density meter elsewhere and minimise the difference in operating conditions between the density transducer and those at the volume metering plane.

Prior to making the final selection, efforts should be made to minimise the errors associated with the following points:

- a) Effects on the density transducer itself;
for example, temperature and pressure coefficients, flowrate effects
attitude, accuracy of calibration, effects of velocity of sound,
etc.

- b) Temperature and pressure differences between the gas at the metering plane and the gas in the density transducer.
- c) Adequate filtration / conditioning to prevent dirt or condensation from causing maloperation of the transducer.
- d) Unregistered gas which passes through the density meter but not the flow meter.

When using a Solarton transducer, the final installation must be designed in order to obtain a representative gas sample stream. This consideration excludes using the upstream flange pressure point as origin for the slip stream. A complete analysis of the flow pattern at this point will not be provided here, but the sudden increase in gas velocity in front of the orifice plate will affect the components in the gas differently thus making the flange taps point unsuitable for obtaining a representative sample

With this in mind we are left to consider the two alternatives shown in Fig.4A and 4B respectively.

In both alternatives the density transducers are located in a thermowell thus making sure that the gas in the density meter has the same temperature as the gas in the main line. However, in figure 4B the sample point is located upstream satisfying the free straight length requirements of ISO 5167, but a pressure drop will occur along the sample pipe to the density transducer. Therefore the pressure in the density will be different to the pressure of the metering plane.

At Frigg it will be necessary to install an additional filter in the sample stream thus increasing this pressure drop. But the most severe limitation with this method is that it is difficult to account for the pressure drop.

In the majority of the offshore installations, the static line pressure is 100 - 200 Bar. Therefore equipment is not available to measure the very much smaller pressure drop in the sample line. Hence, it is not known how to account for this in the massflow calculations.

An additional disadvantage with this method is that the slip stream is not registered by the flowmeter and thus give rise to a systematic error which however small should be accounted for. An estimate of this error may be obtained by referring to the section on drain holes in British standard BS 1042 or similar documents.

The alternative arrangement shown in Fig. 4A provide the same good temperature equalisation as fig. 4B. The sample point in this case is taken downstream of the orifice carrier and the gas slip stream returned to the downstream flange tap. In this arrangement, the length of the sample return line is very short. Therefore the additional pressure drop may be neglected and the pressure in the density transducer is equal to the pressure at the downstream flange tap P2. Additional filtering installed in the sample line will in this case not interfere with the density measurement.

However, the density is determined at the downstream flange tap and the metering plane is as stated previously the upstream flange tap. Using the above arrangement require the basic mass flow equation to be modified, this means that the expansion coefficient at the downstream flange tap must be calculated.

For a given flow meter and flow rate the position where ρ is measured and the corresponding ϵ must give the same answer.

$$W = CZE\epsilon_1 a \sqrt{h\rho_1} = CZE\epsilon_2 a \sqrt{h\rho_2}$$

$$\text{giving: } \epsilon_1 \sqrt{\rho_1} = \epsilon_2 \sqrt{\rho_2}$$

where: ϵ_1 = expansion coefficient at the upstream flange tap.

ρ_1 = Density at the upstream flange tap.

ϵ_2 = Expansion coefficient at the downstream flange tap.

ρ_2 = Density at the downstream flange tap.

If the small differences in density can be considered directly proportional to the small pressure difference.

(iso-thermal approximation)

Giving: $\rho_1 \propto P_1$, $\rho_2 \propto P_2$

Where h = differential pressure across orifice

then one can write: $\epsilon_2 = \epsilon_1 \sqrt{P_1/P_2} = \epsilon_1 \sqrt{\frac{P_1}{P_1 - h}}$

The above method is described in the new draft proposal for the British standard on orifice measurement Part II of BS 1042. In addition AGA 3 provide a formular for calculating ϵ_2 at the downstream pipe tap,

$$\epsilon_2 = \sqrt{1 + X} - \frac{BX}{\sqrt{1 + X}}$$

$$\text{and } B = 0,333 + 1.145 (\beta^2 + 0.7\beta^5 + 12\beta^{13})$$

Note that in this case for $X = h/p$, h is the differential pressure form pipe taps. If h is measured with flange taps the following correlation can be used

$$\frac{h \text{ (pipe taps)}}{h \text{ (flange taps)}} = 1 - 1,08\beta^2$$

CALIBRATION

The procedure to be used on the Frigg field when calibrating the Solartron transducer is still being considered, and I will therefore just offer some general comments on the matter. Normally the calibration of the Solartron transducer is carried out off-line and it is necessary to correct the basic calibration coefficients to take into account the difference in velocity of sound between Frigg gas and the calibration gas.

However, our experience so far is that once the Solartron has been calibrated it will maintain calibration for a very long period of time.

As a continuous monitoring of the density transducer one additional function of the stream micro computer is to calculate density from P & T measurement and compare the theoretical value to the measured density. An alarm will be given if the difference exceed a preprogrammable limit. It is our intention that this facility will enable us to maintain the instrument within the desired accuracy limits.

DIFFERENTIAL AND STATIC PRESSURE MEASUREMENTS

The sensors which are used for these measurement on the Frigg field are the well known transmitters form Rosemount and Foxboro.

Rosemont 1151 HP is used for differential pressure measurements and Foxboro E11GH is used for the static pressure measurements.

Calibration

The static pressure transmitter is calibrated on line using a deadweightster from Chandler engineering Ltd.

The instrument is rugged and well suited for the accuracy required. (< 0.3%).

However, the calibration of the dp cells is more elaborate as there is no equipment available with the required accuracy suitable for field use.

The sensor has therefore to be disconnected and brought to a place where it can be calibrated under controlled environmental conditions using a Degranges & Huot dead-weight tester.

It turned out that the only place suitable for this activity was the living quater platform QP. Even there under difficult weather conditions the vibrations in the structure is such that it is impossible to carry out the calibration.

TEMPERATURE MEASUREMENT

Gas temperature is measured by a Platinum Resistance Thermometer manufactured to BS 1904 Grad 1 having a resistance of 100.0 Ohms at 0°C and a fundamental interval of 38.5 Ohms.

The PRT will be connected by a four-wire, screened arrangement to the metering cubicle terminals and within the cubicle to the SMM.

Temperature measurement is self powered, no external power supplies shall be required. Two of the four PRT wires and the PRT element itself form a current loop carrying some 2.8mA circulated by a current source which forms part of the ADC. The remaining two wires channel the p.d developed across the PRT (caused by the 2.8mA flow through the resistance) into the voltage sensing input of the ADC which has a very high input resistance resulting in a negligible voltage drop in the cabling. The ADC ratios the current magnitude to sensed voltage and yields a count which proportional the PRT resistance yet insensitive to cabling resistance therefore accurate.

The SMM software will convert the measured PRT resistance to a temperature reading using the Calender Van Deusen relationship.

Accuracy: + 0.1^oC in resistance measurement terms, but
+ 0.2^oC taking transducer interchangeability errors into
account.

Calibration

The calibration of the temperature measurement loop is performed by simulating resistance values from the field according to the British Standard BS 1904. Using a high precision certified resistance bridge. The simulated values can then be compared to the printout from the CCM or DBM micro machines.

Although the platinum elements themselves are more stable than thermo-elements, their temperature / resistance relationship do change with time. Exposure to high temperature will accelerate his change.

Very often this is overlooked and systematic errors in the order of 0,5^oC - 1^o can occur if recertification of Pt 100 elements are not performed as part of the regular maintenance plan.

The frequency of this recertification can vary from one installation to another depending on how high temperature the elements are exposed to.

But recertification at 12 months intervals could be used as a starting point.

An easy check which should be performed on the field on a regular basis is to insert the Pt 100 element into a thermo bottle containing mixture of ice / water, noting the temperature measured by the stream microcomputer.

Changes in the Platinum characteristics can in this manner be detected at any early stage.

Stavanger 14.5.1982

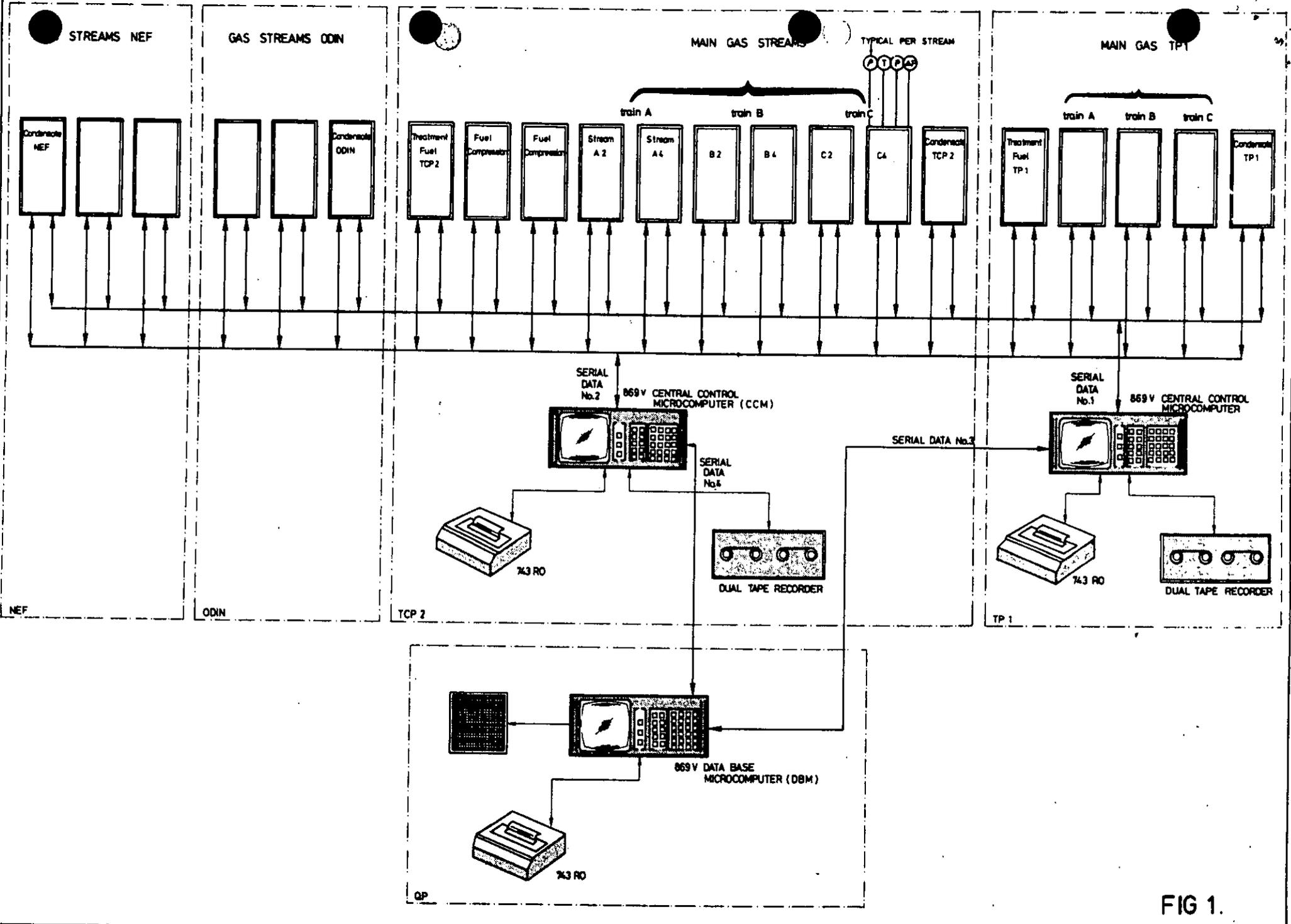
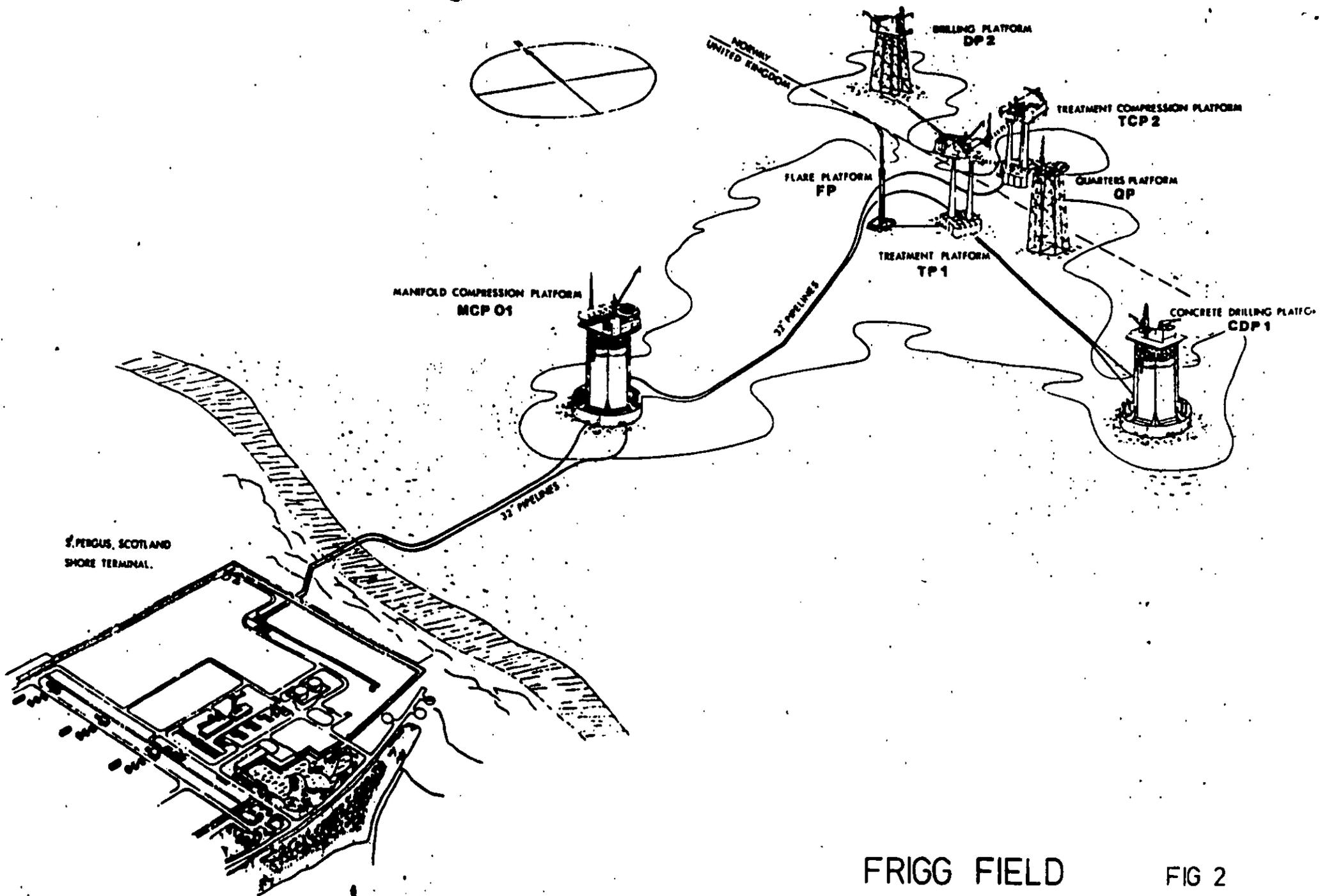


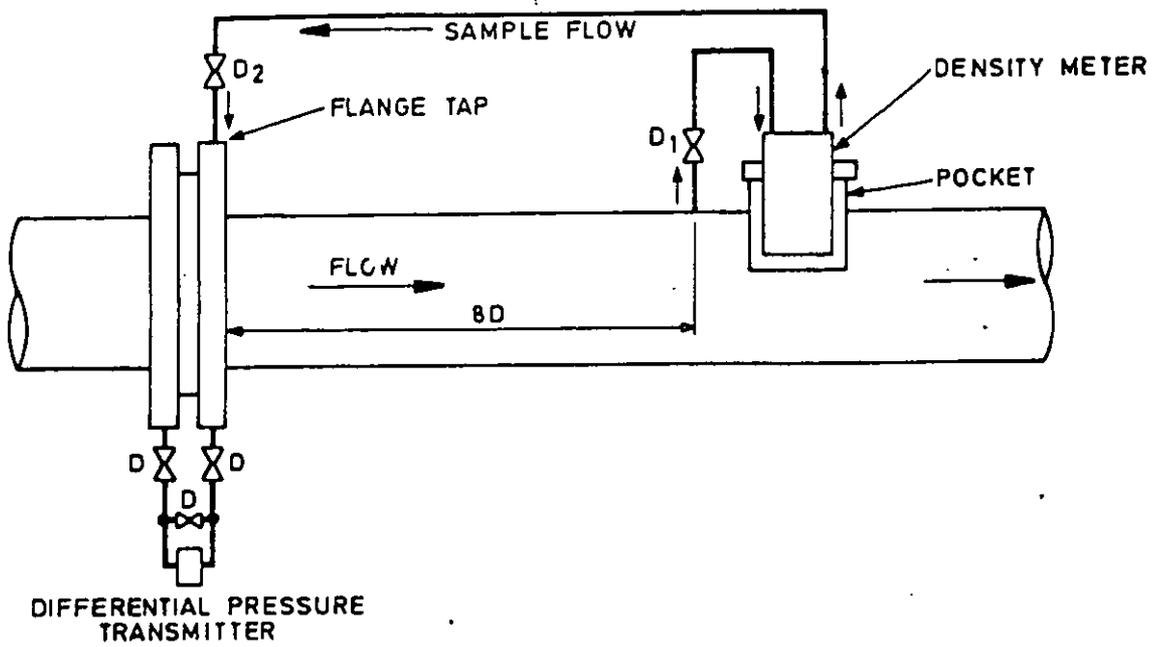
FIG 1.



FRIGG FIELD

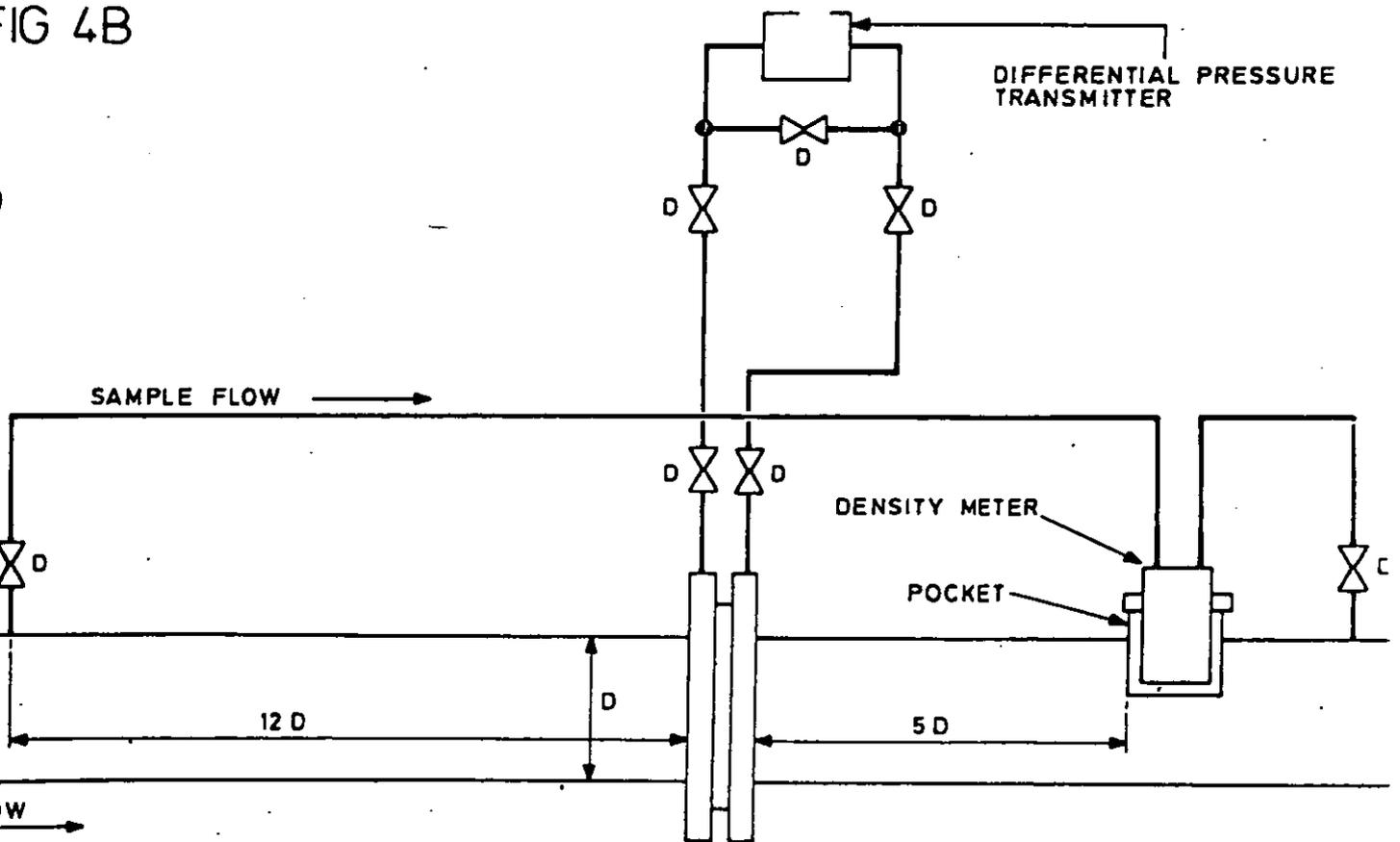
FIG 2

FIG 4A



INSTALLATION OF GAS DENSITY METER ON AN ORIFICE PLATE SYSTEM USING THE PRESSURE RECOVERY METHOD

FIG 4B



BY-PASS INSTALLATION - POCKET METHOD

MEASUREMENT OF GAS AND LIQUIDS

June 7-10, 1982

Rogaland Regional College
Stavanger

MECHANICAL DISPLACEMENT METER PROVERS

- 2 - en

Lecturer: Peter A.M. Jellfs
Technical Director
Moore, Barrett & Redwood Ltd.
England

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MECHANICAL DISPLACEMENT METER PROVERS

1. INTRODUCTION

The conventional pipe provers were first introduced into crude oil measurement for custody transfer purposes in the Middle East in the early 1960's. Since then they have become the recognised calibration standard for all fiscal metering systems throughout the world. Recent improvements in the detection of displacers and pulse interpolation techniques have made it possible to design "compact" provers with greatly reduced swept volumes.

The main factors which influence the design and repeatability of both the conventional and compact are considered in this paper.

2. CONVENTIONAL PROVERS

a. Principle of Operation

The conventional pipe prover uses an oversized rubber or plastic sphere, which is filled with water, in commercial grade industrial pipe. The internal surface of the pipe is usually coated with a baked-on phenolic resin to protect it from any corrosive elements in the crude oil or products. The advantage of the rubber sphere is that it automatically changes its shape to follow the varying contours of the pipe as it moves between the detector switches. The over-sizing of the sphere (2 to 4 per cent larger than the internal pipe diameter) ensures that it acts like a squeegee and prevents by-pass leakage. (See figure 1).

b. Repeatability

The repeatability of the pipe prover is mainly a function of the repeatability of the detectors in establishing the horizontal location of the sphere or displacer. (See figure 5).

