G.G. Nasr · N.E. Connor

Natural Gas Engineering and Safety Challenges

Downstream Process, Analysis, Utilization and Safety

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Preface

Natural gas has been a valuable energy commodity for many centuries. According to Encyclopaedia Britannica, the ancient Chinese were the first to observe the seeps and the use of natural gas in 600 BC with the first utilisation of it in the home during the great empire of Persia, now Iran, in 100 AD. It was not until 1803–1812 when the first gas lighting was patented in London by Frederick Winsor and the first company was then founded in London, England. It was during this century (19th) that natural gas for home lighting was also used directly from the wellhead in Fredonia, New York. Although natural gas was unpopular prior to the eighteenth century due to the use of manufactured gas such as 'coal gas', it was towards the latter part of the nineteenth century that most industrial countries started using natural gas and thus large transmission and distribution pipelines were constructed in transferring the gas to the required areas. The steady utilisation of natural gas grew to the peak during the 1960s to mid-1970s when the shortage of crude oil enforced most industrial nations to find alternative ways of harnessing energy and natural gas has since become one of the main fossil fuel energy sources. Natural gas is colourless with high flammability and energy value and together with its convenience has resulted in a rapid rise to extensive use as a fuel today.

As the utilisation of natural gas became more frequent as one of the main alternative choice of energy source it enabled rapid technological advancement and attainment of knowledge and understanding in various related disciplines of natural gas. Particularly over the last decades, there has been constant progress in research and innovation with regard to the production of natural gas, transmission, distribution, utilisation, safety and management in both upstream and downstream processes. The authors, whose backgrounds are outlined below, independently recognised that whilst there are numerous academically orientated books as well as conference publications and standards available that address the upstream process of natural gas and certain specialised texts addressing narrower areas of application, there is an absence of an academically and industrially oriented book that covers, as far as possible, the downstream process, that is, after the wellhead to gas processing plants and finally to consumers.

The book starts in a logical manner with the opening Chap. 1 describing the fundamentals of natural gas. Subsequent to the wellhead the gas must be transmitted and distributed to its final destination, that is, the consumers. These transmission and distribution processes require thorough understanding of their systems and design which are described in Chap. 2. The gas should also be stored or transported for later use as Liquefied Natural Gas (LNG), this is exposed in Chap. 3. Natural gas which contains certain physical characteristics should flow through various transmission and distribution designed network systems as is described in Chap. 4. No matter where the gas is being transferred to, the accurate control of it together with understanding its quality are pertinent which could eventually reflect on the overall capital expenditures of the gas. It is thus with this in mind that a comprehensive understanding of instrumentation and measurement systems have been provided in Chap. 5. Although natural gas has become one of the main energy sources, the accidental release and subsequent ignition of flammable gas and vapour clouds has led to a number of incidents with catastrophic consequences on oil and gas platforms. Chapter 6 therefore provides inclusive understanding of fire and explosion and safety aspects, where appropriate, of the natural gas. The utilisation of natural gas, including an overview of the heat transfer and heat exchangers, has also been given in Chap. 7. Within almost all the downstream processes the viability of the natural gas is dependent on how the gas business and the related projects should be managed and sustained, which is the subject matter of Chap. 8. In the last Chap. 9, the authors have provided various innovation management models from their own experience and borrowed from various disciplines, with a few case studies which over the last decade have become vital ingredients in the future sustainability of the gas industry.

Prior to commencing their cooperation on this book, GGN and NEC cooperated for many years as committee members of Institution of Gas and Mangers (IGEM) in Continuous Professional Development (CPD) and organized various conferences and short courses in gas safety and technology. Also, GGN and NEC had cooperated in research and consultancy projects, particularly involving gas processing and metering systems. The incentive to cooperate and write the book came from frequent requests from those in academia and industry for a text that was suited to their applications-oriented needs in the downstream process and yet which covered a wide breadth of knowledge. Although, together, the authors have experience in a wide range of gas engineering and safety applications, the very large number of concerns that exist in industry has meant that expertise has been sought from specialist companies and individuals, where appropriate. These specialists are thanked in a later section of this introduction.

Ghasem G. Nasr is Professor of Mechanical Engineering and Innovation at University of Salford, Manchester, England and he is the head of Petroleum and Gas Engineering in the School of Computing, Science and Engineering (CSE) and Director of sprays and petroleum technology research groups. Graduating in Mechanical Engineering, Fuel Technology (PgDip) and Energy Science (MSc) from the University of Middlesex and Sussex University, respectively, and Heat Transfer and Fluid Mechanics (Ph.D.) from the University of Swansea, he moved

into project engineering management and energy utilization between 1988 and 1995 to Tata Steel, Wales, and subsequently acted as a Senior Consultant in Europe and the Middle East before moving to the University College of Manchester (now Stockport College) managing the engineering department. He then joined the University of Salford in 2001 as a leader of Gas Engineering and Management. He has over 25 years' experience in innovation and research in many areas of spray production and gas utilisation and has been consultant to over 120 companies in New Product Development (NPD), innovation and research. He was a founder of 'Spray Research Group' and has authored 130 papers, including editorship of books and journals, and acquired 7 patents and was a lead author of the book *Industrial Sprays and Atomisation (SV, 2001).* He has also held a number of executive appointments in various academic, professional and steering committees. Currently, he is a member of PDC of IGEM and Executive Member of steering committee of Praxis-Global. He has also been guest speaker on many international platforms and delivered over 100 advanced short courses in various related subjects. He is a Chartered Engineer, a Fellow of the Institution of Gas Engineers and Managers (FIGEM) and of the Institute of Mechanical Engineers (FIMechE) and Member of FEANI EurIng and Institute of Liquid Atomisation and Spraying Systems (ILASS). Recently, he has been appointed as a director of Technology and Innovation at the Salvalco Ltd.

Norman E. Conner joined the North Western Gas Board as a student engineer in 1950 and began studying Mechanical Engineering at Warrington and St. Helens Technical Colleges. In 1953 he was awarded a Whitworth Society prize and a Technical State Scholarship to study Gas Engineering at the University of Leeds, graduating in 1957 with an Honors B.Sc., Degree. He returned to the North Western Gas Board as a production engineer in the South Lancashire group and was appointed Chief Chemist at the Warrington Production Station in 1958. In 1964 he left to take up an appointment as a Lecturer in Gas Engineering at the Royal College of Advanced Technology, Salford. He is a Chartered Engineer, a Fellow of the Institution of Gas Engineers and Managers and of the Energy Institute. He represented the University on the IGEM Education, Training and Academic Committees for many years and has been Chairman of the North Western Section of E.I., and the Manchester Gas Association. He received an M.Sc., Degree and Senior Lectureship in 1977. During his career he was actively involved in running Conferences/Symposia in gas engineering, fuel utilisation and chemical engineering at the University. He is also a co-author of: Industrial Gas Utilisation Engineering Principles and Practice Bowker 1977. He was involved in setting up and teaching on the M.Sc., course in Gas Engineering and Management with colleagues (A.L. Bowler and Dr. R. Pritchard) in the late 1980s, later becoming Course Director. He is still actively involved in the Gas and Petroleum Engineering programme at the University.

Acknowledgments

This book reflects a total of some 80 years' experience of the authors in gas engineering and utilisation in both industry and academia. The successful completion of the book, however, must be shared with those who provided the authors with invaluable advice and material. These are from a number of key companies involved in the field of gas engineering. The authors are very grateful for the willing cooperation of these companies and individuals within them. Specifically these include the following, where the sections of the book particularly relevant to each company are given in brackets:

- Scotia Gas and Network, Mr. David Macleod (Sect. 4.2)
- Prakash Bhikaji Morje, Engineering Officer, Shell Trading and Shipping Company, (Chap. 3)
- Dr. R. Pritchard, Previous Senior lecturer University of Salford (Sect. 6.1)
- Abubakar Abbas, Lecturer, Ahmadu Bello University, (Chaps. 1 and 2)
- Dr. Salah Ibrahim, Senior Lecturer, University of Loughborough (Chap. 7)

Where illustrations have been reproduced with permission from other sources, this is acknowledged in the titles of the figures. Again, the authors express their thanks for kind agreement of the copyright holders. A number of individual industrial and academic colleagues are also greatly thanked, who kindly used their expert knowledge in the final manuscript of various chapters. In addition, all our past and present students in the field whose comments and supports are gratefully acknowledged. They include Mr. David Macleod, Late Charles Hazel Dean, Lewis Mather, Mr. Steve Johnson and Andy Bowler (Chap. 5). Also, Dr. R. Pritchard for his encouragement and supports together with academic colleagues who made valuable contributions to various chapters. Particularly Dr. Salah Ibrahim from the University Loughborough who brought expertise and focal contribution to the Fire and Explosion Chap. 7, Mr. Alan Wells to Chap. 8 and Dr. Godpower Chimagwu Enyi to Chap. 9 from the University Salford. The authors are also grateful to the senior research assistant Mr. Abubakar Abbas (Ahmadu Bello University, Zaria, Nigeria) who brought a wealth of experience into Chaps. 1 and 2. The contribution of Dr. Amir Nourian of the University of Salford in the preparation of CAD