Most of the provers used for fiscal metering purposes have mechanically operated micro-switches for detecting sphere location. A distance between detectors of 20 metres is usually chosen in order to achieve a repeatability (expressed as the range) of 0.05% for 10 consecutive runs when proving a turbine meter on crude oil.

Recent experience with piston provers using two proximity type detector switches at 10 metres apart have indicated that on a Round Trip (Forward plus Backward runs) a range of 0.05% can be achieved for 5 consecutive proving runs. (See Figure 2).

The sphere hardness can marginally influence the repeatability of the prover but not sufficiently so that the swept volume can be significantly reduced.

The use of pairs of detectors in place of one detector at either end of the prover has many advantages:

- The repeatability of the detectors can be monitored.
- The mean of the two swept volumes, individually calibrated, can be used to improve the uncertainty. (Fig.6).

c. Reduction in Length

One feature of the conventional prover was the long distance between the launching chamber and the first detector. This was to ensure that when the sphere was launched, it did not arrive at the first detector before the main flow valve was sealed.

This run up distance was virtually eliminated by the use of a launching system which used a ram for holding back the sphere until the main flow valve was fully sealed. (See figure 7).

d. 4-Way Valve

A major item of cost has always been the 4-way valve which represents as much as a third of the total cost of the pipe prover.

The main reason for this cost is that the valve has to be designed so that the integrity of the seals can be monitored during a proving run. Also special consideration has been given to ensure that the seals are not damaged during the cycling of the valve. Most valves are designed so that the seals are withdrawn before rotating. Two well known valves raise the plug containing the ports vertically before rotating. (See figure 8).

Any leak across the seals can be observed by monitoring the pressure in the valve casing when it is in the fully seated condition.

3. COMPACT PROVERS

a. Principle of Operation

The swept volume of a pipe prover can be reduced by using a piston and a rod with special detectors. The pistons slide in precision machined cylinders with lip seals around the periphery of the piston. The piston rod slides through a seal in the end chamber and the detector switches are situated on the rod outside the prover barrel. These detectors have such a high resolution that the distance between them can be reduced from the usual 20m_{ns} with conventional provers to less than 1m with a corresponding decrease in the swept or calibrated volume.

b. Non-Linearity of the Meter

As the swept volume is decreased the inter-rotational non-linearity (IRL) of the meter being proved becomes significant.

The linearity of the turbine meter is dependent on the evenness of the spacing between adjacent magnetic points in the rotor rim (or shroud) or tips of rotor blades. The IRL is sometimes defined as half the difference between the maximum and minimum time periods between adjacent blades during one revolution. (See figure 9).

In France a piston prover with two rods is used for proving small nutating disc type displacement meters on LPG. As these meters are non-linear in their throughput the prover calibration volume - and hence the distance between the external detectors - is so sized that the volume corresponding to the non-linearity represents less than 0.02 per cent of the proving volume (See figure 3).

There are two alternative methods for overcoming the non-linearity problem of turbine and displacement meters with small volume provers. These are as follows:

(i) Multiple Chronometry and Pulse Interpolation

Pulses generated by the meter are fed to a signal processor which times precisely each individual pulse during one complete cycle or revolution of the meter. Also the processor identifies each pulse with a particular rotor blade.

By timing the prover in and out gating signals in relation to the meter pulses it is possible to remove the error due to any non-linearity in the meter under test. (See figure 9).

(ii) Prover Pulse Generator

By using a device which converts the linear travel of the piston rod into pulses - sometimes called a linear motion translator - the pulses generated by the meter can be synchronized with the pulses emitted by the prover. A micro-computer can then correct for any non-linearity in the meter pulse output, viz counting the prover pulses between whole revolutions of the meter. (Fig.10).

Also this system would measure accurately the average flow rate during the proving operation.

c. Leakage past the Piston

In the conventional prover the sphere is 2-4% oversize and is highly flexible so that it acts as an energised seal. However, with a piston it is not possible to make an energised seal without introducing the problem of deformation which could contribute to a significant error in the short swept volume. In order to obtain an adequate seal therefore, it is necessary to have a very small interference fit in the precision made prover barrel. This seal could be damaged by fine particles and it is considered necessary by certain authorities to monitor the integrity of the seals during the prover run.

One method is to observe the pressure drop across two parallel seals by means of a suitable transient analyser.

d. Valves

External valves or an internal poppet valve (see figure 4) are used with compact provers for directing the flow but all of these devices should be monitored for leakage during the proving runs.

e. Means for By-Passing Flow

It is essential to ensure that the prover has some means of by-passing flow at the beginning and end of the proving run. Some designs rely on the rapid closing of external valves for directing flow. This can create unacceptable pressure surges due to the need for extremely rapid cycling in order to reduce the run up distance before the calibration section.

One method used by a number of manufacturers of compact provers is to have large end chambers and a means of controlling (externally) the movement of the piston. The 4-way valve or individual valves can then be cycled and the

seating monitored before releasing the piston for the proving run.

The large end chambers also ensure that most of the non-organic particles such as pipe rust etc., do not enter the main prover barrel. (See figure 10).

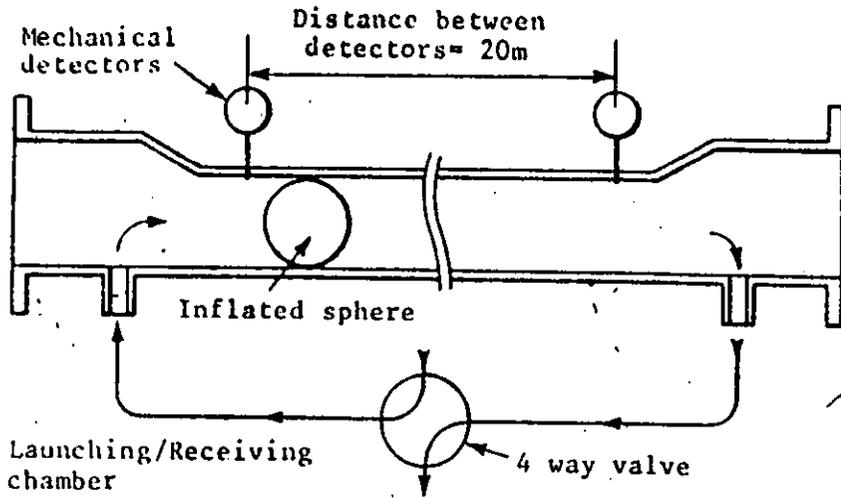


FIG 1. CONVENTIONAL PROVER (BI-DIRECTIONAL)

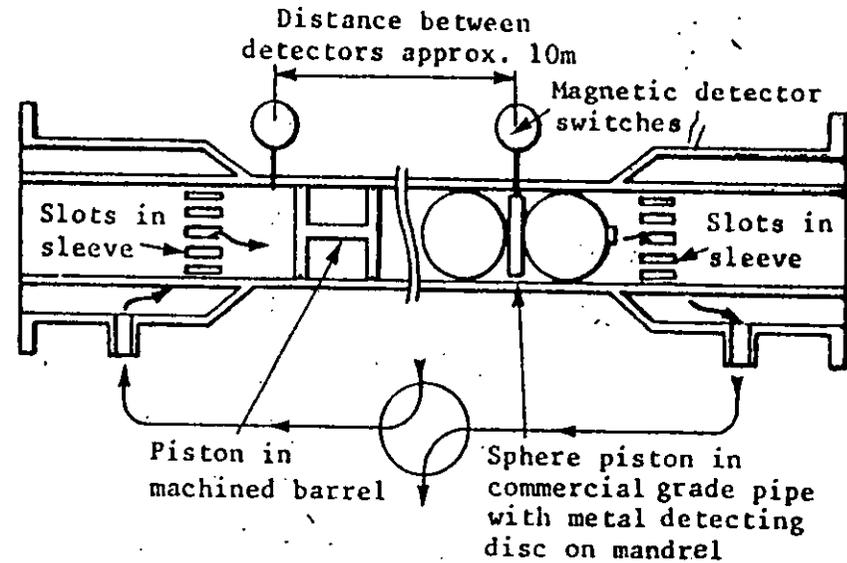


FIG 2. SHORT PISTON PROVERS (INTERNAL DETECTORS)

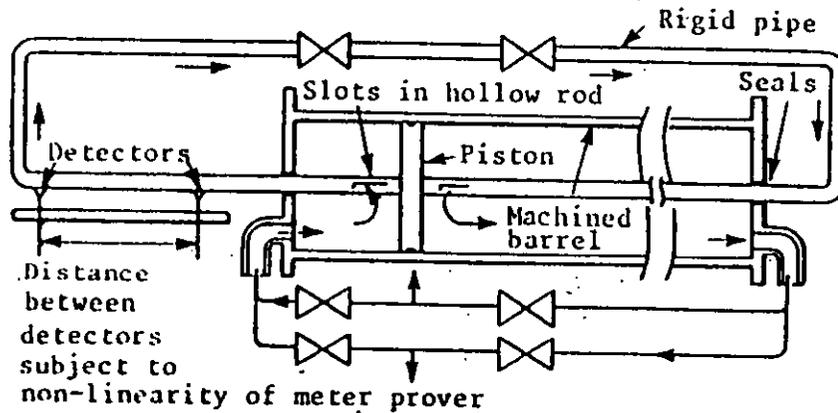


FIG 3. SHORT PISTON PROVER (EXTERNAL DETECTORS AND VALVING)

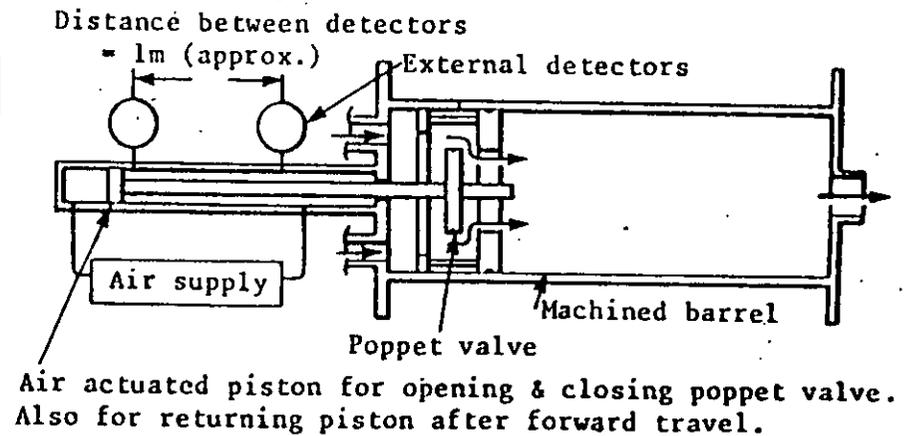


FIG 4. SHORT PISTON PROVER (EXTERNAL DETECTORS AND INTERNAL VALVE)

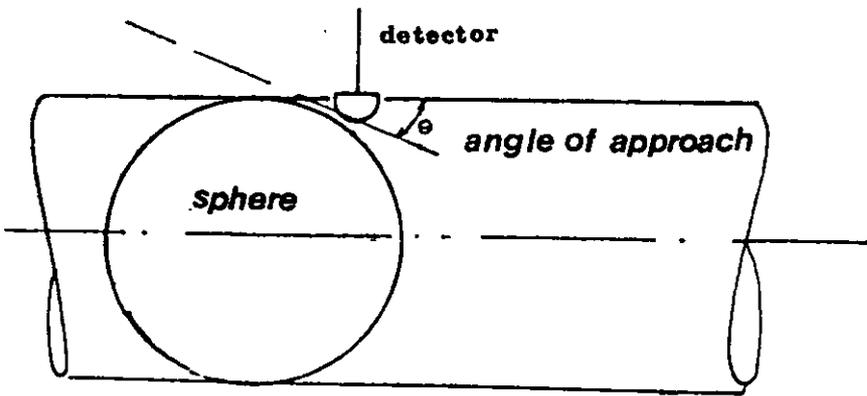


FIG 5
SWITCH REPEATABILITY

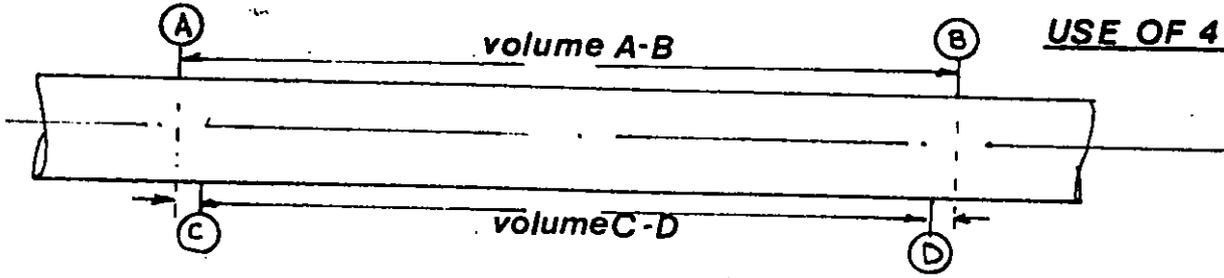


FIG 6
USE OF 4 DETECTORS

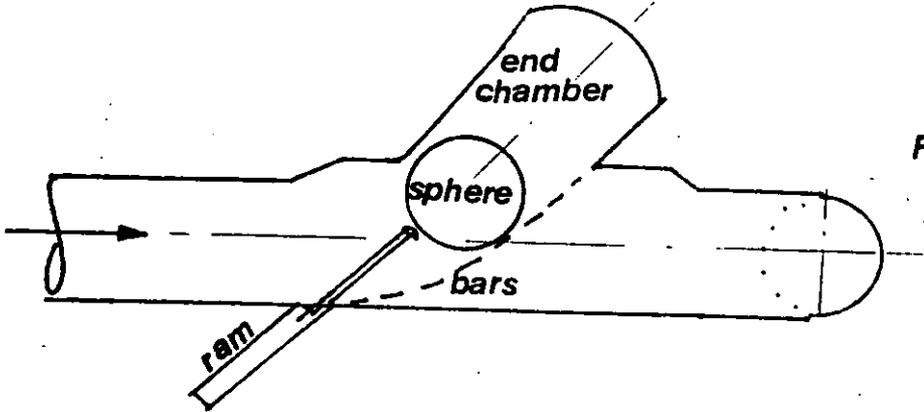
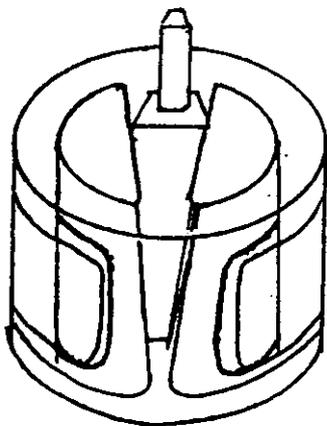
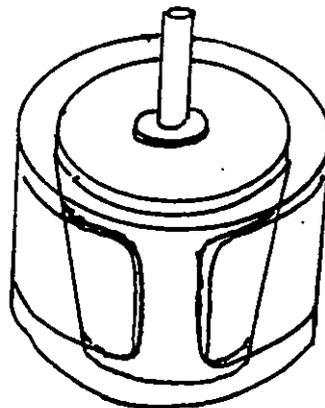


FIG 7
LAUNCH RAM



Split Seal



RISING PLUG

FIG 8 4 WAY VALVE DESIGN

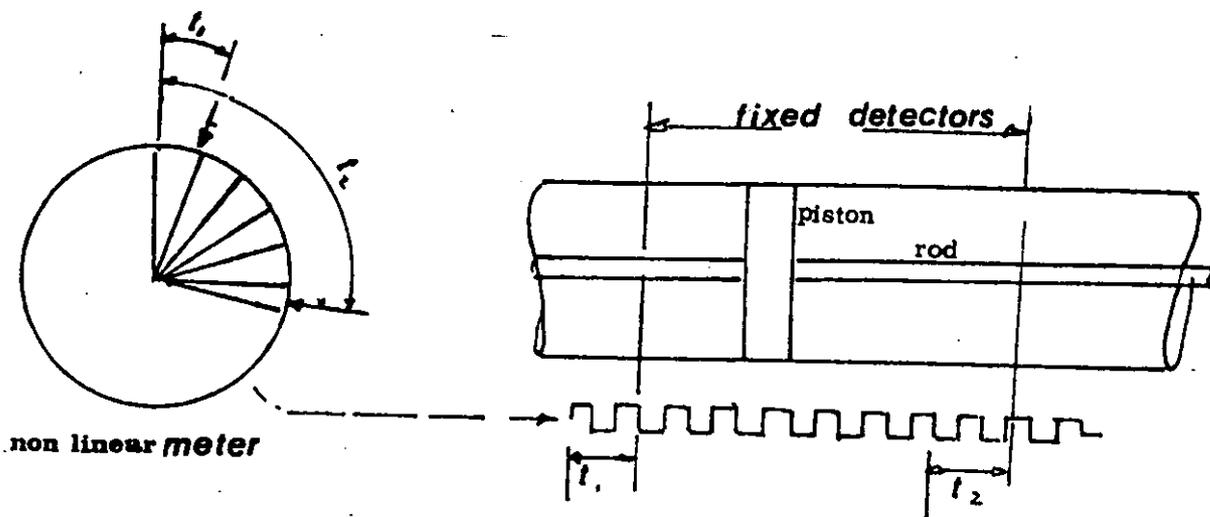


FIG 9 MULTIPLE TIMING METHOD

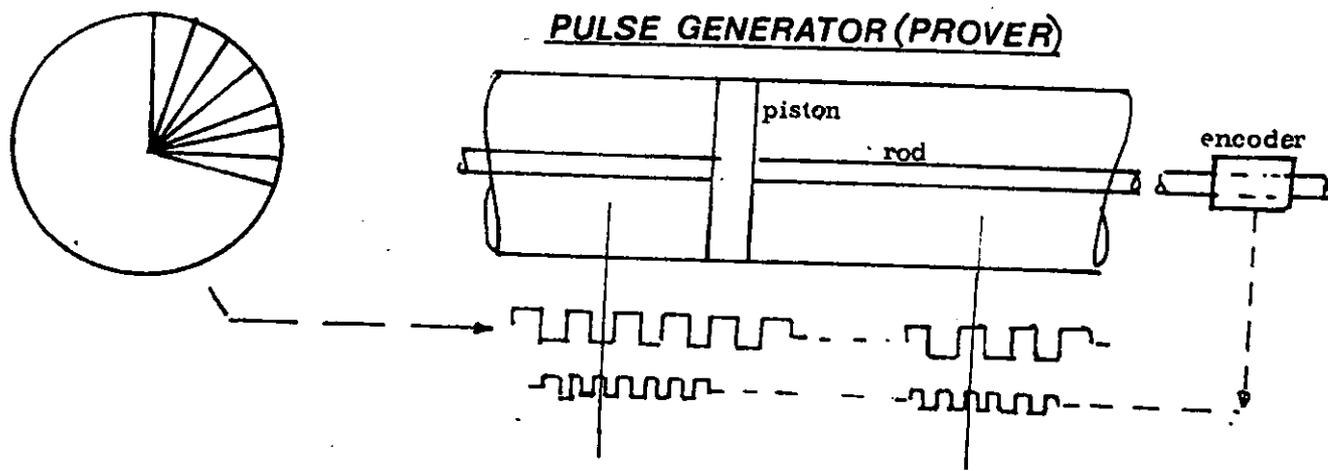
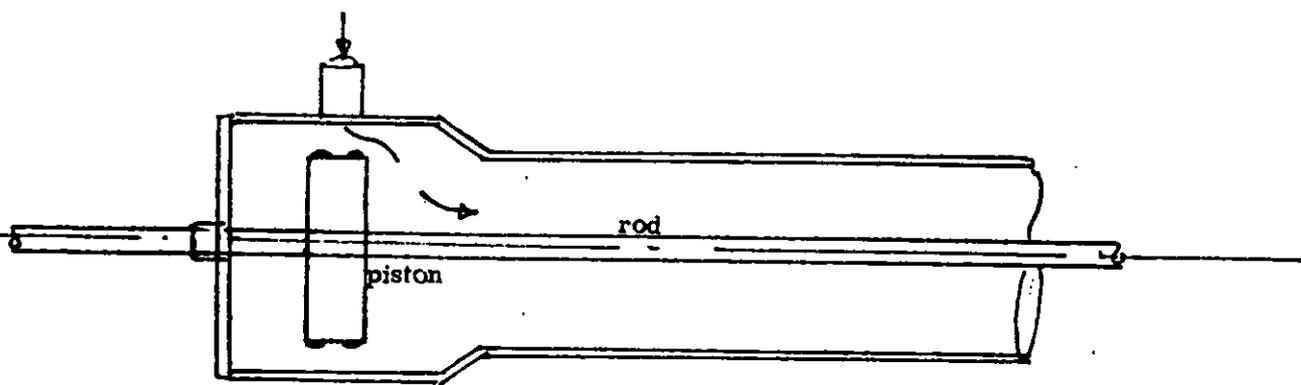


FIG 10 END CHAMBER



MEASUREMENT OF GAS AND LIQUIDS

June 7-10, 1982

Rogaland Regional College

Stavanger

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:

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

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:

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.