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Nomenclature

A	Cross sectional area (m^2)
A_p	Burner port area (mm)
BPL	Burner Port Loading (MW/m^2)
c	Constant (–) Eq. 2.1
CV	Calorific value (MJ/m^3)
C	Overall design coefficient (–)
C_d	Discharge coefficient (–)
cP	Centipoise (–)
D	Diameter (m)
d	Diameter (m)
E	Efficiency factor (–)
F	Force (kgm/s^2)
f	Friction factor (–)
g	Gravity (m/s^2)
h_f	Frictional head loss
h	Height of gas column (m) Eq. 4.5
IGU	International Gas Union (–)
K	Constant (–)
k_v	Velocity coefficient (–)
L	Length (m)
NTS	National Transmission Systems (–)
P	Pressure (Pa)
$psig$	Pound per square inch gauge (–)
ppm	Parts per million (–)
Q	Flow rate (m^3 (st)/hr)
q	Heat transfer rate (W/m^2)
Re	Reynolds number (–)
r	Air/gas ratio (–)
R	Molar gas constant (J/kg.K)
s	Specific gravity (–) Eq. 4.7
S_u	Burning velocity (m/s)

T	Temperature (K)
t	Thickness (s)
v_g	Gas rate (m^3/hr)
v_e	Erosional velocity (m/s)
x	Coordinate (m)
Z	Compressibility factor (m)

Greek Alphabets

δ_0 Film thickness (m)
ε Emissivity (–)
Δ Delta (–)
λ Wave length (m)
μ Dynamic viscosity (kg/ms)
ν Kinematic viscosity (m^2/s)
ρ Density (kg/m^3)
$\Delta\theta$ Temperature difference (K)
σ Stress (N/m^2)
τ Shear stress (N/m^2)

Subscripts

a	Air
c	Cold
g	Gas
h	Hot
i	Inlet
j	Jet
L	Liquid
m	Mass
m, 0.5	Refers to mass median value
max	Maximum value
p and q	Subscript in generalized mean diameter relationship
R	Refers to Rosin Rammler size distribution
st	Standard
s	Atmospheric
sp	Smooth pipe
v	Volume
o	Initial value

Note on Units

Unless stated in the text all equations are presented in SI units, that is, kg, m, s, K and derived units; J, N, Pa and W. In figures, tables and text more convenient or commonly used units may be used and these units are always clear, for example: μm, mm and kJ.

It is, however, normal practice in oil and gas industries that measurements are made using imperial units. This book adopted the SI units of measurement and where it was absolutely necessary, the imperial units were used. The following list provides some of the conversion factors of the units that were used in the respective text.

Some Relationships

Length; 10^6 μm (microns) = 1 km = 0.62137 mile = 3281 ft = 1000 m
Area; 1 m^2 = 10.76387 ft^2 = 1550 in^2
Volume/Capacity; 1 m^3 = 1000 litres = 6.28983 bbl = 1bbl/day = 0.1589873 m^3/day
Mass; 1t = 1000 kg = 2204.6223 Ib = 1 kg = 1000 g
Density; 1kg/m^3 = 62.428 Ib/ft^3 = 8.3304 Ib/gal
Viscosity Dynamic (μ); 1 $kgm^{-1}s^{-1}$ = 1 Pas = 10 Poise = 10^3cP (centipoise)
Viscosity Kinematic ($\nu = \mu/\rho$); 1 m^2s^{-1} = 10^4 Stokes (St) = 10^6cSt
Pressure 1 bar = 10^5Pa = 14.50 lb f $inch^{-2}$ = 750 mmHg = 10.2 m H_2O;
1Pa = 1 Nm^{-2}; Standard Atmospheric Pressure = 1 atm = 1.0132 bar
Force 1N = 1 $kgms^{-2}$ = 10^5 = dynes = 0.2248 lb f
Flowrate; 1 litre/s = 2.119093 ft^3/min
1 MMscf/day = 28316.847 m^3/day
Time; 1day = 24 hr = 86400 s
Velocity; 1 m/s = 3.2808399 ft/s
Power 1 kW = 1 kJs^{-1} = 1.341 HP (UK)
Energy; 1 kJ = 0.0002777 kWh = 0.947817 BTU
Temperature; K = °C + 273
°C = [(°F – 32)/1.8]

Acronyms

ACQ	Annual Contract Quantity
ALARP	As Low As Reasonably Possible
ANSI	Association National Standard Institute
API	American Petroleum Institute
ASME	American Society of Mechanical Engineers
BAT	Best Available Technique
BPL	Burner Port Loading
CEO	Chief Executive Officer
DEA	Diethanol Amine
GIP	Gas Innovative Portal
GS(M)R	Gas Safety Management Regulations
HAZID	Hazard Identification
HAZOP	Hazard and Operability Studies
ICO	Innovative Centre Office
ID	Innovation Director
IGEM	Institution of Gas Engineers and Managers
IGU	International Gas Union
IM	Innovation Manager
LFL	Lower Flammability Limits
LNG	Liquefied Natural Gas
LP	Low Pressure
LPG	Liquefied Petroleum Gas
MAPD	Major Accident Prevention Document
MEA	Mono Ethanol Amine
MSC	Metric Standard Conditions
MW	Molecular Weight
NTS	National Transmission System
NTU	Number of Transfer Units
ORV	Open Rack Vaporizers
PE	Poly Ethylene

PMO	Portfolio Management Office
PRS	Pressure Reduction Station
TEA	Triethanol amine
UFL	Upper Flammability Limit
UK	United Kingdom
W	Wobbe Number

Chapter 1
Fundamentals of Natural Gas

1.1 Background to Natural Gas

1.1.1 Introduction

At the present time, the gas industry in the UK and many other countries is based on the direct supply of natural gas. However, government and energy agencies are already planning for the major changes in energy supply which will be necessary in the medium-term future. It is clear that natural gas is a finite and depleting resource, but the rapid growth of the world's natural gas supply industries in recent years has proved beyond doubt the acceptability on economic and environmental grounds of gas transmission in pipelines over long distances, especially for high-pressure supplies of high calorific value (HCV) of the range 31–52 MJ/m^3 natural gas. Such gas can be stored in a variety of ways: as liquid natural gas (LNG), underground in depleted natural gas reservoirs, aquifers or salt cavities and by compression in the transmission system.

The foregoing means that there is every incentive to maintain gas industries by reverting to a manufacturing base, producing a substitute gas from the dirtier and less convenient fossil fuels, particularly coal because of its relative abundance, but also from crude oils, residual fractions and tar sands. Once the decision to manufacture a substitute gas has been made, the composition of that gas must be decided. It would be easier from a process standpoint to produce a gas with a Lower Calorific Value (LCV), nearer in composition to the town gases previously manufactured. However, apart from the need for appliance conversion, such a change would greatly reduce the capacity of the distribution system. A substitute natural gas consisting mainly of methane would therefore be preferred. This would allow maximum use of substitute gas to·be made in peak or seasonal load.

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1.1.2 Natural Gas Composition and Characteristics

1.1.2.1 Natural Gas Composition

Natural gas, as the name implies, can be defined as any gaseous material, usually combustible, and normally emerging from the ground either without outside assistance, purely under its own pressure, or from a bore hole drilled from the surface into an underground reservoir. Combustible gases are not very numerous and natural gas, therefore, invariably contains methane, usually ethane, sometimes propane and butanes. Figure 1.1 shows an approximate composition by volume. Other occasional combustible components are the unsaturated analogues of ethane, propane and butane, ethylene, propylene and the butylenes, hydrogen and carbon monoxide, hydrogen sulphide, ammonia which can be present in trace quantities.

Apart from combustibles, certain natural gases also contain inert gases, such as nitrogen, carbon dioxide and trace amounts of the inert gases such as helium, argon and xenon. Natural gas composition, to a considerable extent, depends on whether the gas, or condensate, is dry or produced in association with crude oil. In the former case, the main component will be methane and higher hydrocarbons heavier than ethane will only be present in traces; in the latter case, there will be a gradual transition from "permanently" gaseous (methane, ethane) to liquefiable under pressure (propane, butane) and to permanently liquid (pentane and heavier) hydrocarbons. Intermediate hydrocarbons such as propane and butane will thus generally be present in larger concentration in associated gas than in dry gas.