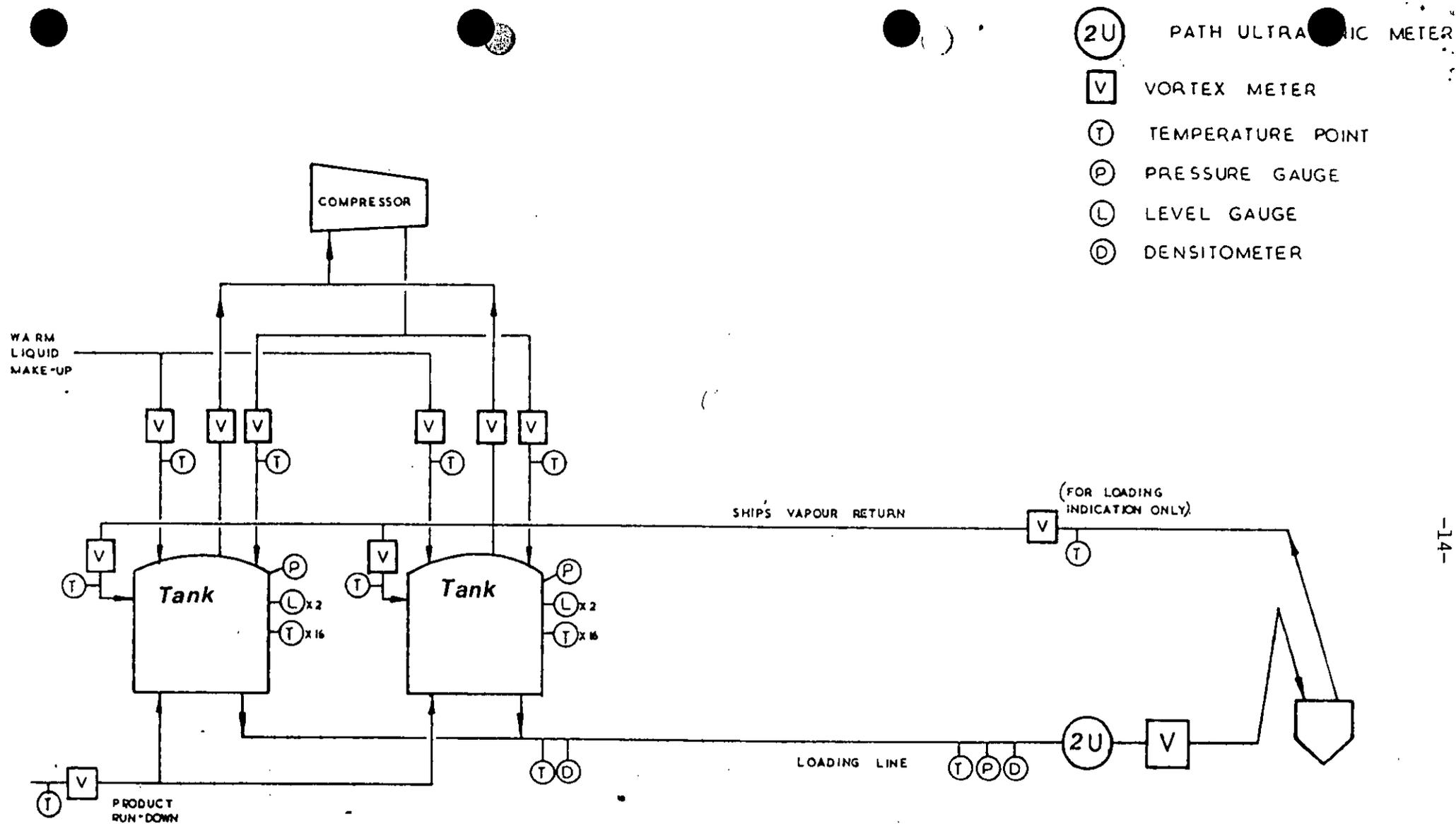


FIGURE 1 SIMPLIFIED SCHEMATIC DIAGRAM ILLUSTRATING THE
STATIC AND DYNAMIC MEASUREMENT SYSTEM.

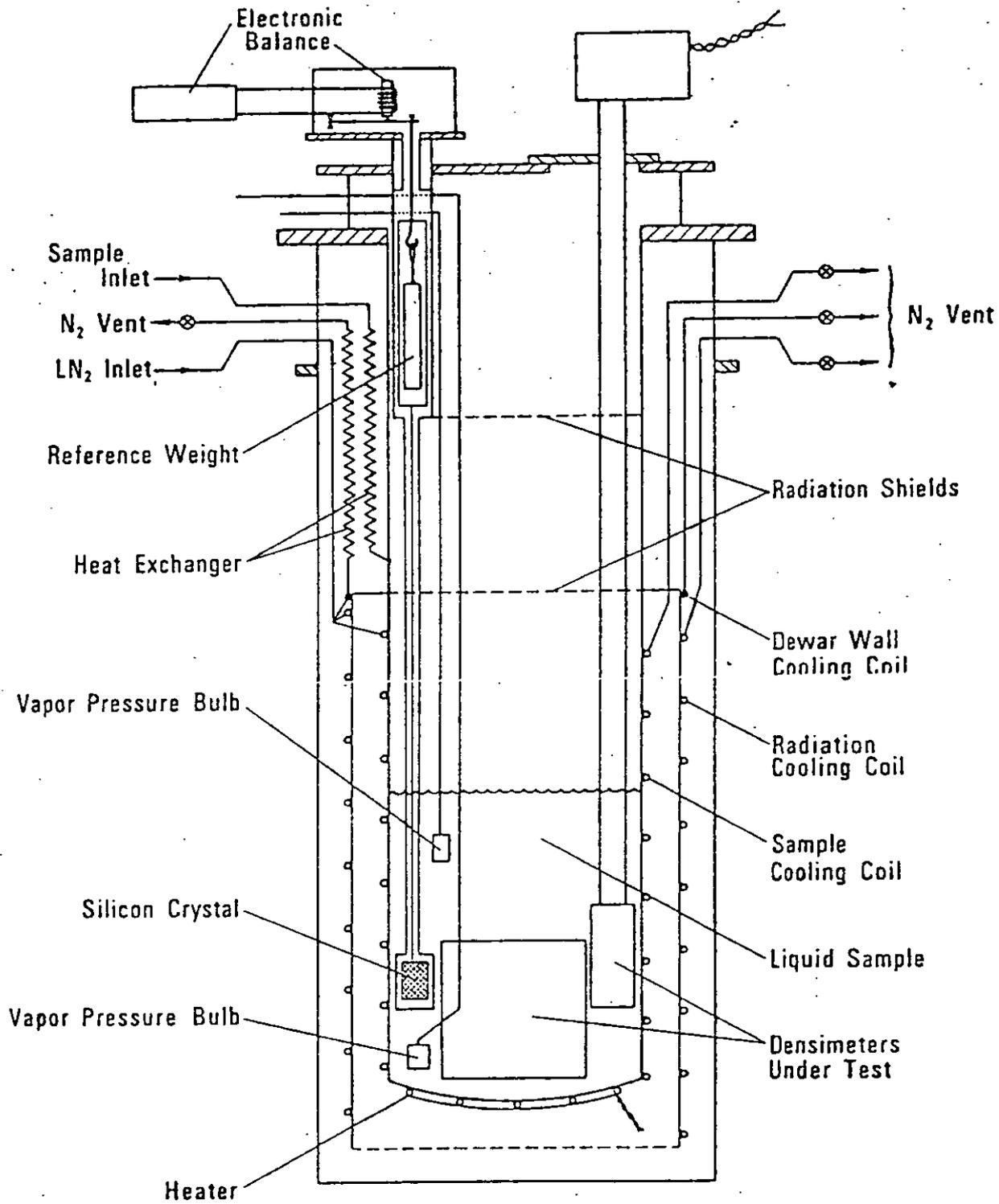
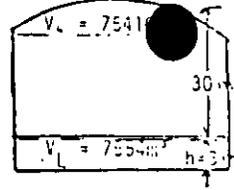
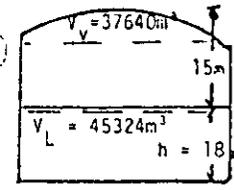
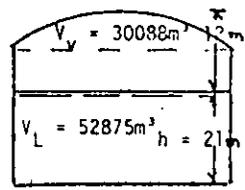
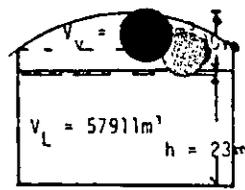


Fig 2 Density Reference System

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

FIG 3



Component Uncertainties	h = 23M	h = 21M	h = 18M	h = 3M
LIQ Level (a) contraction of stillwell for $\pm 5^\circ\text{C}$ $= 5 \times 0.000014 \times 33000 = \pm 2.3\text{mm}$ (3) (b) = repeatability of $\pm 3\text{mm} = \pm 2.1\text{mm}$ $\sqrt{2}$ $= (2.3^2 + 2.1^2)^{1/2} = \pm 3.1\text{mm}$ Volume - Calibration table = $\pm 0.05\%$ Shell Shrinkage = $0.000028 \times 5 \times 100 = 0.01\%$ (4) Densitometric Uncertainty volume = $8 + (29^2 + 6^2)^{1/2} = \pm 38\text{m}^3$ mass (5) = $38 \times 583 = 22.2 \text{ Tonne}$ Density - Calibration = 0.2% (6) - Correction = $\pm 0.1\%$ (6) $(0.2 + 0.1)^{1/2} = \pm 0.22\%$ Total Liquid Uncertainty mass = $(22.2^2 + 74.3^2)^{1/2} = \pm 77.5 \text{ Tonne}$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$ $0.0005 \times 2518 \times 23$ = $\pm 29\text{m}^3$ $0.0001 \times 2518 \times 23$ = $\pm 6\text{m}^3$ = $8 + (29^2 + 6^2)^{1/2} = \pm 38\text{m}^3$ = $38 \times 583 = 22.2 \text{ Tonne}$ = $\frac{2518 \times 23 \times 583 \times 0.0022}{1000} = 74.3 \text{ Tonne}$ = $(22.2^2 + 74.3^2)^{1/2} = \pm 77.5 \text{ Tonne}$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$ $0.0005 \times 2518 \times 21$ = $\pm 26\text{m}^3$ $0.0001 \times 2518 \times 21$ = $\pm 5\text{m}^3$ = $8 + (26^2 + 5^2)^{1/2} = \pm 34\text{m}^3$ = $34 \times 583 = 19.8 \text{ Tonne}$ = $\frac{2518 \times 21 \times 583 \times 0.0022}{1000} = 67.8 \text{ Tonne}$ = $(19.8^2 + 67.8^2)^{1/2} = \pm 70.6 \text{ Tonne}$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$ $0.0005 \times 2518 \times 18$ = $\pm 23\text{m}^3$ $0.0001 \times 2518 \times 18$ = $\pm 5\text{m}^3$ = $8 + (23^2 + 5^2)^{1/2} = \pm 32\text{m}^3$ = $32 \times 583 = 18.7 \text{ Tonne}$ = $\frac{2518 \times 18 \times 583 \times 0.0022/1000}{1000} = 58.1 \text{ Tonne}$ = $(18.7^2 + 58.1^2)^{1/2} = \pm 61.0 \text{ Tonne}$	$3.1 \times 2.518^{(1)}$ = $\pm 8\text{m}^3$ $0.0005 \times 2518 \times 3$ = $\pm 4\text{m}^3$ $0.0001 \times 2518 \times 3$ = $\pm 0.8\text{m}^3$ = $8 + (4^2 + 0.8^2)^{1/2} = \pm 12\text{m}^3$ = $12 \times 583 = 7.0 \text{ Tonne}$ (4) = $\frac{2518 \times 3 \times 583 \times 0.0022}{1000} = 9.7 \text{ Tonne}$ = $(7.0^2 + 9.7^2)^{1/2} = \pm 12.0 \text{ Tonne}$
VAP Volume - calibration table - dome = 0.5% - tank = 0.05% Shell Shrinkage = $0.000042 \times 5 \times 100 = 0.02\%$ volume = $(114^2 + 5^2)^{1/2} = \pm 114\text{m}^3$ mass = $114 \times 2.183^{(6)}$ = $\pm 0.25 \text{ Tonne}$ Density - Temp = $\pm 5^\circ\text{C} = 2.2\%$ Press = $\pm 5\text{mb} = 0.5\%$ Mole Comp = 1.6% } = 2.8% (6) Overall Vapour Uncertainty mass = $(0.25^2 + 1.5^2)^{1/2} = 1.5 \text{ Tonne}$	0.005×22711 } $\pm 114 \text{ m}^3$ 0.0005×2342 } = $0.0002 \times 25053 = \pm 5\text{m}^3$ = $(114^2 + 5^2)^{1/2} = \pm 114\text{m}^3$ = $114 \times 2.183^{(6)} = \pm 0.25 \text{ Tonne}$ = $\frac{25053 \times 2.183 \times 0.028}{1000} = \pm 1.5 \text{ Tonne}$ = $(0.25^2 + 1.5^2)^{1/2} = 1.5 \text{ Tonne}$	0.005×22711 } $\pm 117\text{m}^3$ 0.0005×7377 } = $0.0002 \times 30088 = \pm 6\text{m}^3$ = $(117^2 + 6^2)^{1/2} = \pm 117\text{m}^3$ = $117 \times 2.183 = \pm 0.26 \text{ Tonne}$ = $\frac{30088 \times 2.183 \times 0.028}{1000} = \pm 1.8 \text{ Tonne}$ = $(0.26^2 + 1.8^2)^{1/2} = \pm 1.8 \text{ Tonne}$	0.005×22711 } $\pm 121 \text{ m}^3$ 0.0005×14929 } = $0.0002 \times 37640 = \pm 8\text{m}^3$ = $(121^2 + 8^2)^{1/2} = \pm 124 \text{ m}^3$ = $121 \times 2.183 = 0.26 \text{ Tonne}$ = $\frac{37640 \times 2.183 \times 0.028/1000}{1000} = \pm 2.3 \text{ Tonne}$ = $(0.26^2 + 2.3^2)^{1/2} = \pm 2.3 \text{ Tonne}$	0.005×22711 } $\pm 140\text{m}^3$ 0.0005×52699 } = $0.0002 \times 75410 = \pm 15\text{m}^3$ = $(140^2 + 15^2)^{1/2} = \pm 141\text{m}^3$ = $141 \times 2.183 = \pm 0.31 \text{ Tonne}$ = $\frac{75410 \times 2.183 \times 0.028/1000}{1000} = 4.6 \text{ Tonne}$ = $(0.31^2 + 4.6^2)^{1/2} = 4.6 \text{ Tonne}$
TOTAL Liquid = Vapour =	$57911 \times 583 = 33762 \text{ Tonne}$ $25053 \times 2.183 = 55 \text{ Tonne}$ 33817 $(77.5^2 + 1.5^2)^{1/2} = 77.5 \text{ Tonne} = 0.23\%$	$52875 \times 583 = 30826 \text{ Tonne}$ $30088 \times 2.183 = 66 \text{ Tonne}$ 30892 $(70.6^2 + 1.8)^{1/2} = 70.6 \text{ Tonne} = 0.23\%$	$45324 \times 583 = 26424 \text{ Tonne}$ $37640 \times 2.183 = 82 \text{ Tonne}$ 26506 $(61.0^2 + 2.3^2)^{1/2} = \pm 61.0 \text{ Tonnes} = 0.23\%$	$7554 \times 583 = 4404 \text{ Tonne}$ $75410 \times 2.183 = 165 \text{ Tonne}$ 4569 $(12^2 + 4.6^2)^{1/2} = 12.9 \text{ Tonne} = 0.23\%$

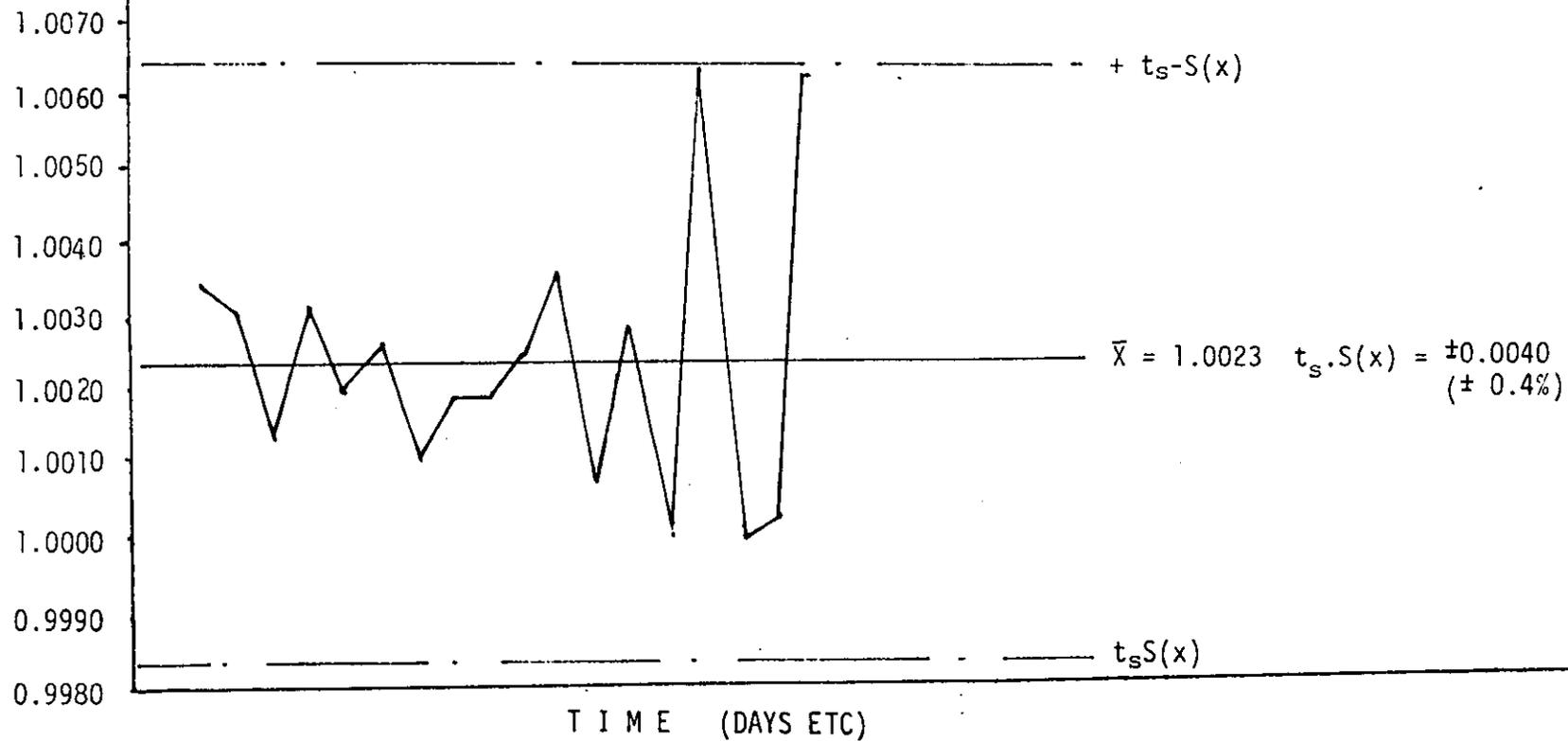
(1) Tank dia = 56.62mm ($\text{m}^3/\text{m} = 2518$)
 $\text{m}^3/\text{mm} = 2.518$
(2) Height of Stillwell = 33000 mm
(3) Av. temp error = $\pm 5^\circ\text{C}$
(4) Density of C3 = $583\text{kg}/\text{m}^3$
(5) Tonne - kg/1000
(6) Uncertainty of densitometer = $\pm 0.2\%$
(7) Error in correcting densitometer to tank temperature
(8) Density of vapour C3 = $2.183\text{kg}/\text{m}^3$
(9) Density of Vapour = $\frac{(288 \times P_v \times 44.097)}{T_v \times 1.013 \times 23.6451}$
(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 $\pm 0.4\%$)



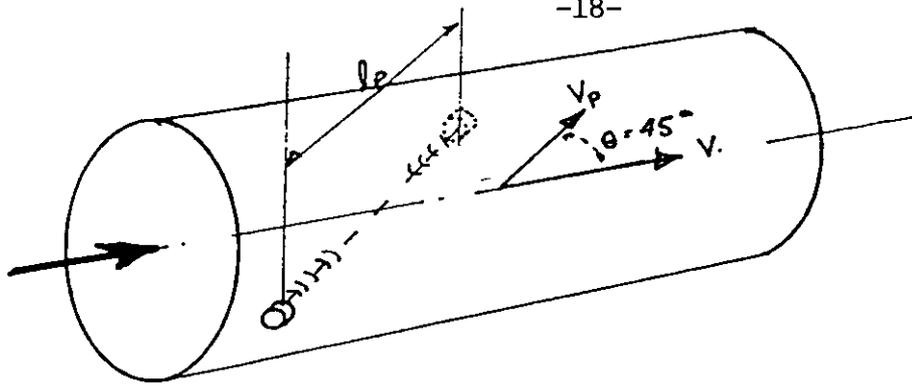
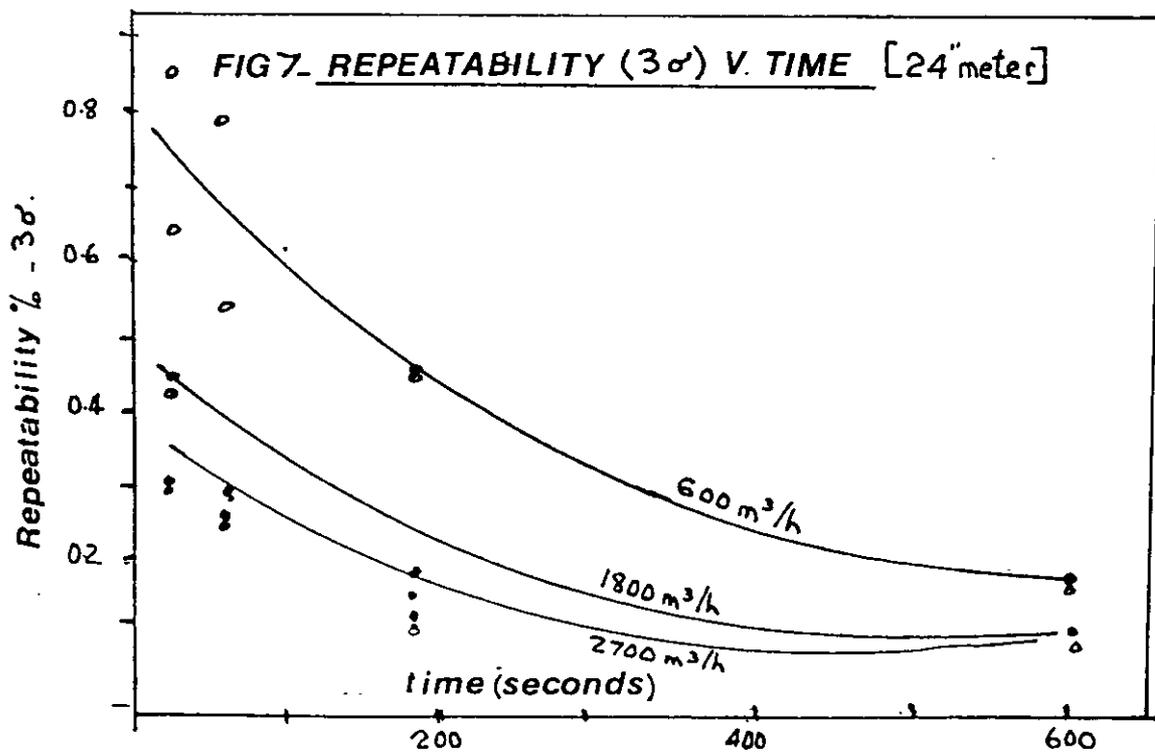
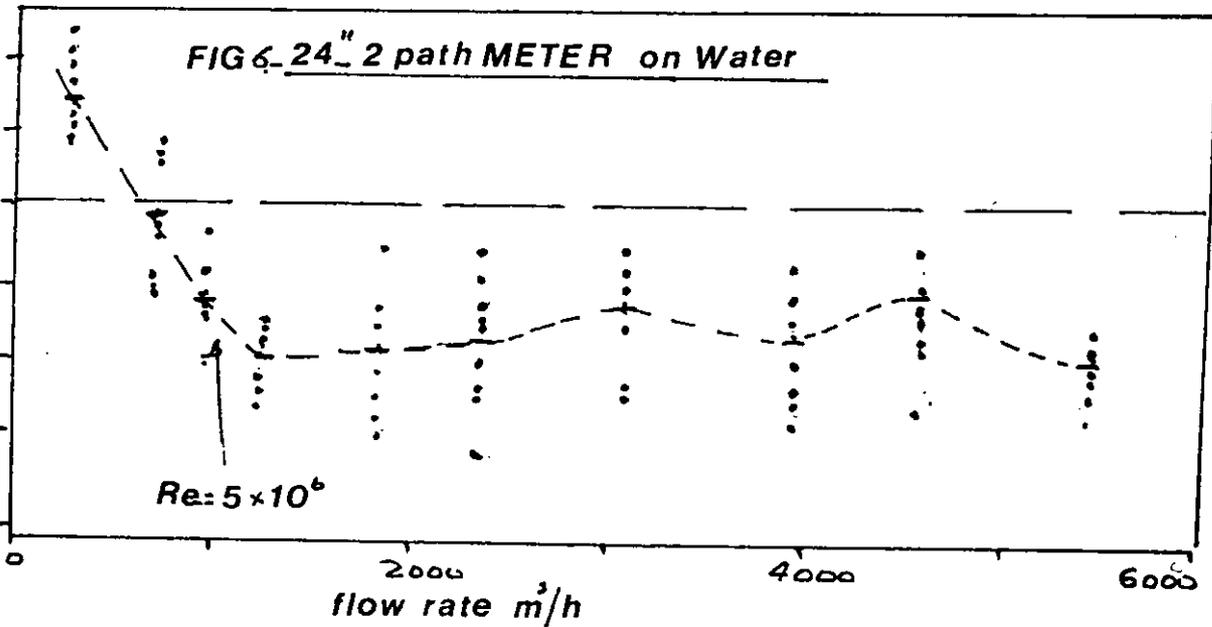