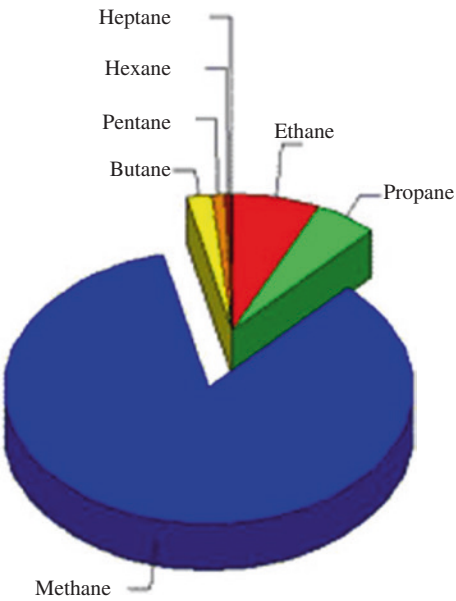


Fig. 1.1 Components of natural gas [1]

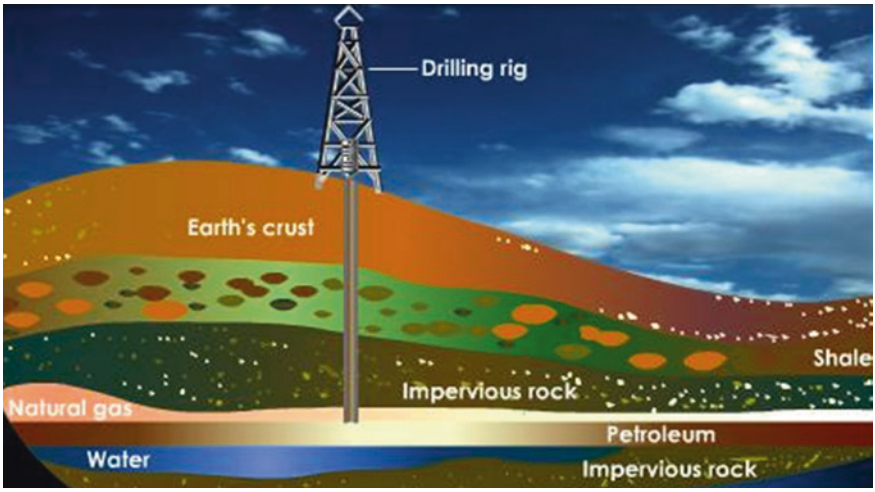


Fig. 1.2 Natural gas production and reserves profile [2]

The presence of inerts such as nitrogen and carbon dioxide is unpredictable and depends on local geological and geophysical conditions. A number of typical natural gas compositions are produced, i.e. before any possible treatment to remove impurities. A schematic diagram indicating the gas reserves and the drilling rig is shown in Fig. 1.2, with profile of different regions of earth crust.

Clearly, in order to arrive at a composition as close to uniformity as possible, removal of components other than methane and ethane would appear to be a reasonable approach. Unfortunately, this is not always possible, since, e.g. the separation of nitrogen from the remaining gases would require cooling to extremely low (−180 °C) temperatures. Other components such as carbon dioxide and hydrogen sulphide, which are acidic gases, can be removed by scrubbing the gas with an alkaline solution, preferably one which can be regenerated and reused. Heavier hydrocarbons, from propane upwards, can be condensed by compressing and cooling the gas and separating the uncondensed part, which is now mainly methane with some ethane, from the liquid components, largely propane/butane with some higher hydrocarbons

1.1.2.2 Natural Gas Characteristics

The importance of a well-defined and unchanging composition of natural gas is not unrelated to the fact that each component has certain chemical physical, physiological and performance characteristics which, in the case of a widely used fuel such as natural gas, should vary as little as possible. However, before we consider the numerical values of the different characteristics of the various components, let us consider for a moment which aspects of gas technology depend on