FIG 5 Principles



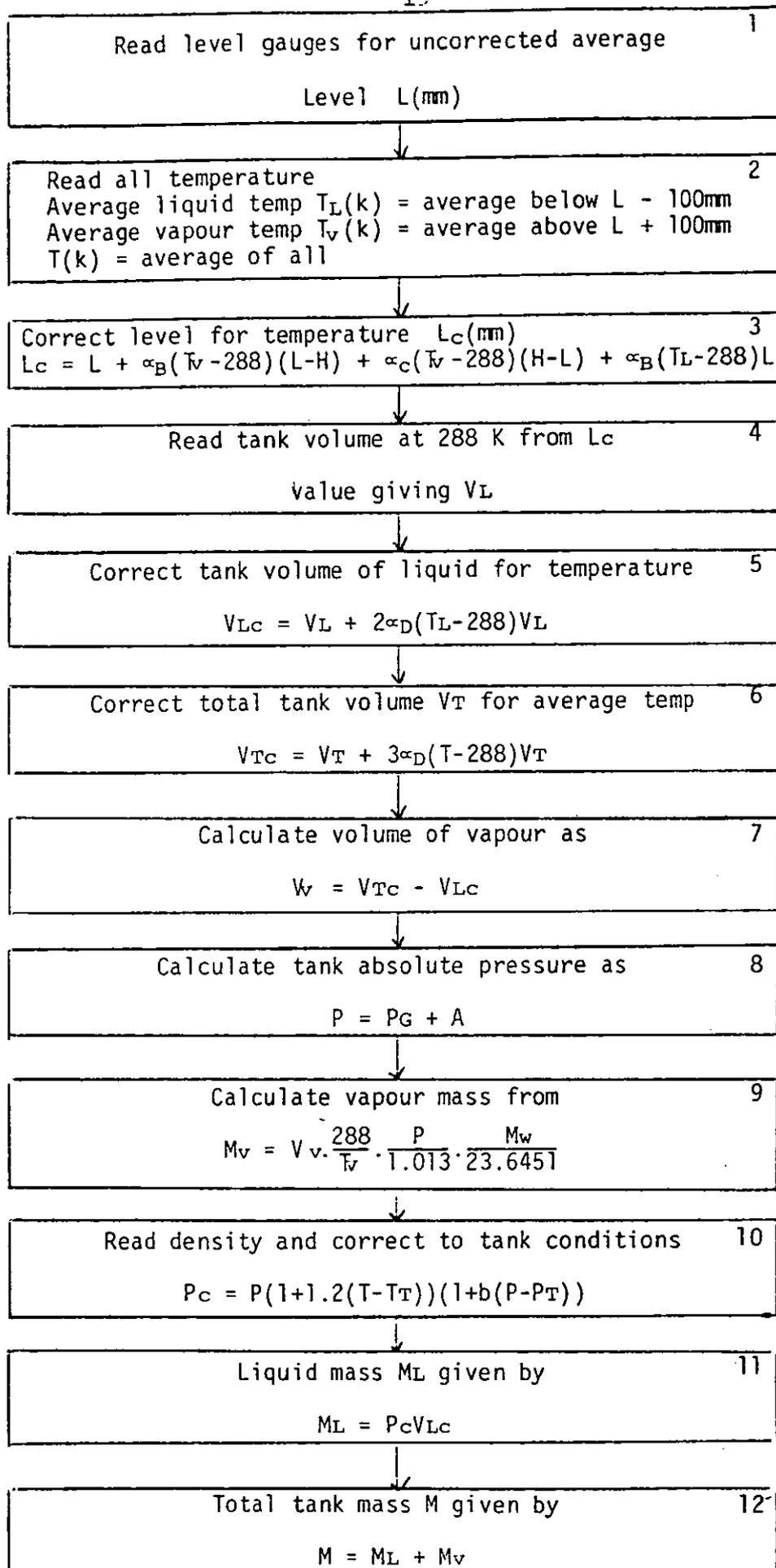


FIGURE 8

TANK INVENTORY CALCULATION

FIG 9. VORTEX METER - Bluff Body Design

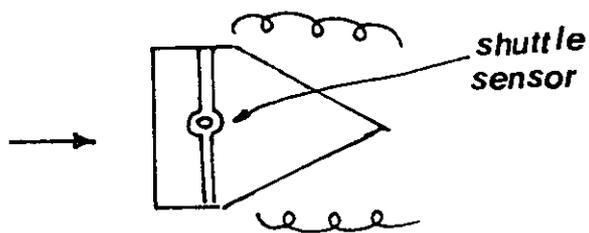
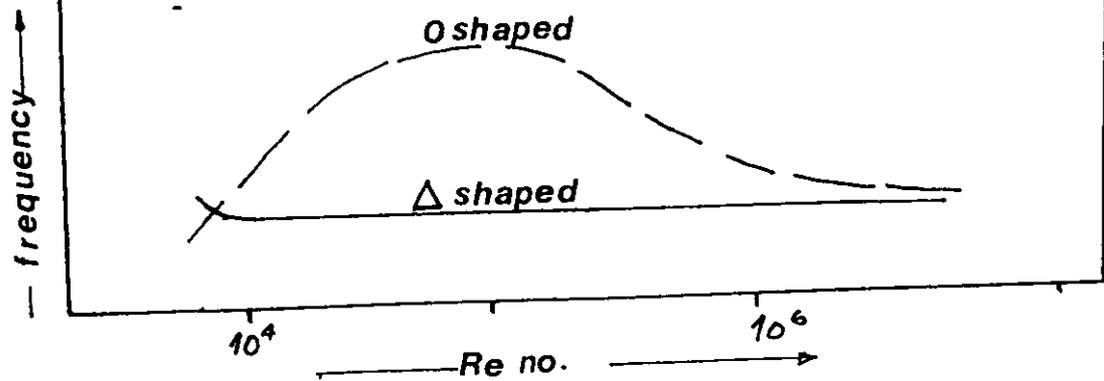
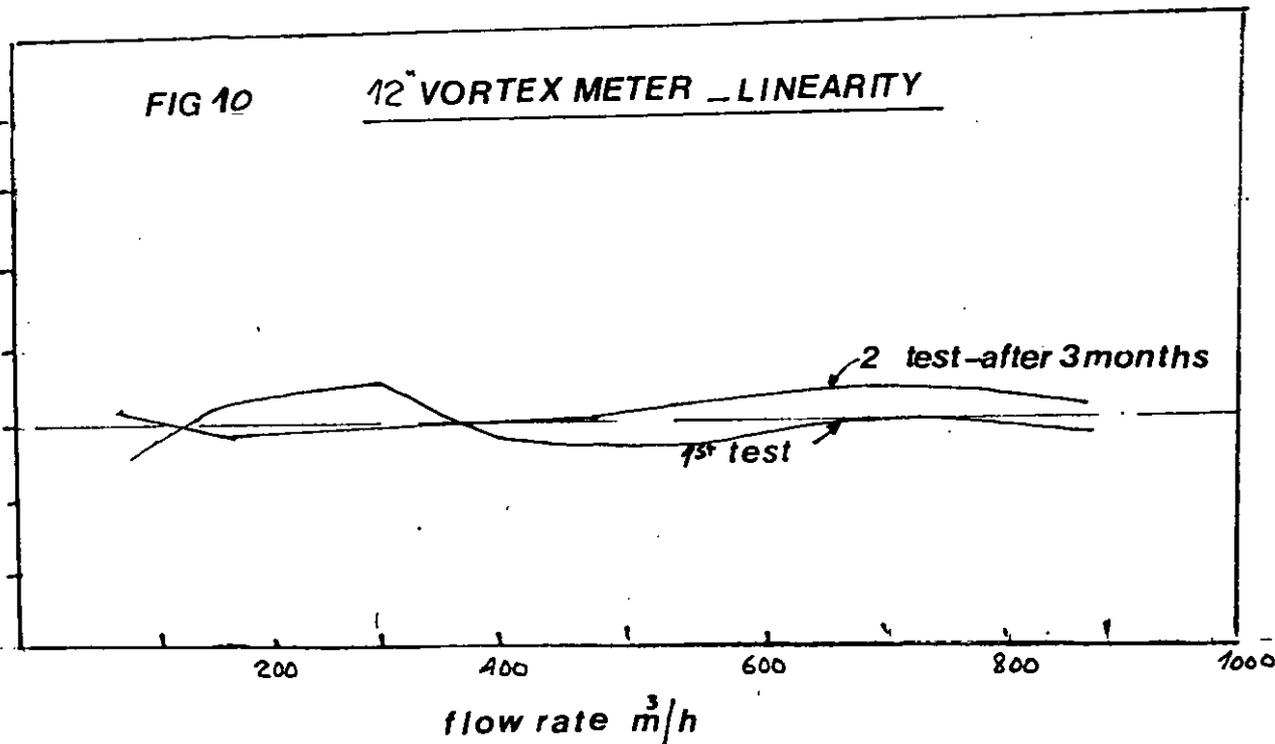
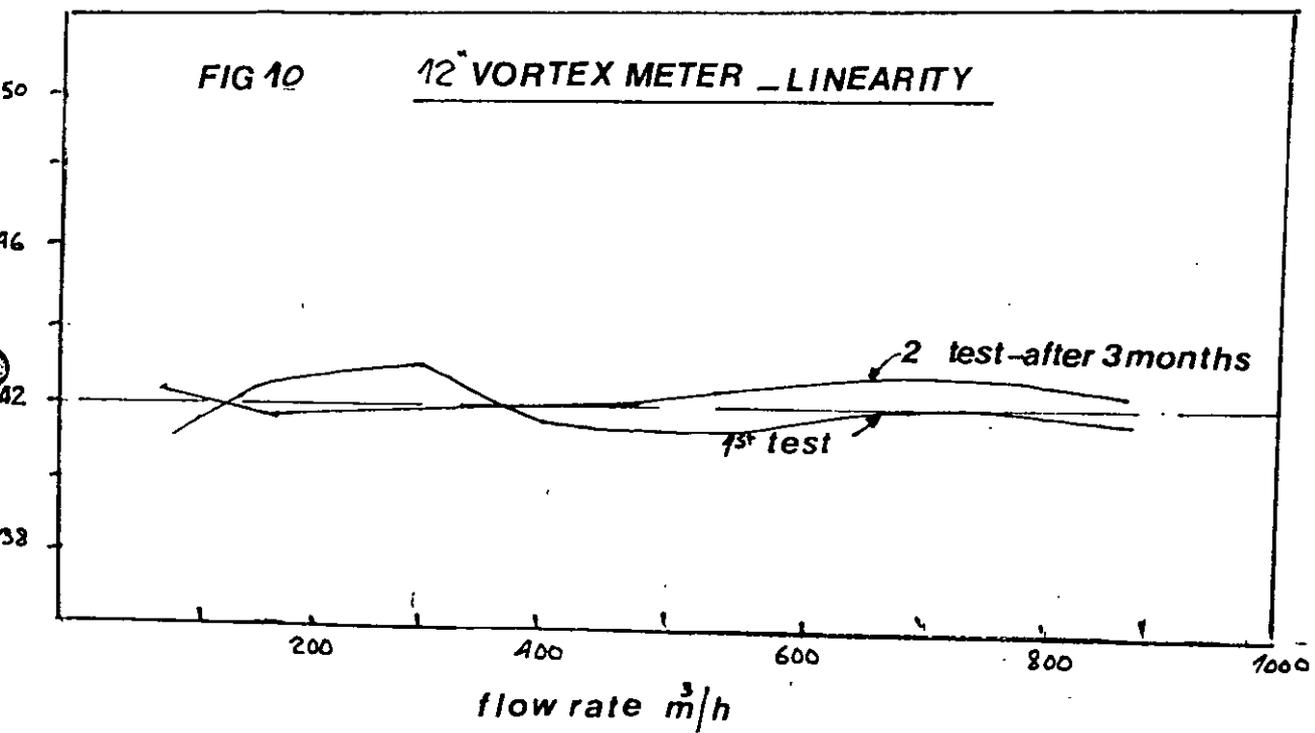
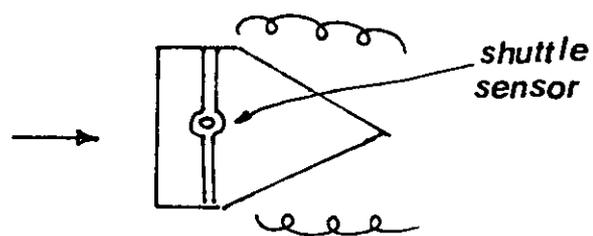
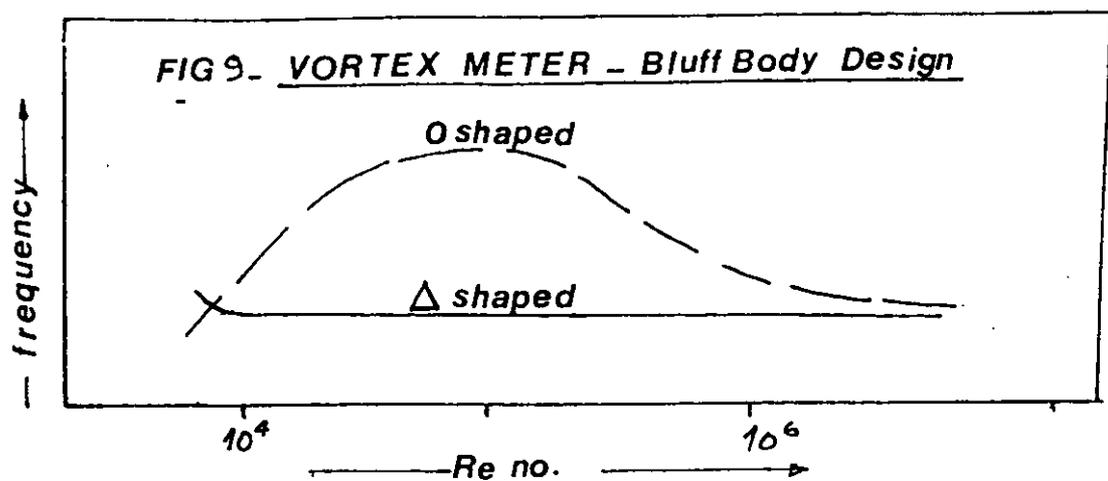


FIG 10 12° VORTEX METER - LINEARITY





MEASUREMENT OF GAS AND LIQUIDS
Rogaland Regional College, Stavanger
7.-10. June 1982

Vortex Meter and
Coriolis Mass Flow Meter

Geir Magne Nesbakken
Norsk Hydro a.s
Porsgrunn Fabrikker

Vortex meter

The vortex meter represents one of the most promising developments in flow metering over the last ten years.

It is a very robust meter and for smaller dimension also among the cheapest. Possibly the greatest advantages is the linear digital output and the possibility to calibrate the meter accurately under other than normal operating conditions.

Fig. 1 shows a typical vortex meter. It consists of a straight pipe with a bluff body obstructing the flow. The bluff body causes vortices to be shed alternately on each side as shown in fig. 2. This is the same effect as is seen by a flag waving in the wind. The flag pole is producing vortices in the flowing air and the flag is thrown alternately to each side. The frequency is proportional to the wind velocity.

Any obstruction in a pipe will cause vortices to be produced. The frequency and the minimum and maximum flow rate for stable vortex generation is given by the shape of the bluff body. A few different designs is shown in fig. 3.

For a particular design the calibration factor is given by the ratio of the width of the bluff body to the length of it. This is a rather teoretical approach since the bluff body is installed in a circular pipe and it is therefore difficult to determine the correct length to be used in the calculation.

On the other hand, when one vortex meter is calibrated the resulting calibration factor can be applied to other equal meters. Since the mechanical construction is very simple, it is not difficult to maintain the same dimensions within the appropriate tolerances.

When the mechanical dimensions are given the calibration factor is only depending on the Reynolds number. For most designs the calibration is linear within 1 or 2 percent for Reynolds number larger than 30 000 up to a limit given by the onset of cavitation. Its this property that is used when calibrating under conditions very different from normal operating conditions. Once the calibration curve is determined versus Reynolds number for example under water calibration, the meter can be used for any gas or liquid as long as the Reynolds number is known.

In the early days of vortex meter it was often claimed that the calibration was very little dependent on the flow profile, and straight upstream lengths of as little as three pipe diameters were recommended by some manufacturers. Over the years the requirements have steadily increased and today most manufacturers specify 20 diameters upstream straight length.

In a general description it is difficult to give accuracy figures, but under the conditions outlined above the accuracy should be in the region of $\pm 1\%$. For more accurate results calibration of each individual meter under normal operation conditions is necessary.

Calibration of a vortex meter is complicated by the fact that the vortex generation is not regular. This means that a large number of pulses have to be counted to get the correct average. Also it makes the use of pulse splitting techniques very difficult.

It is very difficult to get manufacturers comments on installation effects. Only one is known to give details of calibration shift caused by the installation of meters in pipes with internal diameter that is not exactly equal to the meter internal diameter. In practise this is an almost unavoidable problem since the inner diameter of pipes with nominally the same diameter is depending on pressure rating.

At the calibration laboratory at Norsk Hydro, Porsgrunn, three different meters have been tested on installation effects. Since the time schedule did not permit the results to be discussed with the meter manufacturers, the results will not be reported in writing at this stage.

The results will be given in the lecture without referring to the manufacturers names.

Coriolis mass flow meter

As we have seen earlier in this seminar, we are normally interested in the mass rather than in volume and density. For a long time different methods of direct mass determination have been tried, with more or less success.

An instrument working by a completely new principle has now been on the market for 4 or 5 years. In basic principle it only measures mass flow independent of how this mass is composed. Of course there are practical limits to this statement, but it is felt important to stress that the principle is a real mass flow measurement.

The instrument is shown in fig. 4. The only wetted part is a U-tube vibrating at its resonant frequency excited by a magnetic coil. If there is no flow through the tube the two legs of the U-tube will have parallel oscillations, but as soon as there is flow the two legs will twist relative to each other. This twist angle is proportional to mass flow rate.

To explain the mechanics behind the meter, let's first consider the Coriolis force generally. This is the force caused by earth rotation that determine the wind directions around high and low pressure areas on the weather map. For the same reason a train on the northern hemisphere will always lean against the right rail on a straight track.

If we look at the boy on the turntable in fig. 4 we see that he will be leaning to the left if he is going to walk along a straight line painted on the turntable. This is because his velocity in the direction perpendicular to the direction he is walking is increasing proportionally to the distance from the center of the turntable. This velocity increase requires an acceleration force (Coriolis force) and to take up this force the boy has to lean.

Exactly the same would happen if we mounted a straight pipe on the turntable. Any volume of mass that was passing through this pipe would be accelerated perpendicular to the pipe axis. The acceleration force would have to be taken from the pipe, and if the pipe was fixed at only one end it would therefore bend. If we change the flow direction the bending direction will also change.

Each leg of the U-tube is fixed at only one end and since the flow direction is opposite in the two legs the whole U-tube will twist. Because the U-tube itself is rotating in alternating directions (oscillating), the twist angle is oscillating at the same frequency. The angle itself can be detected as a phase difference between the two "corners" of the U-tube.

Basically the only factor influencing the tube twisting is the mass flow, independent of density, viscosity, temperature and so on.

However since the U-tube is not rotating at a constant angular velocity, but oscillating, the meter is density dependent. This is because with changing density the oscillation frequency will change.

The vibration amplitude is determined by the magnetic force and the elasticity of the U-tube. The elasticity is temperature dependent causing the meter itself to be temperature dependent.

Both density - and temperature effects are predictable and can therefore be compensated.

The meter is virtually independent of viscosity, the limiting factor being the pressure loss through the U-tube.

Its also independent of deposits on the tube wall if the density of the deposit is the same as the fluid density.

Corrosion or erosion may cause the elasticity of the U-tube to change, thereby changing the calibration.

The most important limitation on the use of the meter is the maximum tube diameter of 50 mm (at least at present time), and the requirement of a minimum mass flow rate compared to the U-tube mass, making the meter unsuitable for low pressure gas.

The original idea behind this lecture was to present results from a test of one of these meters. The reason this cannot be done is that the U-tube broke after four days operation in the flow laboratory.

Based on technical discussions with the inventor, visit at a user and theoretical study of the meter, I still regard it to be a break-through in direct mass measurement.

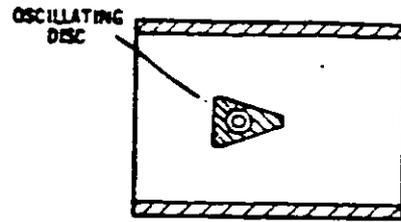
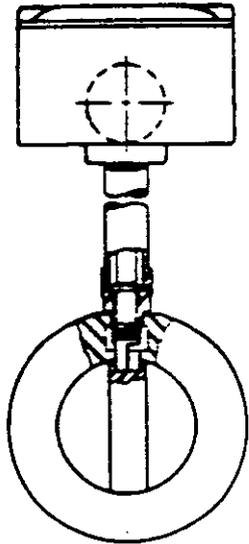


Fig. 1 Typical vortex meter

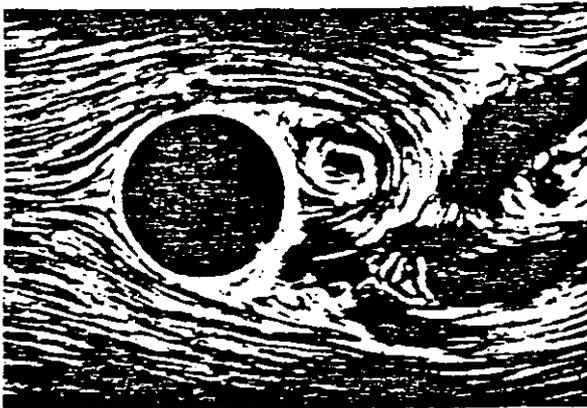


Fig. 2 Vortices behind a circular body

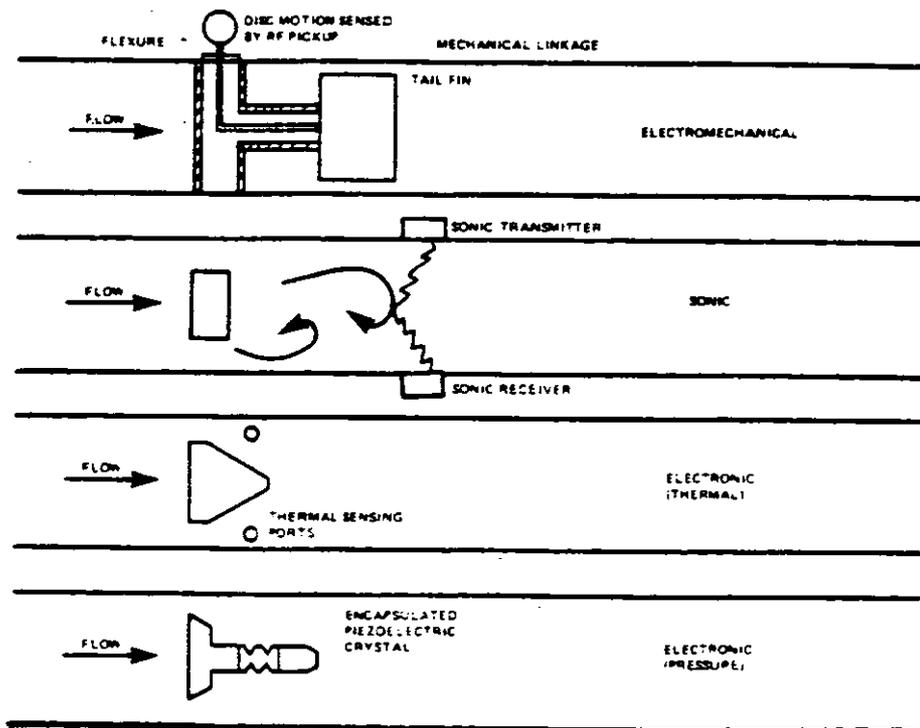
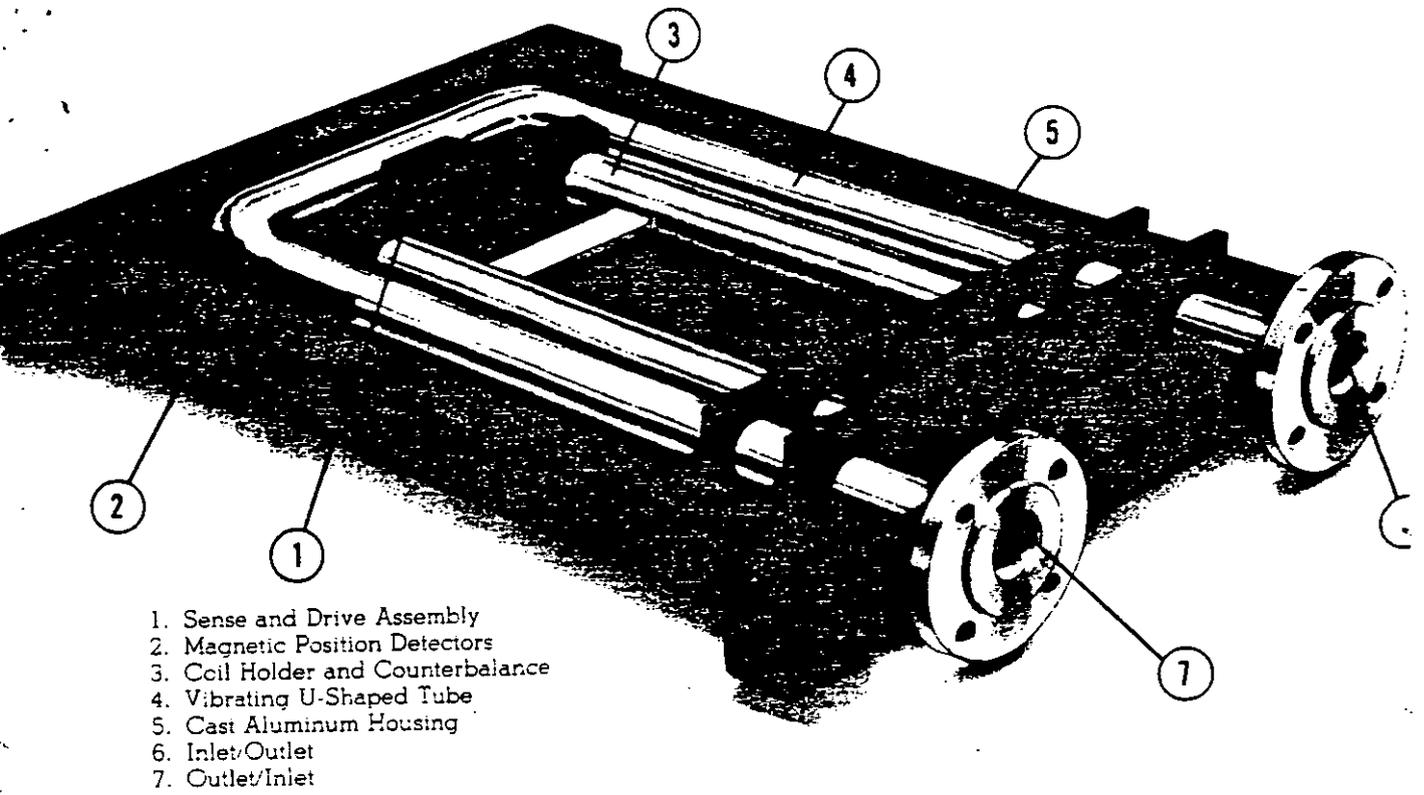


Fig. 3 Bluff bodies of different shapes



1. Sense and Drive Assembly
2. Magnetic Position Detectors
3. Coil Holder and Counterbalance
4. Vibrating U-Shaped Tube
5. Cast Aluminum Housing
6. Inlet/Outlet
7. Outlet/Inlet

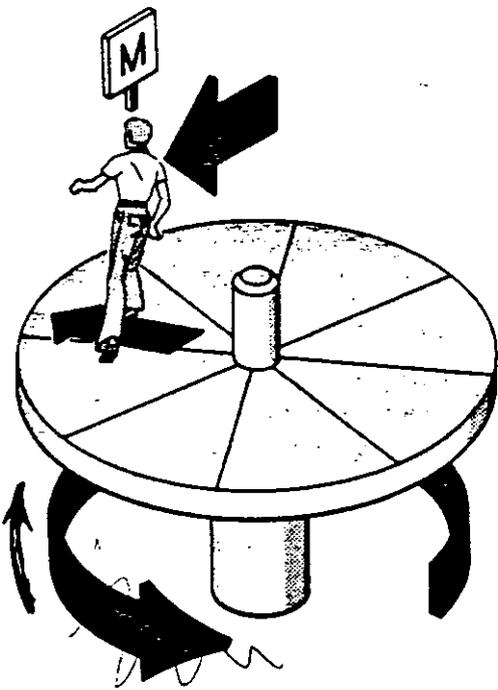


Fig. 3 If a person were standing at the center of a merry-go-round and tried to walk in a straight line toward the edge, he or she would have to lean sideways against the Coriolis force to stay on line. The Coriolis force, F_c , can be calculated using the mass of the person's body, M , the velocity of travel, V , toward the edge, and the angular velocity of the merry-go-round, ω . $F_c = 2M\omega \cdot V$.

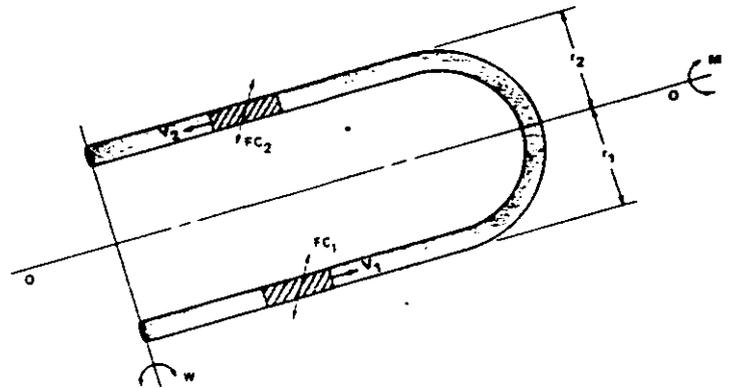


Fig. 5 In operation, the mass flow meter forces, f_{c1} and f_{c2} , create an oscillating moment, M , about axis, O .

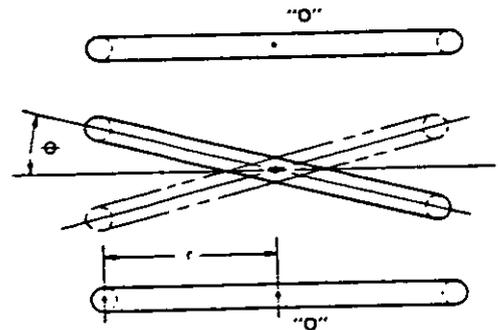


Fig. 6 End view of the U-shaped tube showing the parameters for calculating the torque and its relationship to the mass flow rate. The torque depends on the deflection angle, θ , of the pipe and its spring constant, K .

Fig. 4. Mass flow meter

MEASUREMENT OF GAS AND LIQUIDS

Teknologidagene i Rogaland
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ULTRASONIC FLOWMETERS

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ULTRASONIC FLOWMETERS

The term "ultrasonic flowmeter" covers a variety of different methods and meters because ultrasound may be used in a number of ways to extract information from a flow-process [1]. Here I will only deal with single phase fluids.

A division into three main groups, based upon principles of operation might be as follows:

- Reflection type or Doppler type meters.
- Transmittance type meters
(including time-difference, phase-difference and frequency-difference/sing around-meters)
- cross-correlation meters.

The latter emphasise cross-correlation rather than ultrasonics.

Common to all these is the use of ultrasonic sensor devices and also the advantage of nonintrusiveness. Apart from that they utilize totally different principles to measure the velocity of flowing fluid and should thus be regarded as different types of instruments.

I will now look at each one in particular to point out

- what they are measuring
- indicate how they do it
- how well they perform on the base of the principle used.

For the sake of completeness we should first repeat what is going on inside the pipe and what to measure.

In practice, most flows are turbulent and the instant picture could be somewhat like Figure 1a. Eddies are generated and move along the pipe forming a rather complicated flowpattern. If particles, bubbles etc. are contained they will be moved correspondingly. The time-average velocity in each particular position of the cross-section, however, describes a continuous distribution

(velocity profile) across the pipe, Figure 1b.

This profile is subject to change by Reynolds number, changes in wall-roughness, upstream bends and fittings causing swirl and non-symmetric profiles. The quantity to be measured is the exact integral of the velocity profile or its exact mean value.

Reflection type or Doppler flowmeter [2], [3], [4]

These meters utilize the effect of a shift in frequency when a sound wave is reflected from a moving object, Figure 2.

The shift is proportional to the velocity component of the reflecting source in the direction of the ultrasonic wave. Thus a Doppler meter is in fact measuring the velocity of bubbles, particles etc. moving along with the fluid. Due to turbulence as mentioned this is bound to be an averaged value, but also because reflections will occur from various locations in the flow dependent on distribution of reflecting sources.

This measuring principle implies the following features and problems:

- The sensor could be clamped or bonded to the exterior of the pipe without any cutting or drilling of holes. This may be the most outstanding and attractive feature of this type of meter.
- Unlike most other meters, the Doppler meter operates only if there are particles, suspended solids, fluid interfaces or gasbubbles in the process fluid. It will not work with very clean fluids.
- The meter is profile dependent, like most other meters. The measured volume is normally 10-20 mm inside the pipe-wall and thus a real mean pipe-velocity measurement is not performed. In small pipes (< 2" diameter) you will probably read high, in larger pipes (> 6"), low. To correct for profile variations is thus difficult [3].
- The meter depends on distribution of reflecting sources. Gas in liquids tends to be concentrated near the wall, nonhomogeneous mixtures may give strange results and the larger the

reflecting sources the less likely it is that they will move with the same velocity as the surrounding fluid [3].

- The external mounting does not relax the need to know the pipe cross-section area which should be determined according to required and achievable accuracy.
- The meter provides a much better repeatability than accuracy which means that if in situ calibrations can be made, the accuracy in certain applications can be greatly improved.

Thus the Doppler flowmeter should be regarded as a meter of

- low accuracy ($\pm 2\%$ F.S.D. or more)
- high repeatability ($< \pm 1\%$ F.S.D.)
- low initial purchase price (10.000 NOK)
- low installation and maintenance cost
- and requiring in situ calibration to achieve best accuracy.

Transmittance type

These meters utilize in slightly different ways the fact that an ultrasonic wave travelling downstream will go faster than one travelling in the opposite, upstream direction, Figure 3.

The difference in propagation velocity is proportional to the fluid velocity component in the direction of the ultrasonic beam. This component varies across the pipe due to the velocity profile of the flowing medium, but also because of turbulence. Thus what is achieved by measuring the difference in propagation velocity is the mean value or the integral of velocity along the acoustic path.

The different methods achieve this in the following way (Fig. 4):

- The transit time difference method measures difference in time of arrival of two pulses transmitted simultaneously but in opposite directions along the same path. This is the most used method and also regarded as most accurate. Time resolution is down to 0.1 ns (10^{-10} s) corresponding to 1 mm/s in a 4" pipe. Sampling rate depends on transit time which is

approximately 0.1 ms per inch of pipe-diameter for a gas-meter [4], [5], [6].

- The differential phase method uses the phase difference between two waves continuously transmitted in opposite directions across the pipe. This is essentially the same measurement as the time difference above but uses other techniques in detection electronics, and produces an average over several wavelengths of the acoustic signal.

High quality state-of-the-art electronics technology is required to achieve high accuracy [5], [6].

- The frequency difference method or sing-around technique transmits several pulses in both directions across the pipe. Each new pulse is transmitted at the arrival of the previous one. The frequency difference between the pulse-train transmitted downstream and that upstream is the averaged value of the mean velocity along the path. To achieve high resolution this implies averaging over say seconds, increasing as pipe-diameter increases.

As shown, the various methods all measure the same basic quantity. The difference is in the electronics and shows up mainly in response time and to some extent also in accuracy [5], [6].

As already indicated the transit time difference seems to be the most preferable method and is most the one used in instruments claiming high accuracy.

Of the various configurations of transducers in the meter section, the single path is most common (Figure 5). The path is located along the pipe diameter.

Transducers could be mounted withdrawn from the pipe wall or slightly intruding as shown (Figure 6).

The measuring principle and practical configurations of single path meters imply:

- none or negligible obstructions of flow-profile.
- relatively clean fluids required.

Particles, bubbles etc. will introduce scatter and increase acoustic damping and thus making it more difficult to get signals through to the receiver. Large objects may even block the signals completely. Signal processing rejecting "bad measurements" will to some extent overcome these problems, depending on size and density of the disturbing elements.

- profile dependency exists.

The measured mean velocity is not the mean pipe-velocity, but the mean of a narrow path along the diameter. A meter factor in the range 1.08 to 1.04 is thus required for Reynolds number from 10^4 to 10^7 [7]. This corresponds to a change of 1% for each magnitude of 10 in Reynolds number.

Additional changes of profile caused by roughness and upstream disturbances resulting in swirl and non-symmetrical profiles could introduce even more serious errors. Errors of 3-4% 6 diameters downstream of a 90° elbow have for example been reported [8].

- Transducers could be mounted in existing pipe-works, but to achieve high accuracy meters should be installed as a separate section.
- Cross sectional area should be known. This will be the case when a separate pipe section is delivered.
- When electronics are properly designed meters will show very high repeatability.
- Calibration is unaffected by temperature, pressure, viscosity, and composition of fluid. As a physical property of non-flowing fluid which will depend on these factors, sound velocity is continuously measured and accurately compensated for. Thus the requirement is only to get the signal through to the receiver.

To summarize: The single-beam ultrasonic flowmeter is a meter of

- medium accuracy ($\pm 1\%$ F.S.D. or more)
- high repeatability ($\pm 0.1\%$ F.S.D. claimed)
- initial purchase price relatively independent of pipe size

Transducer + electronics 35.000 NOK

4" pipe section 1.500 NOK

48" 11.000 NOK

- maintenance costs are relatively low because of the absence of moving parts and simple mechanical design.
- relatively low calibration cost.

Multipath meters [2]

The main limitation of a single path meter is its flow profile dependence. The logical solution to this is to apply multiple paths at different locations in pipe cross-section. Each path results in a mean value through a defined part of the velocity profile (Figure 7). Combining these multipath measurements in proper ways will give the flowrate of the pipe.

Various computational schemes have been proposed, based on numerical integration formulas and knowledge of the extreme limits of profile-variations, and integration errors in the region of 0.1% are achievable for Reynoldsnumber ranging from 10^4 and upwards [10], [11], [12]. Errors less than 1% are reported even when measuring only 4 diameters downstream of a 90° elbow [13].

The multipath ultrasonic flowmeter thus possesses the same properties as the single beam meter considering non-intrusiveness, high repeatability, necessity for relatively clean fluids and independence of pressure, temperature, viscosity etc.

The meter section and transducer mountings should be precisely machined to achieve high accuracy. Mounting of transducers in existing pipe is not therefore recommended.

However, the multipath ultrasonic flow-meter is supposed to be the most accurate available for medium to large pipes [14].

It should be regarded as a meter of

- very high accuracy ($\pm 0.5\%$)
- very high repeatability ($\pm 0.1\%$ F.S.D. claimed)
- high purchase cost although relatively independent of size (> 100.000 NOK! This I think is closely related to low production volume and partly to expensive electronics.
- relatively low maintenance costs.

The cross-correlation flowmeter

This meter is based upon a comparison of "state" between two cross-sections of pipe located close to each other. The "state" defined as particles, bubbles, eddies etc. present at a given instant. The "state" of each cross section can be detected by an ultrasonic path sensing acoustic damping or velocity along the diameter. Like fingerprints in received signal the various state-pattern will be sensed in the two cross-sections with a time difference corresponding to the velocity of eddies, particles, bubbles etc. (Figure 8) [15], [16].

The flow-measurement is done through cross correlation of the two signals/states and thus the classification "ultrasonic" is secondary, as other types of sensor can be used instead.

The measuring principle implies

- external mounting of transducers
- clean or dirty liquids. Able to handle very difficult process fluids, such as corrosive fluids, slurries, vapour etc.
- profile dependence although more difficult to predict than for single-path meters.
- slow response due to cross-correlation calculations.

The meter is a new device on the market and the following characteristics could be suggested:

- low/moderate accuracy (supposed better than Doppler-meters)
- high purchase cost, but independent of pipe-size
- low installation and maintenance cost.

Applicability

So far, I have not distinguished between gas and liquid meters. The Doppler meter is for several reasons not commercially available for gas-metering but for multiphase measurements it could very well be used.

Transmittance types and cross-correlation meters will in principle work both in gas and liquid. In practice, however, the problem of making efficient transducers for gaseous media has to be overcome. This seems to be in the process of being solved now. A few single beam meters are on the market. At CMI we have been building gas-transducers for some years and recently the techniques have been applied in a 4" gas-meter with 3 paths. The meter is still in an experimental stage but accuracies better than 1% are indicated [17]. The meter will even work down to atmospheric pressure. This in fact is often a problem for this sort of instrument.

For crude oil multipath meter is used in the Alaska pipe-line for leak-detection where differences between meters are reported better than 0.1% [11].

The transit time difference meters combine high resolution and high bandwidth thus enabling measurement of transients and oscillations as for example reported by Dordain, ONERA in [18], and as has been mentioned, velocities down to a few mm/s may be measured.

The "clamp on" property of Doppler meters and cross-correlation meters make them very suitable for metering difficult fluids at extreme temperatures, corrosive fluids etc.

Use of temperature, chemical etc. protected transducers in meters of the transmittance type will also provide accurate instruments for these applications.

When referring to their properties I will suggest ultrasonic flowmeters to find an increased use

- where flare gases and exhaust fumes should be monitored, conditions requiring high dynamic range, non-intrusive meters and tough operational conditions.
- where large quantities of valuable fluids have to be metered to the highest level of accuracy over a long period.

Future trends

The ultrasonic meter is already a highly sophisticated electronic instrument and as such will surely benefit from the continuous improvement in technology. This implies

- higher accuracy
- accuracy will be relatively less expensive
- more extensive use of μ -processors to perform
 - corrections for flow-profile variations
 - self-testing and diagnostics
 - self-calibration and self-adjustment
- multipath meters could be made less expensive providing a high accuracy meter at comparable prices. High prices are closely related to electronics and low production volume
- more integrated meters where additional sensors could be added to compute mass flow and composition.

As the multipath meter will be able to provide increased accuracy and perhaps even more: to give accurate readings when distorted flow profiles exist, I think these meters will be paid increasing attention by both producers and users.

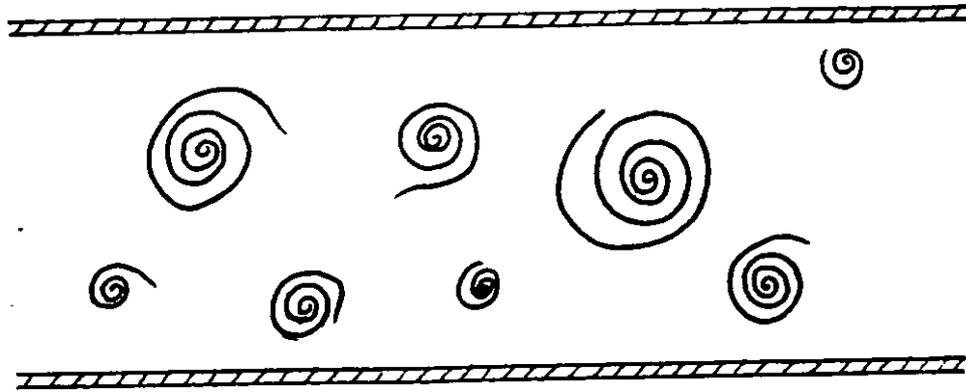
To conclude:

The ultrasonic sensor technique offers a variety of flowmetering methods from the Doppler flowmeter at the low accuracy end to the very accurate multipath meter. As a result of their many advantages these should become even more used in flowmetering both for gases and liquids. As multipath meters will reach reasonable price levels, non-intrusive meters of very high accuracy and repeatability will become available.