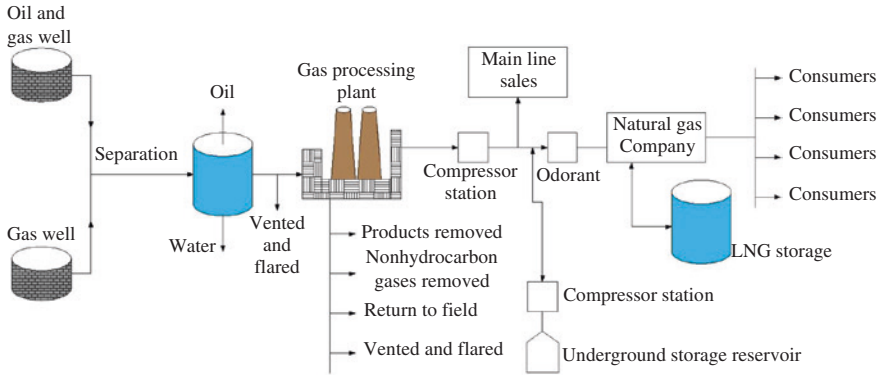


Fig. 1.3 Natural gas supply networks [3]

each characteristic. Natural gases are produced, purified, compressed, shipped and stored—in compressed or liquefied form distributed and eventually burned or converted into chemicals. Natural gas characteristics that are of particular significance for each operation are usually obtained from the gas well and then processed to meet the process requirements as shown in Fig. 1.3.

Combustion properties

In order to ensure regular and complete combustion of the gas, it is essential that those characteristics which have been grouped together under the heading of "combustion properties" should be both constant and within an acceptable range (which will depend on the combustion equipment used). The more important properties and the flame characteristics which they control are the following:

- Calorific value
- Wobbe index
- Flame speed
- Flammability range

The importance of constant gas composition can be demonstrated by listing certain properties and especially the combustion characteristics of the more usual natural gas components as shown in Table 1.1.

Liquefaction gas characteristics

Another important set of gas characteristics are those which have a bearing on natural gas liquefaction. Since the volume of methane can be reduced by a factor of about 600 by cooling and compressing the gas until it turns into a liquid, there is considerable incentive to ship, store and possibly even distribute it in the liquid

Table 1.1 Composition of natural gas [6]

Fuel composition and properties	Natural gas	Energy crop (switch grass)	Typical wood gas	Agricultural wastes
Methane (CH_4)	93.77	4.50	4.00	4.40
Ethane (C_2H_6)	2.60	0.00	0.00	0.00
Propane (C_3H_8)	0.37	0.00	0.00	0.00
Butane (C_4H_{10})	0.10	0.00	0.00	0.00
Higher hydrocarbons	0.00	1.89	0.00	2.80
Carbon monoxide (CO)	0.00	15.25	18.00	10.20
Carbon dioxide (CO_2)	1.02	17.77	10.00	14.80
Hydrogen (H_2)	0.00	4.18	19.50	11.70
Water vapour	0.00	0.00	4.00	5.70
Nitrogen (N_2)	2.14	56.41	44.50	50.40
Wet fuel molecular weight (kg/kmol)	17.1	29.2	23.7	
Lower heating value (kJ/kg)	46,787	3,913	5,622	5,094
Lower heating value (kJ/Nm^3)	35,679	5,103	5,977	6,053
Stoich air/fuel ratio (volume based)	9.51	1.16	1.31	1.41
Energy content of stoich mixture (kJ/Nm^3)	3,396	2,358	2,583	2,514
Approximate temperature rise of stoich mixture (K)	2,411	1,669	1,882	1,788

form; however, this is not an easy matter; the boiling point of methane is −162 °C and such low temperatures are difficult to achieve clearly.

Liquefaction of ethane, propane and butane is far easier. Importantly, also the critical temperature of methane, i.e. the temperature above which one cannot liquefy it simply by compression of the gas, is low (−83 °C), and methane must be pre-cooled to that level before the ultimate liquefaction pressure is applied.

Pipe Transmission characteristics

A different set of characteristics have a bearing on natural gas transport by pipeline at elevated pressures (50–100 bar). Under these conditions, problems can arise in sections of a transmission system, due to the presence of water and, or, condensable hydrocarbons. The gas supplied from the gas company's fields or terminals is treated to meet a rigid specification for water and hydrocarbon dew point, so that ideally, it cannot deposit any water or hydrocarbon liquids in transmission lines or at pressure-reducing stations. However, appreciating that problems with gas treatment plant at terminals may occur and that the decision to accept gas with water and/or hydrocarbon dew points above the specified limit becomes necessary, systems are normally set up to monitor water and hydrocarbon dew points together with the methanol content of the gas. Responsibility for carrying out these determinations usually rests with company research or scientific sections and liaison with the terminals, producers and regulatory bodies.