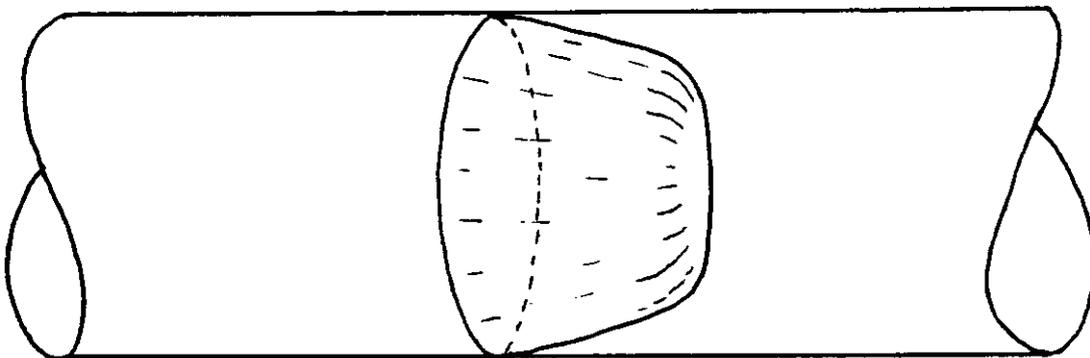
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a) Turbulent flow. Instant picture.



b) Turbulent flow profile. Time average.

Figure 1. Pipe flow.

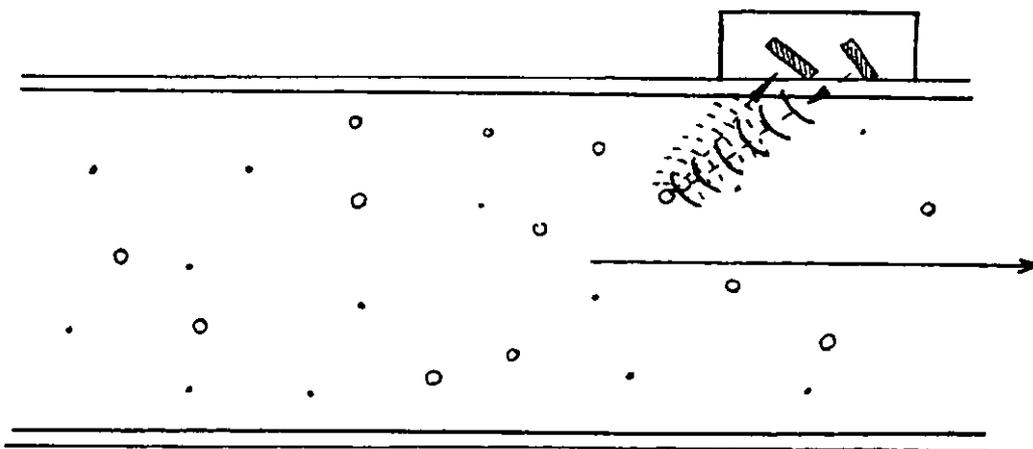


Figure 2. Doppler flow-measurement.

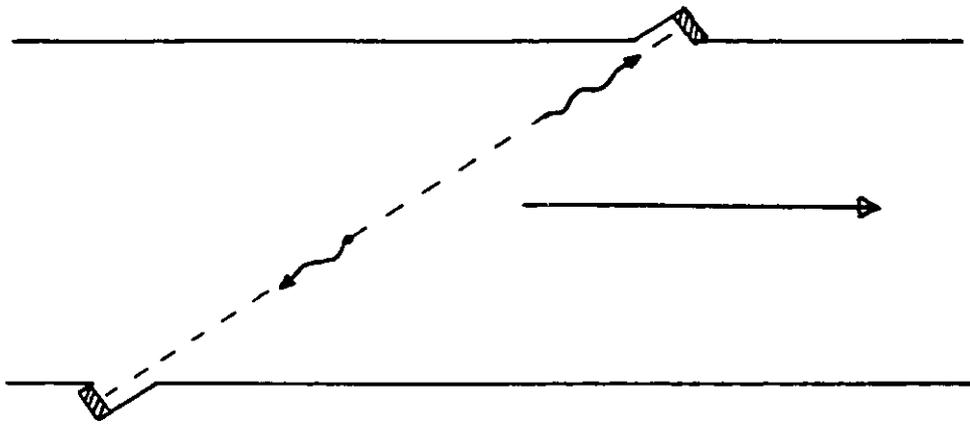


Figure 3. Transmittance type flow-measurement.

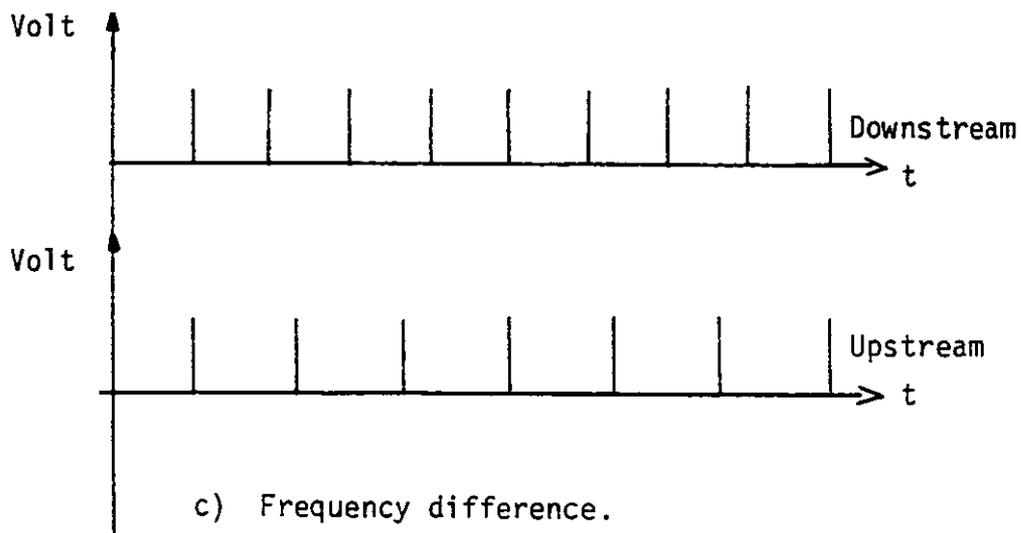
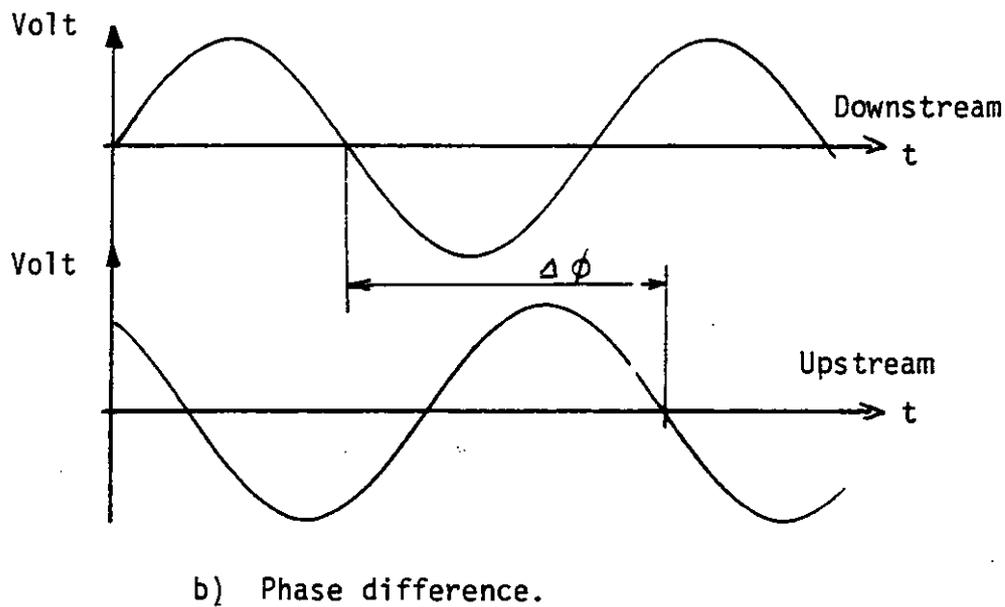
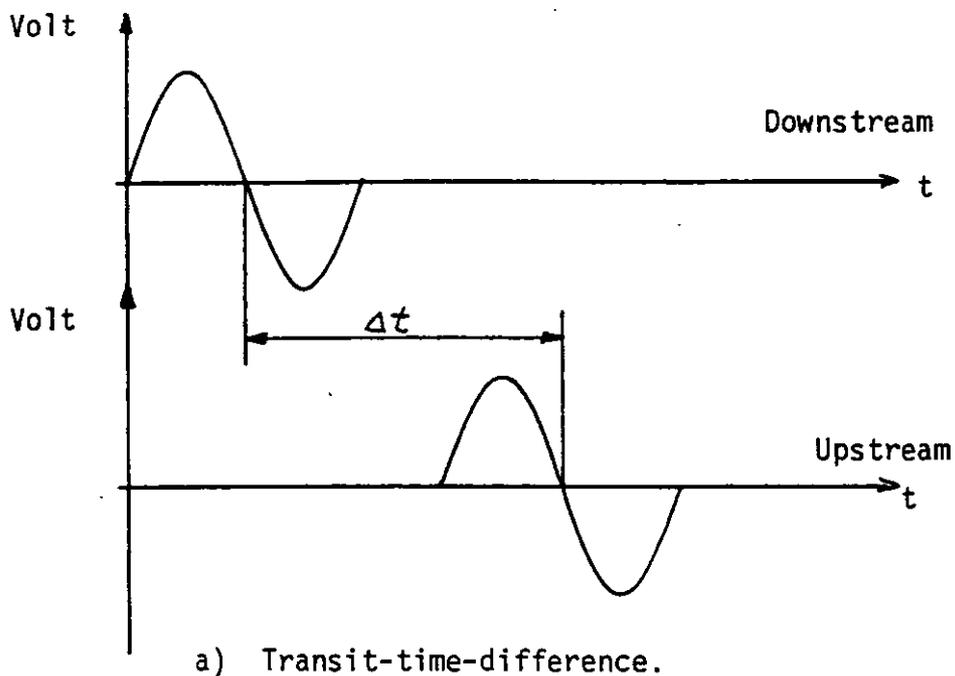


Figure 4. Transmittance type principles.

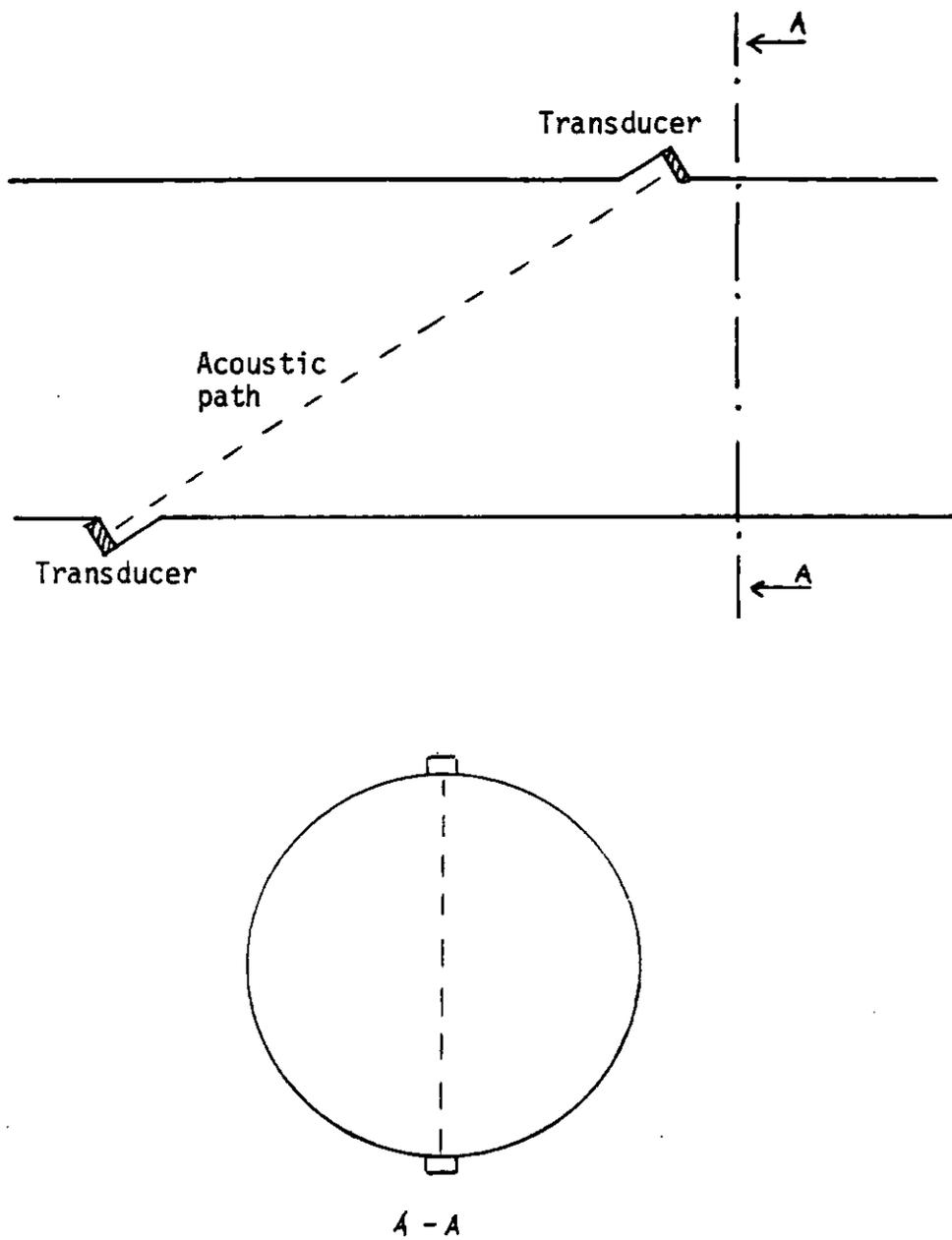


Figure 5. Singlebeam ultrasonic flowmeter.

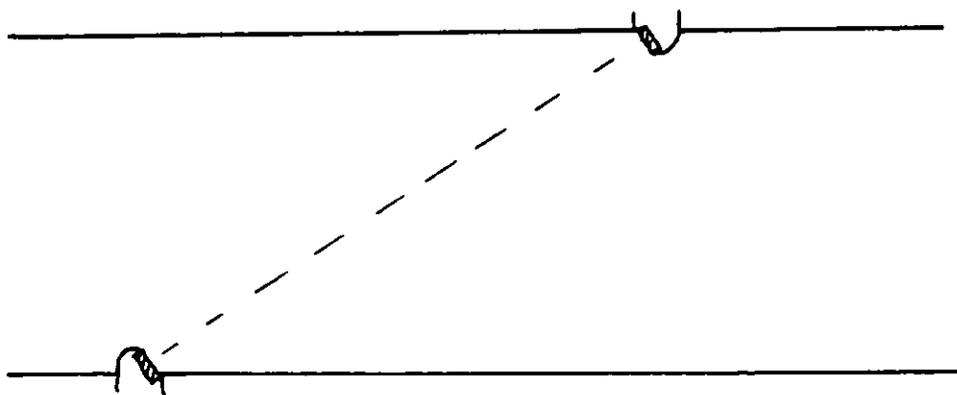


Figure 6. Singlebeam ultrasonic flowmeter.
Transducers slightly intruded.

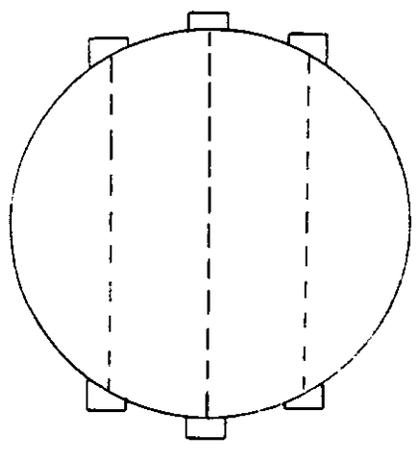
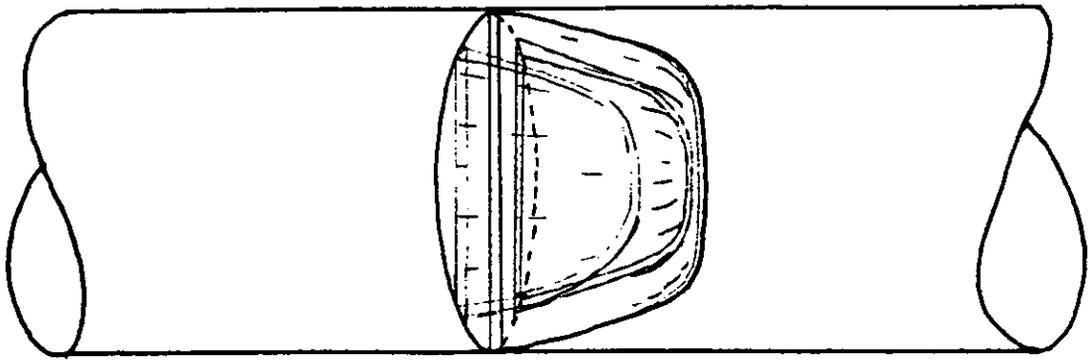


Figure 7. Multipath ultrasonic flowmeter.

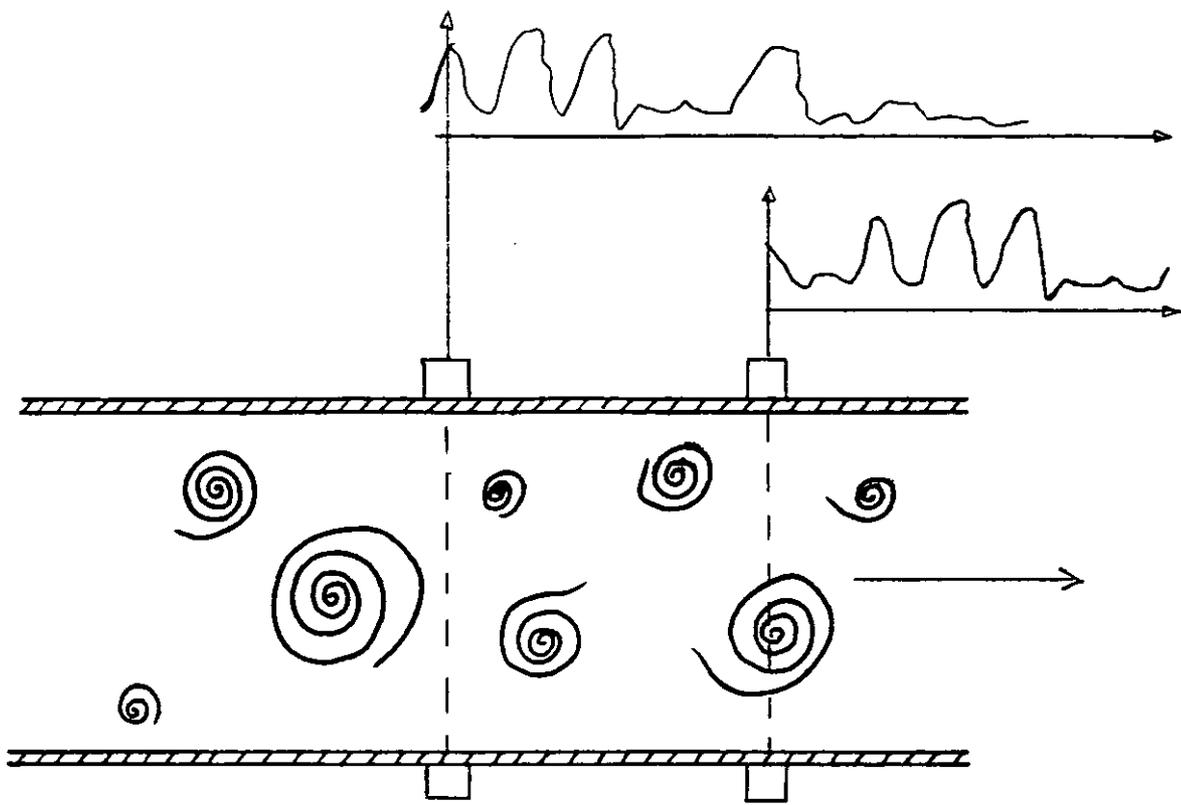


Figure 8. Ultrasonic crosscorrelation flowmeter.

FUTURE METERING SYSTEMS
SONIC NOZZLES

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FIGURES

1. INTRODUCTION

The phenomenon of critical flow, although established analytically and experimentally before 1900, did not become popular in flow measurement until the 1950's. At that time critical flowmeters were used as a test device in performance evaluation of gas turbines and within the aerospace industry in its rocket propulsion projects. From these applications came an optimized critical flow meter, the critical flow venturi, also called sonic nozzle. After many years of effort in different laboratories in Europe and USA, the International Standardization Organisation has decided to publish a draft international standard (DIS) for measurement of gas flow by means of critical flow venturi nozzles.

2. TYPICAL FEATURES OF A CRITICAL FLOW VENTURI NOZZLE

A critical flow venturi nozzle (CFVN) or sonic nozzle (SN) is shown in figures 1 and 2. It consists of an inlet convergent, a toroidal or cylindrical throat and a divergent outlet. A CFVN is a constriction type flow meter in which the phenomenon of critical flow occurs when the gas velocity in the throat of the meter is accelerated to the local value of the velocity of sound.

2.1. Advantages

System simplicity

The CFVN requires less instrumentation and is less sensitive to installation and pulsation errors than subcritical devices. As long as critical flow conditions are maintained, downstream perturbations are not propagated upstream and do not introduce any error in the flow rate measurement. Only pressure and temperature need to be recorded in order to predict the flow rate through the CFVN.

Recovery

Due to the divergent outlet section downstream of the throat the CFVN is able to recover up to 90% to 95% of the stagnation pressure.

Repeatability

The CFVN has no movable part which may create performance drift. Experiences have demonstrated repeatability well within +0,1%.

Predictability

The flow rate through a CFVN can be theoretically predicted with high enough accuracy for all practical purposes.

2.2. Limitations

Rangeability

The rangeability of the CFVN is limited to the possible variations of inlet pressure. It is convenient when large flow rates are required to instal a battery of CFVN as shown in figure 3.

Pressure loss

Critical flow conditions in the nozzle generate pressure loss of at least 5% to 10% of upstream pressure. For system design purpose this pressure loss should be put to 10%.

Real gas properties

For real gas calculations there is still a lack of knowledge, but for design purpose the model developed by R.C. Johnson prevails (1,2).

Construction

Due to the particular inlet, throat and outlet sections the construction requires very skillful machinist to build and inspect according to the tolerances. The larger the throat diameter, the easier the construction.

3. THEORETICAL CONSIDERATIONS

3.1. Ideal conditions

The theoretical calculations of flow through a CFVN is based on three main hypothesis:

- the flow is one-dimensional
- the flow is isentropic
- the gas is perfect

If these conditions are satisfied, the value of the mass flow rate through the CFVN is predicted to be:

$$q_{mi} = A^* C_i^* p_o \left(\sqrt{\frac{R}{M} T_o} \right)^{-1} \quad (1)$$

where

C_i^* is the critical flow function for one dimensional isentropic flow of a perfect gas, defined as:

$$C_i^* = \left[\gamma \left(\frac{2}{\gamma-1} \right)^{\frac{\gamma+1}{\gamma-1}} \right]^{\frac{1}{2}} \quad (2)$$

The other parameters are defined in the nomenclature.

3.2. Real conditions

In order to compensate for the deviation from non-one dimensional and non isentropic flow one introduces the notion of discharge coefficient, C . The discharge coefficient of a CFVN is less than unity since the flow is not one-dimensional and a boundary layer exists due to viscous effects. C may be determined by direct calibration or from an empirically determined function of the Reynolds number.

By direct calibration the nozzle discharge coefficient is obtained from the equation:

$$C = \frac{q_{act}}{q_{mi}} \quad (3)$$

where

q_{act} is the actual mass flowrate calculated from a primary calibration. From series of primary calibrations the following equation has been developed:

$$C = a - b Re_d^{-n} \quad (4)$$

where

Re_d is the CFVN throat Reynolds number defined as:

$$Re_d = \frac{4q_m}{\pi d \mu_o}$$

The coefficients a and b are given in table 1. They are the same as those indicated in the forthcoming DIS for CFVN (3).

The flowrate, q_m , in real conditions becomes:

$$q_m = A_* C C_* P_O \left(\sqrt{\frac{R}{M} T_O} \right)^{-1} \quad (5)$$

where

C^* is the critical flow function for one dimensional gas flow.

$$q_m = A_* C C_R \sqrt{(P_O \rho_O)} \quad (6)$$

where

C_R is the real gas critical flow coefficient defined as

$$C_R = C_* \sqrt{Z} \quad (7)$$

R.C. Johnson (2) has developed an empirical correlation for the critical flow metering of natural gas mixtures valid up to 70 bar. According to Johnson:

$$C_R = a_c f + b_c \quad (8)$$

the compressibility, Z , being calculated by means of the Benedict, Webb, Rubin equation of state.

The coefficients a_c and b_c are given in tables 2 and 3 as function of pressure and temperature.

The gas composition factor, f , is defined as:

$$f = -\frac{1}{2} X_{N_2} + X_{CO_2} + X_{C_2H_6} + 2 X_{C_3H_8} + 3 X_{C_4H_{10}} \quad (9)$$

where

X is the mole fraction of the gas whose chemical symbol appears in the subscript.

According to Johnson the composition factor, f , should be in the range 0 to 0.2. It should be noted that the composition range of the gas mixture, where this correlation is applicable, is quite limited. Table 4 gives the permissible range for all components in mole fractions. Although it has been recognized that this technique is not as accurate as one would desire, it is the only one which is currently available for practical use.

4. STANDARD CFVN AND INSTALLATION REQUIREMENTS

As indicated in figures 1 and 2, CFVN consists of a convergent inlet followed by the throat and divergent outlet. The divergent is shaped to provide a maximum pressure recovery.

A large number of inlet shapes and throat geometries have been proposed and studied. Discussions by ASME and ISO committees have reduced these shapes down to two which are currently considered as standard devices (3).

One is proposed by Hillbrath (4); it is identical to the design of Smith and Matz (5). It consists of a toroid of $R/d=2.0$, and a contraction ratio of 2,5. This device has no cylindrical throat and the diffuser is a cone of 4 degree half angle tangent to the continuation of the inlet toroid. Brain and Reid (6), Arnberg et al. (7) and Stratford (8) have also largely contributed to the toroidal CFVN development and design.

The second one is based on investigations performed by Jaumotte (9), Castillon (10), Masure et al. (11), Grenier (12) and Peignelin (13). It consists of a quarter of a torus tangent both to the inlet plane and to the cylindrical throat. The radius of curvature of the torus is equal to the throat diameter. The cylindrical throat has a length equal to the throat diameter. The diffuser is the same as for the toroidal throat venturi.

The inlet conduit up to 3 pipe diameter (3D) upstream of the venturi nozzle shall not deviate from circularity by more than 1% of its diameter and shall have an average roughness height which shall not exceed $75 \cdot 10^{-4} \cdot D$ (m/m) of conduit diameter.

To avoid corrections for the dynamic pressure upstream of the nozzle, a CFVN shall never be used with a diameter ratio (throat/pipe) larger than 0.25 when placed in a circular conduit. In any other case it is recommended that the throat area shall never exceed 6% of the upstream area.

Pressure taps for upstream static pressure measurements shall be located 0.9 -1.1 D from the inlet plane of the nozzle. The diameter of pressure tap should preferably be 1.3 \pm 0.3 mm. Downstream pressure shall be measured with a pressure tap 0.5 D downstream of exit plane. Temperature shall be measured 2D upstream and the diameter of the sensing element should not exceed 0.04 D.

5. UNCERTAINTIES

5.1. General

As mentioned previously, operation at critical flow conditions is characterized by a continuous acceleration of the flow from the venturi inlet to some location downstream of the throat.

Figure 4 shows an ideal Mach number distribution along venturi length at typical subcritical and critical flow conditions. Ideal weight flow rate per unit area at the venturi throat is shown in figure 5 as a function of throat Mach number and throat pressure ratio. Variation of venturi throat static pressure as a function of maximum venturi Mach number is indicated in figure 6. Operation at critical conditions, as compared with operation at subcritical conditions, results in a marked reduction of the error in flow rate resulting from errors in venturi pressures.

The rate of change of air flow with respect to Mach number is large at low Mach number (Ma from 0.2 - 0.4). The rate of change is zero at Ma=1. At critical flow conditions the throat static pressure is constant.

5.2. Draft international standard

In the DIS document (3) it is indicated that the relative uncertainty of the discharge coefficients calculated according to equation (4) is $\pm 0.5\%$. This uncertainty is for a confidence level of 95%.

European laboratories as NEL and Gaz de France which operate tests facilities for primary calibration of CFVN (gravimetry and volumetry) claim an accuracy better than ± 0.25 to $\pm 0.3\%$ on their discharge coefficients.

5.3. Uncertainty calculations

Details for practical uncertainty calculations are indicated in the standard ISO-5168: Estimation of uncertainty of a flow-rate measurement.

The practical working formula for mass flow uncertainty calculations is:

$$E_{q_m}^2 = 4 E_d^2 + E_c^2 + E_{p_o}^2 + \frac{1}{4} E_{T_o}^2 + \frac{1}{4} E_M^2 + E_{C^*}^2$$

Assuming the following relative uncertainties:

E_d	=	0.05%	throat diameter
E_c	:	see table 5	discharge coefficient
E_{p_0}	:	see table 5	stagnation pressure
E_{T_0}	=	0.15%	stagnation temperature
E_M	:	see table 5	molecular weight
E_{C^*}	:	see table 5 and figure 7	critical flow factor

Based on these above indicated uncertainties, the relative error on the flow rate has been calculated. Table 5 shows that e_q varies between $\pm 0.45\%$ and $\pm 0.76\%$

6. APPLICATIONS

6.1. General

The CFVN has two main application fields. One concerns secondary standards for gas flow meter calibration and control, the second is for turbine testing.

Turbine testing requires the accurate measurement of air flow. CFVN's are normally used because they are approximately 3 times more accurate than subsonic metering devices (15).

The CFVN has been used as secondary standard by Gaz de France who uses this type of device to control the flowmeters installed on their grid. In UK NEL has actively promoted its use within gas metering.

In the US, the Natural Gas Pipeline Co. of America has also used CFVN to verify their line meters (16).

Besides, tests have shown that the method of using a set of sonic nozzles (figure 3); arranged in parallel in a package of short length, can prove to be a particularly effective means of obtaining performance traceability for flowmeters which measure flow rates well in excess of those which can be covered on existing primary standard test facilities.

6.2. Practical criteria

When a CFVN is intended for use, for instance with a natural gas whose composition factor, f , is within the previously mentioned validity range (0 to 0.2), the following procedure is recommended:

- determine roughly the nozzle capacity, i.e. A^* . Figure 8 shows the mass flowrate variation as function of stagnation pressure for different throat diameters.
- manufacture or buy the right nozzle (either with toroidal or cylindrical throat). Refer to DIS recommendations.
- have the CFVN primary calibrated in order to determine the discharge coefficient. Alternatively determine the discharge coefficient through a secondary calibration or by means of equation (4).
- apply Johnson method to calculate the real gas critical flow coefficient, C_p and use an appropriate state equation to calculate Z .
- calculate the mass flow rate from equation (5) or (6).

Particular attention must be paid to the gas composition. The above mentioned formula is only valid for gases whose composition corresponds to what is indicated in table 4. For other gases and for pressures higher than 70 bar there is no methods which are directly applicable. However, the fundamental procedure used by Johnson is a general one and can also be used for other gas compositions. It should be noted that when the gas contains important quantities of heavy components (for instance more than 0.4% of C_4) precautions should be taken to prevent possible condensation effects.

7. CONCLUSIONS

Recommendations have been given for two types of standardized CFVN. Essential features of these designs are given as indications. A final document on that topic will be issued by the International Organization for Standardization (3).

The CFVN is normally not suited for on line flow rate measurements in field installations, but it is very useful as secondary standard for gas flow meter calibration and for verification of line meters (16). The CFVN is used for testing of turbines (15).

The accuracy and the repeatability are two main advantages of the CFVN. The mass flow rate through a CFVN is easily predicted from theoretical calculations. Uncertainties of the order of $\pm 0.7\%$ on the flow rate may be obtained when an $\pm 0.5\%$ uncertainty on the discharge coefficient is considered. Improvement on the flow-rate accuracy may be obtained by direct calibration of the CFVN. In this case accuracy better than $\pm 0.5\%$ is achieved.

Methods are available to calculate the mass flow rate through a CFVN for natural gas mixtures which have up to 0.4% of C_4 components.

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1. NOMENCLATURE

The nomenclature used in this report is shown below:

A^*	Area of critical flow venturi nozzle throat	L^2	m^2
C	Coefficient of discharge for the venturi nozzle	Dimensionless	
C_d	Coefficient of discharge for the orifice	Dimensionless	
C^*	Critical flow function	Dimensionless	
C_i^*	Critical flow function for one dimensional flow of a perfect gas	Dimensionless	
d	Diameter of orifice or throat of primary device	L	m
D	Upstream internal pipe diameter	L	m
e_x	Absolute uncertainty of the quantity X	$[X]$	
E	Velocity of approach factor $E=(1-\beta^4)^{-\frac{1}{2}}$	Dim.less	
f	Gas composition factor	Dim.less	
k	Pressure loss coefficient	Dim.less	
\dot{m}	Total mass rate of flow in the loop	M/t	kg/s
Ma	Mach-number	Dim.less	
M	Molecular weight	M	kg
p_o	Stagnation pressure of the gas at nozzles inlet	$ML^{-1}T^{-2}$	Pa
p	Static pressure of the gas	$ML^{-1}T^{-2}$	Pa
Δp	Differential pressure	$ML^{-1}T^{-2}$	Pa
Q	Total volume rate of flow in the loop	L^3/t	m^3/s
q_m	Mass rate of flow through a CFVN	M/t	kg/s
q_v	Volume rate of flow through a CFVN	L^3/t	m^3/s
Re	Reynolds number	Dim.less	
R	Universal gas constant	$L^2t^{-2}T^{-1}$	$Nm/^\circ K$
T	Temperature of the gas	T	$^\circ K$
T_o	Stagnation temperature of the gas	T	$^\circ K$
U	Mean axial velocity of the fluid in the pipe	L/t	m/s
Z	Compressibility factor	Dim.less	

β	Diameter ratio, $\beta = \frac{d}{D}$		Dim.less	
γ	Ratio of specific heat capacities	••	Dim.less	
κ	Isentropic exponent		Dim.less	
μ	Dynamic viscosity of the gas		$ML^{-1} t^{-2}$	kg/ms^2
ρ	Density of the gas		ML^{-3}	kg/m^3
φ	Product coefficient $\varphi = A \cdot CC \cdot \sqrt{\frac{R'}{M} - 1}$		$Lt(MT)^{\frac{1}{2}}$	$ms(^{\circ}K/kg)^{\frac{1}{2}}$
θ	Sensitivity coefficient		-	
ε	Expansibility factor		Dim.less	

TABLE 1

Toroidal throat		Cylindrical throat	
$10^5 < Re_d < 10^7$	$a = 0.993\ 54$	$10^4 < Re_d < 4 \times 10^5$	$a = 1$
	$b = 1.525$		$b = 7.24$
	$n = 0.5$		$n = 0.5$
		$4 \times 10^5 < Re_d < 2.8 \times 10^6$	$a = 0.9886$ $b = n = 0$
		$2.8 \times 10^6 < Re_d < 2 \times 10^7$	$a = 1$ $b = 0.2215$ $n = 0.2$

TABLE 2

VALUES OF COEFFICIENT α_c

```

*****
* TEMP *           INLET STAGNATION PRESSURE - MEGAPASCALS           *
* DEG  *           *           *           *           *           *           *
* C    *           *           *           *           *           *           *
*****
* 0    * -.0293 * -.0331 * -.0371 * -.0407 * -.0437 * -.0452 * -.0442 *
* 5    * -.0298 * -.0336 * -.0373 * -.0408 * -.0436 * -.0452 * -.0447 *
* 10   * -.0304 * -.0340 * -.0375 * -.0409 * -.0436 * -.0452 * -.0450 *
* 15   * -.0309 * -.0343 * -.0377 * -.0410 * -.0436 * -.0451 * -.0452 *
* 20   * -.0314 * -.0347 * -.0380 * -.0412 * -.0436 * -.0451 * -.0454 *
* 25   * -.0319 * -.0351 * -.0383 * -.0413 * -.0436 * -.0452 * -.0456 *
* 30   * -.0324 * -.0355 * -.0385 * -.0414 * -.0437 * -.0452 * -.0457 *
* 35   * -.0328 * -.0358 * -.0387 * -.0414 * -.0437 * -.0452 * -.0458 *
* 40   * -.0332 * -.0361 * -.0390 * -.0416 * -.0437 * -.0453 * -.0459 *
*****

```

TABLE 3
VALUES OF COEFFICIENT b_c

```

*****
* TEMP *           INLET STAGNATION PRESSURE - MEGAPASCALS *
* DEG  *           *           *           *           *           *
* C    *           0 *           1 *           2 *           3 *           4 *           5 *           6 *
*****
*  0  * .6709 * .6708 * .6709 * .6714 * .6722 * .6737 * .6756 *
*  5  * .6707 * .6706 * .6708 * .6713 * .6722 * .6736 * .6755 *
* 10  * .6704 * .6704 * .6706 * .6712 * .6721 * .6734 * .6753 *
* 15  * .6701 * .6702 * .6704 * .6710 * .6720 * .6733 * .6751 *
* 20  * .6699 * .6699 * .6702 * .6709 * .6718 * .6731 * .6749 *
* 25  * .6695 * .6697 * .6700 * .6706 * .6716 * .6729 * .6746 *
* 30  * .6692 * .6694 * .6698 * .6704 * .6714 * .6727 * .6744 *
* 35  * .6689 * .6691 * .6695 * .6702 * .6712 * .6724 * .6741 *
* 40  * .6686 * .6688 * .6693 * .6700 * .6709 * .6722 * .6738 *
*****

```

TABLE 4

Methane	0.840 - 1.000
Ethane	0 - 0.11
Propane	0 - 0.020
2-Methyl Propane	0 - 0.004
Butane	0 - 0.004
Nitrogen	0 - 0.023
Carbon Dioxide	0 - 0.017

E_c	E_{p_0}	E_M	E_{c^*}	E_{q_m}
0.3	0.1	0.25	0.25	0.44
			0.40	0.54
		0.5	0.25	0.49
			0.40	0.58
	0.2	0.25	0.25	0.47
			0.40	0.57
		0.5	0.25	0.52
			0.40	0.61
	0.3	0.25	0.25	0.52
			0.40	0.61
		0.5	0.25	0.57
			0.40	0.65

0.5	0.1	0.25	0.25	0.59
			0.40	0.67
		0.5	0.25	0.63
			0.40	0.71
	0.2	0.25	0.25	0.62
			0.40	0.69
		0.5	0.25	0.66
			0.40	0.73
	0.3	0.25	0.25	0.66
			0.40	0.73
		0.5	0.25	0.69
			0.40	0.76

TABLE 5 RELATIVE UNCERTAINTIES

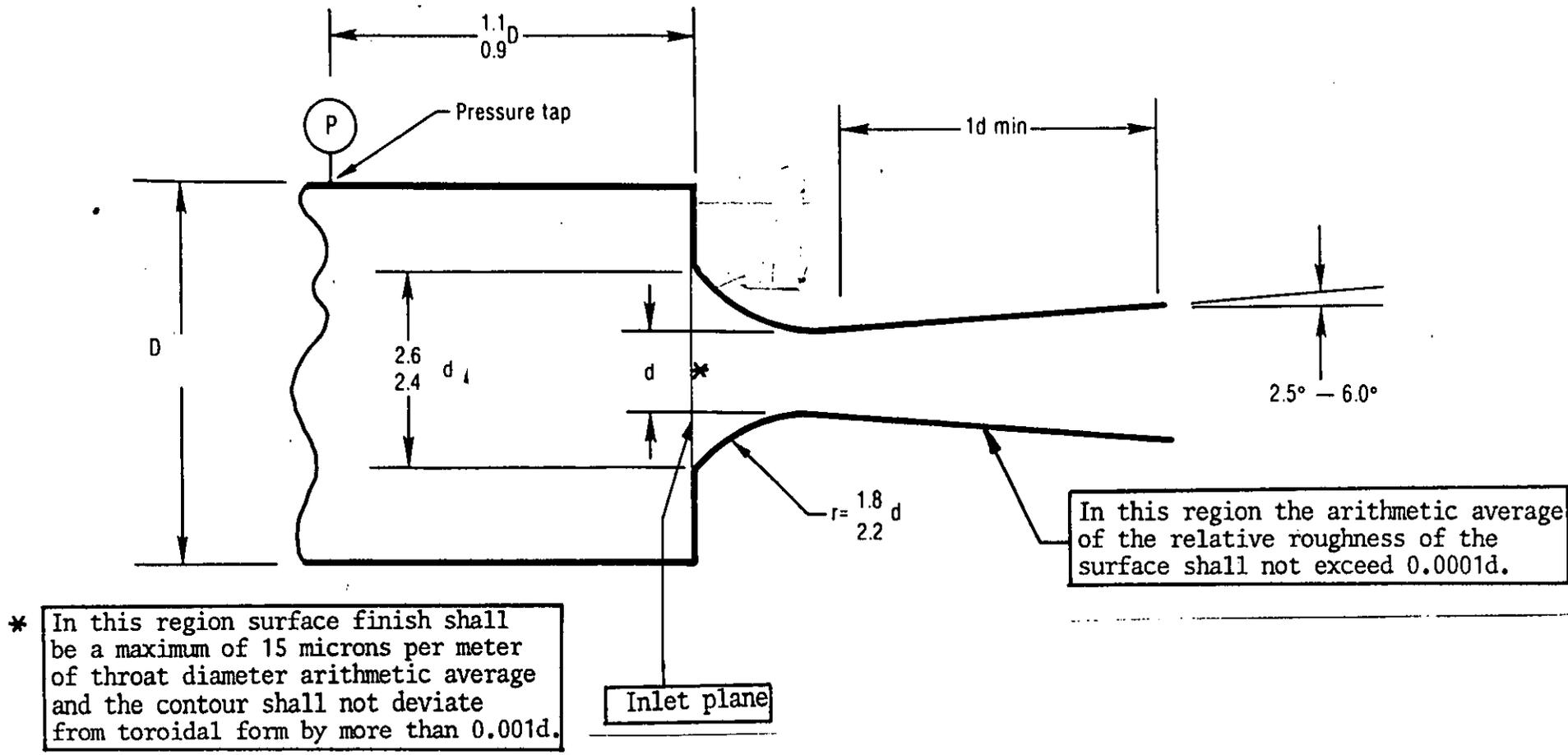


FIGURE 1 TOROIDAL THROAT VENTURI NOZZLE

* In this region surface finish shall be a maximum of 15 microns per meter of throat diameter arithmetic average and the contour shall not deviate from toroidal and cylindrical form by more than 0.001d.

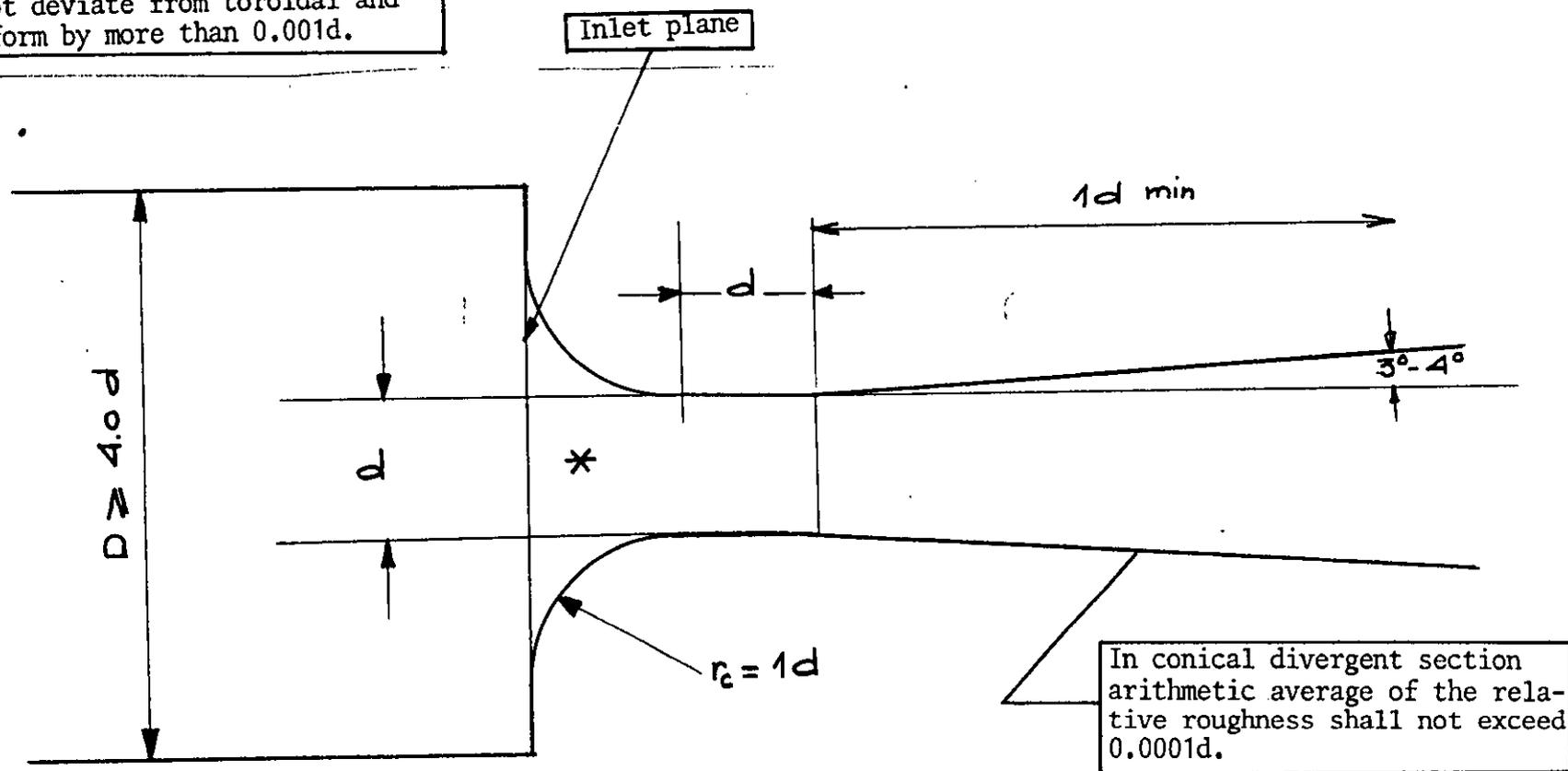


FIGURE 2 CYLINDRICAL THROAT VENTURI NOZZLE

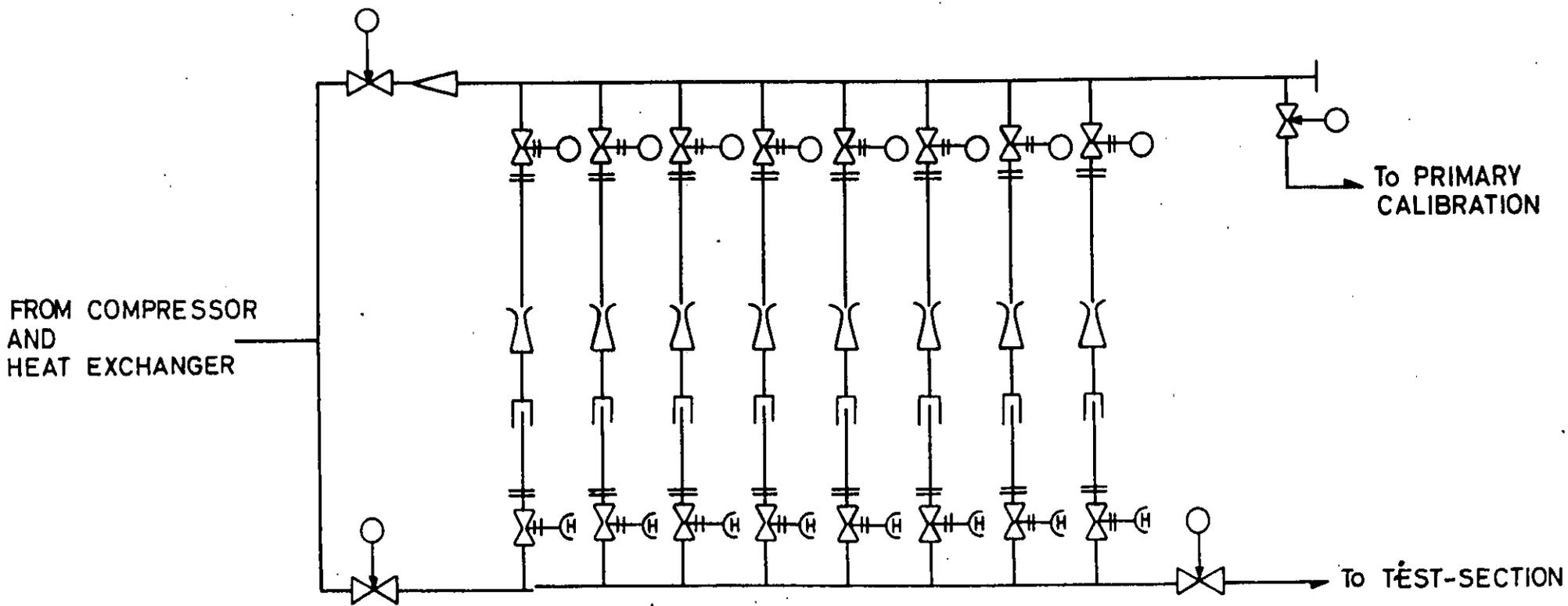


FIGURE 3 BATTERY OF VENTURI NOZZLES

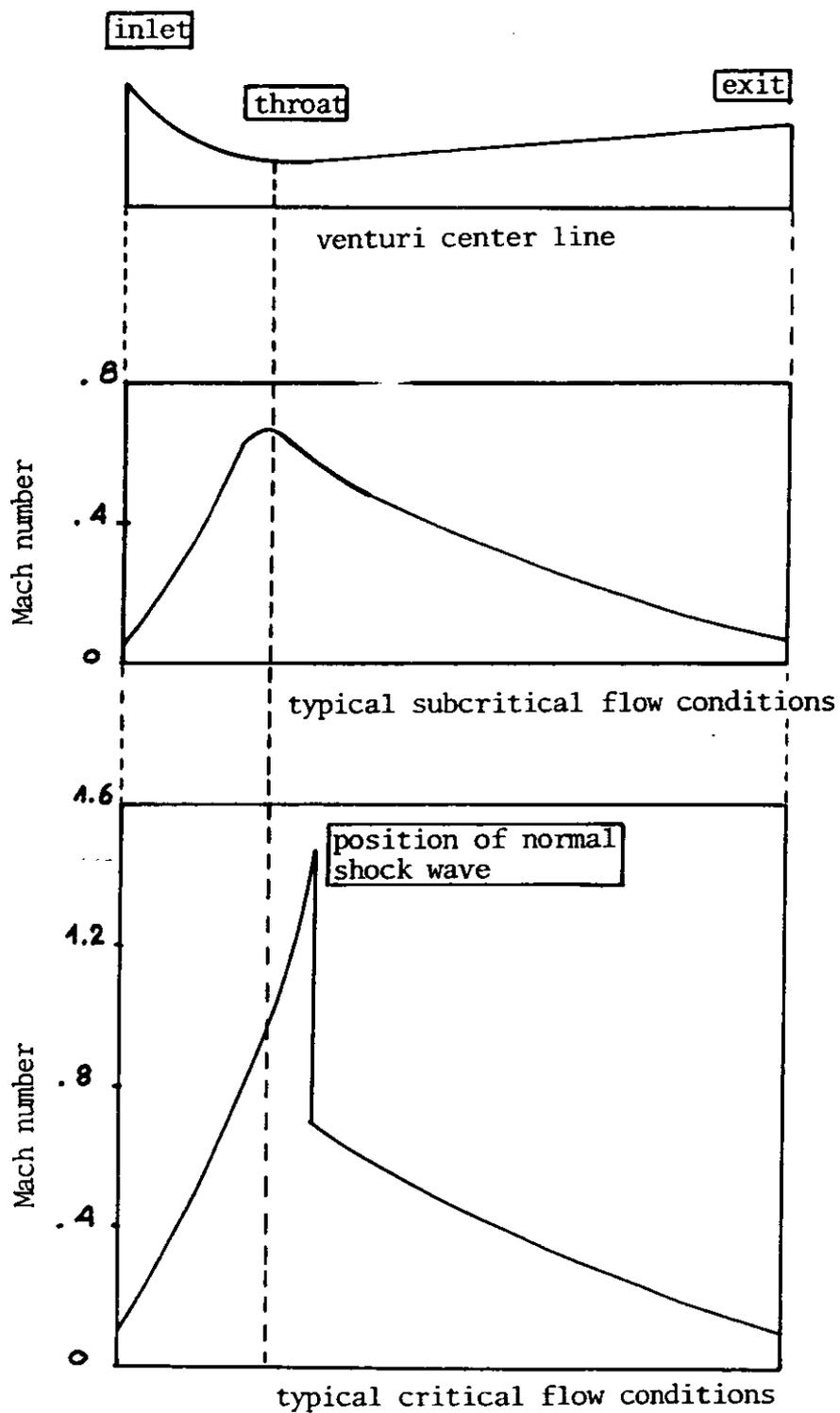


FIGURE 4 IDEAL MACH NUMBER DISTRIBUTION ALONG VENTURI LENGTH AT TYPICAL SUB/CRITICAL FLOW CONDITIONS (ref.5)

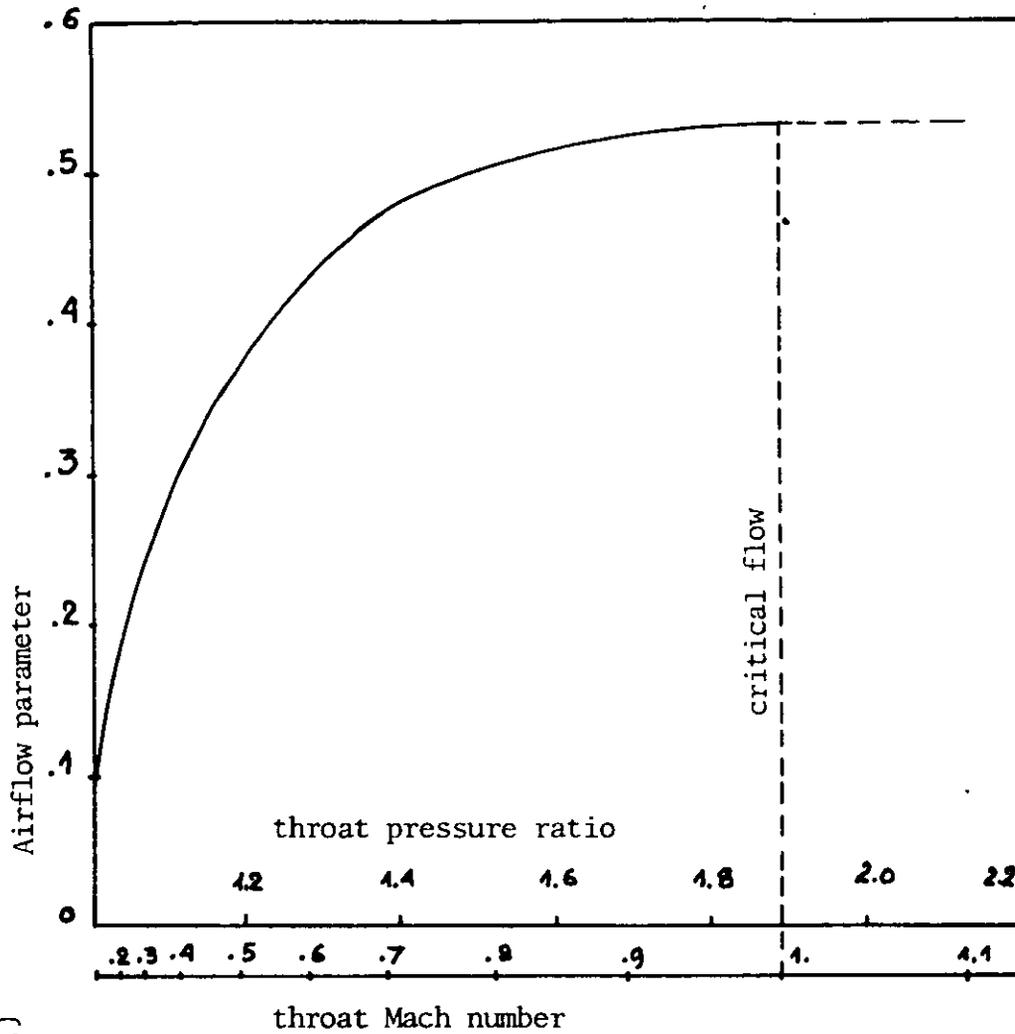


FIGURE 5
(ref. 5)

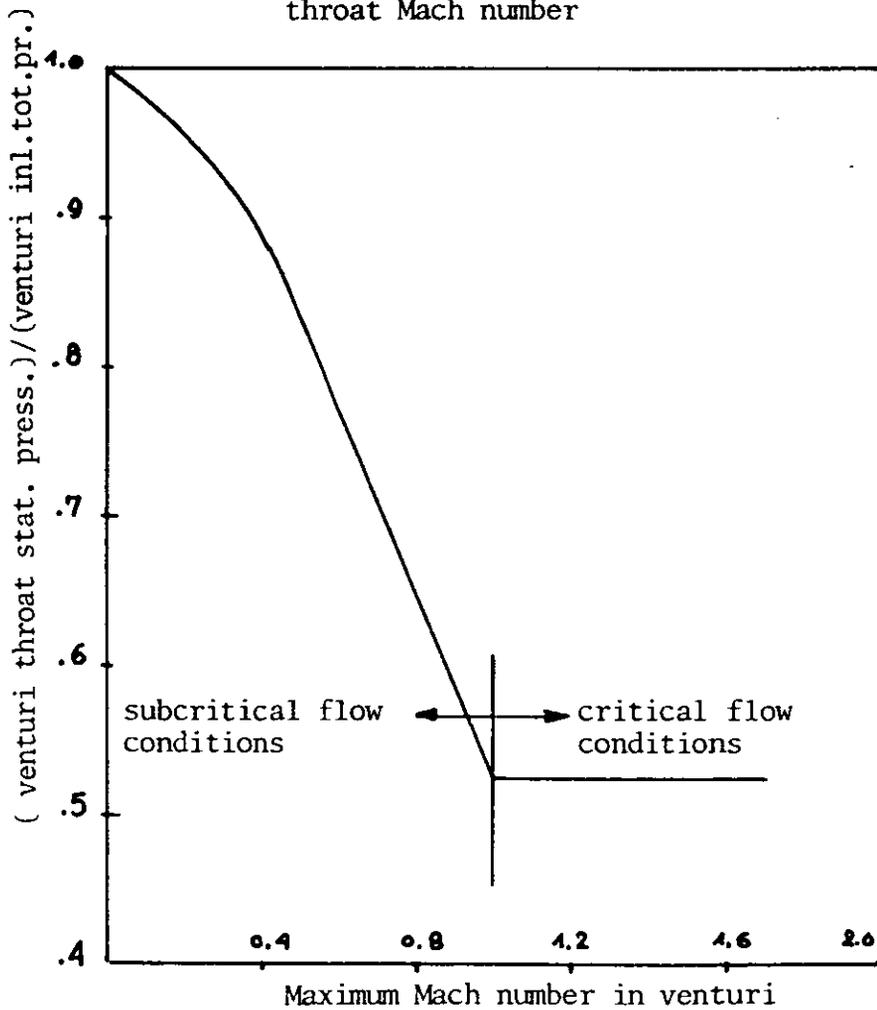


FIGURE 6
(ref. 5)

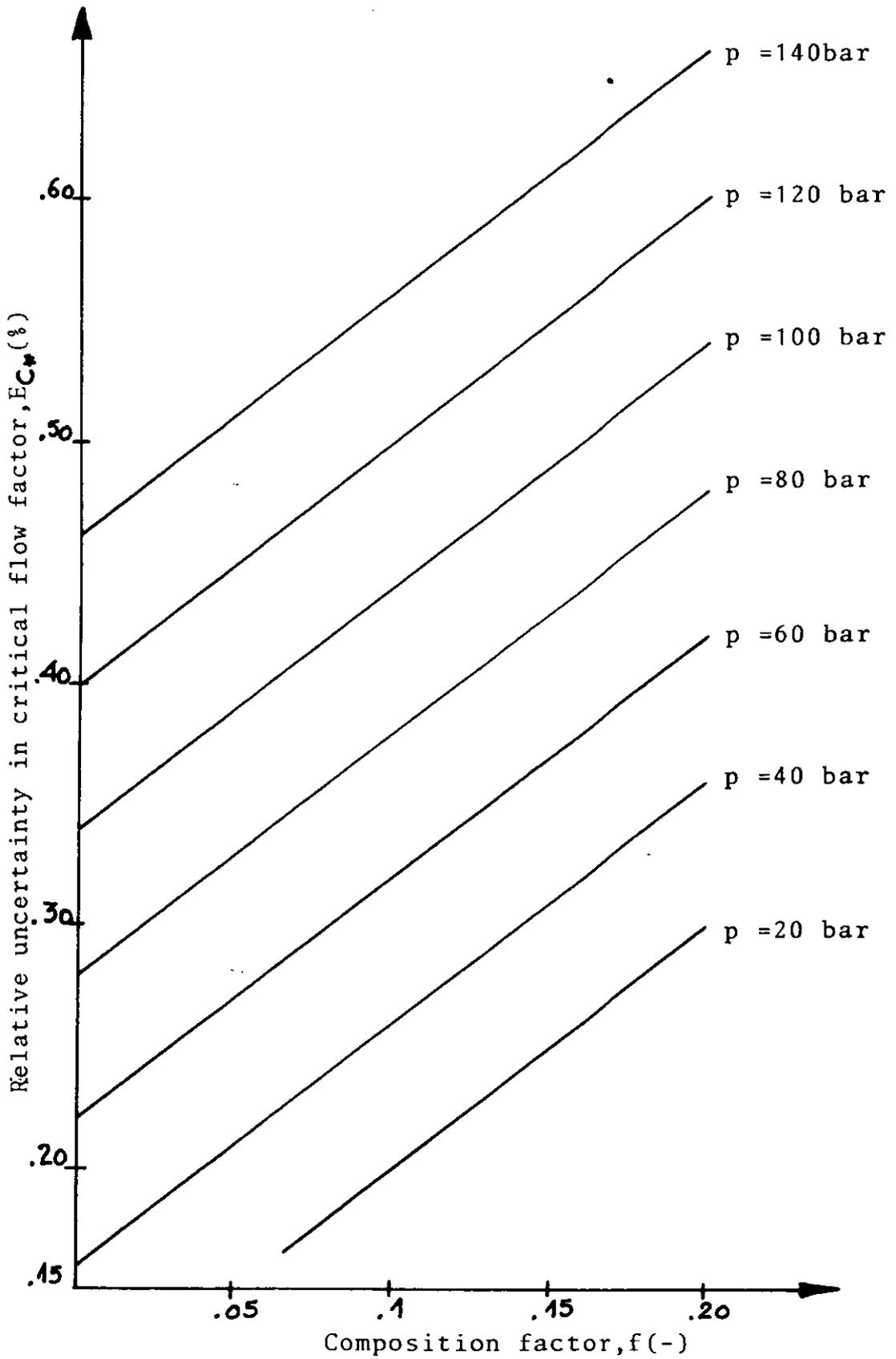


FIGURE 7 RELATIVE UNCERTAINTY IN CRITICAL FLOW FACTOR VS. THE COMPOSITION FACTOR AT DIFFERENT PRESSURES.

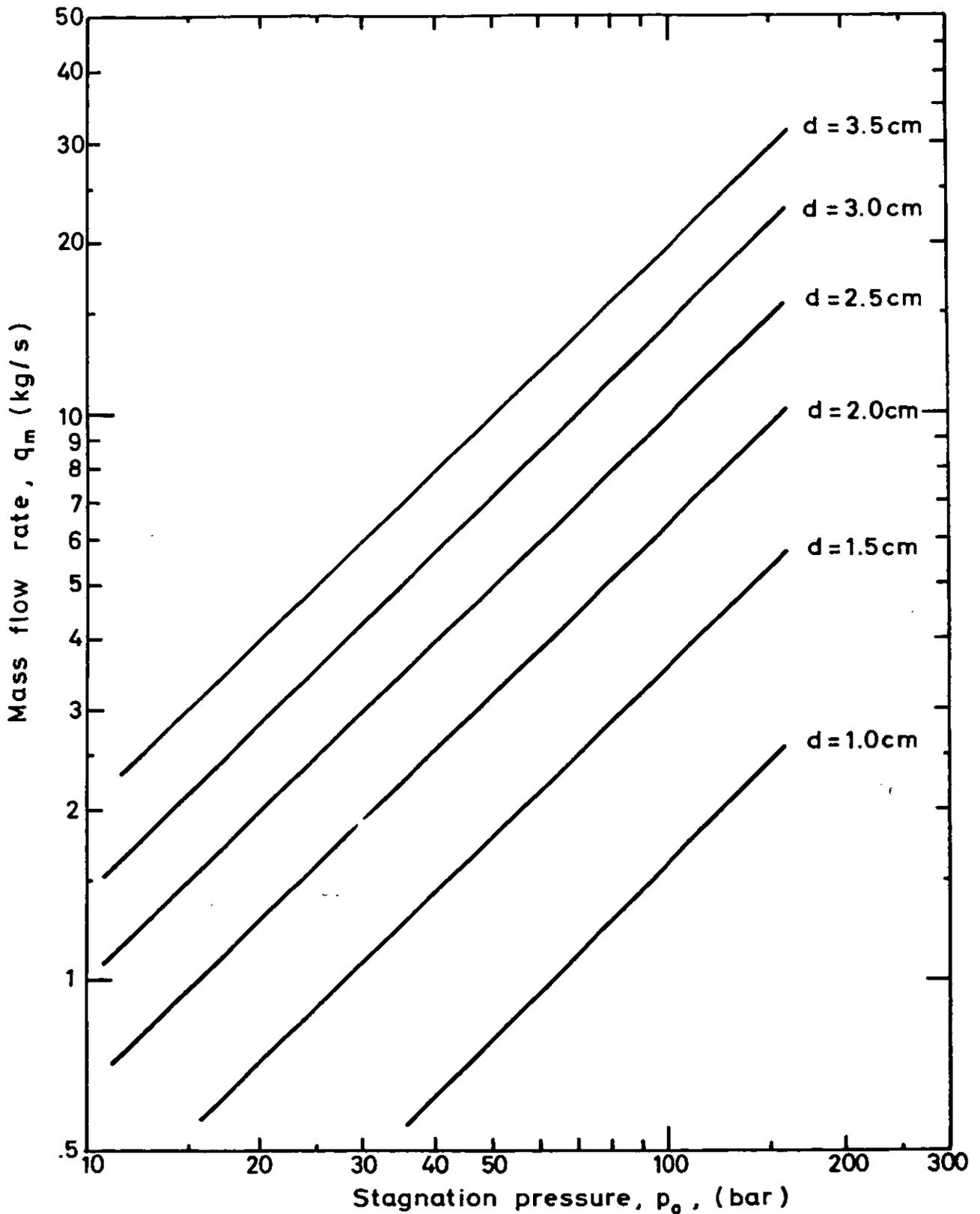


FIGURE 8.

MASS FLOW-RATE AS FUNCTION OF STAGNATION PRESSURE FOR DIFFERENT THROAT DIAMETERS. ($T_0 = 30^{\circ}\text{C}$). THE CURVES ARE ONLY FOR ROUGH ESTIMATES.