It is also necessary to carry out the above-mentioned determinations when new transmission lines are being commissioned. A new line is first tested hydraulically and then swabbed with pigs to remove most of the residual water. To prevent the water still remaining in the line from forming gas hydrates, the pigging procedure is repeated using a large slug of methanol. This removes some more residual water and leaves a film of aqueous methanol which will not form hydrate so readily. From the volume of methanol used, the volume of liquid pigged out and the percentage of methanol in this liquid, estimates can be made of the volume of water still in the main, how long it will take to become dry and whether the methanol content of the aqueous film is high enough to prevent hydrate formation. Decisions can then be taken to either gas the line up or pig again with methanol. After commissioning, the concentration of methanol in the aqueous film must be maintained until the main is dry. This can be achieved by either adding enough methanol to the gas to achieve this or adding more methanol at intervals to the line.

The foregoing paragraphs have shown why instruments for measuring methanol concentration in gas and the water dew point of gas now play an important role in assessments of gas quality. Hydrocarbon dew point measurement is another important criterion of quality, because of the phenomenon of retrograde condensation. For example, at the terminals of the UK gas transmission system, the gas has its hydrocarbon dew point depressed by refrigeration to about −l5 °C at 70 bar, which ensures that no hydrocarbon condensation can occur in any part of the national transmission or distribution system. Any failure to treat in this way results in the hydrocarbon dew

point rising to a temperature in the range of 5–l5 °C at 70 bar and the occurrence of retrograde condensation when the pressure is reduced. Similarly, the presence of water must be strictly controlled in order to ensure satisfactory high-pressure transmission. Most hydrocarbons, higher hydrocarbons more so than methane, form solid hydrates which are stable below a given temperature and above a given pressure. Such hydrates that have the consistency of snow or ice can block valves and small diameter pipes. Their formation can be inhibited by the addition of methanol or glycol; however, the total absence of water, shown by a very low water dew point, is clearly the most effective means of preventing hydrate formation. The next criterion of quality to be considered is odorosity which is extremely important. The main problem is usually to find an odorant or odorant mix which can be applied to all the natural gases encountered in the supply mix to a particular gas company and give satisfactory odorisation control in all its aspects. This is only relevant for gas distributed below 7 bar.

Moving to the quality criteria of interest to utilisation engineers and gas consumers, it is again necessary to point out that these criteria may differ between gas industries due to natural gas composition, regulations and control legislation or may be changed after the introduction of natural gas into a system previously carrying manufactured gas. For example, in the UK, the specified upper limit for hydrogen sulphide in the gas was raised to 3.3 ppm which meant the development of analytical methods and instruments to measure to the new limit for official testing purposes. The measurement of total sulphur concentration also received a lot of attention.

The field of component analysis is also extremely important, and methods of analysis must be agreed on by the producers and the gas company as the analyses have an important bearing on the price structure for the natural gas. Problems may also be experienced in the measurement of CV which is again very important to the producers, the gas company and the consumers. The remaining quality criteria that are of interest are density and relative density (which are important also in metering), Wobbe number, in complete combustion factor (ICF) and soot index (SI).

Compression Gas Characteristics

The need to compress natural gas arises where the latter is transmitted by pipeline, re-injected into underground or above-ground reservoirs for storage under pressure, or where the gas is to be liquefied. Compression on a large scale usually implies the use of fast running centrifugal fans, and the blades of the latter can easily be damaged by solid particles of liquid droplets.

It is, therefore, essential that natural gas that is to be compressed should be free of solid or liquid contaminants, and in this context, its dust, water and higher hydrocarbon content are significant. Especially water, which can be converted into solid ice or hydrocarbon hydrates, must be carefully measured, and both hydrocarbon and water dew point of the incoming gas must be established. Performance of compressors obviously also depends on gas relative density and to a lesser extent on compressibility and viscosity. It is also necessary to design these machines for a given inlet temperature and inlet pressure range.

1.1.3 Natural Gas Specifications

Specification relating to the transmission and distribution of natural gas must have items relating to the integrity of both the pipeline and the flow of gas. The composition of gas entering the pipeline must be such that the pipeline is not damaged. Corrosion, stressing and abrasion must be avoided. The composition must also ensure that, under all temperature and pressure conditions of distribution, the flow of gas can occur. Accumulation of liquid in the pipeline will reduce the capacity of the pipeline and interfere with instrumentation and control equipment. Such interference can lead to erroneous measurements and instrument failure.

The specification has aspects covering:-

i Safe use of domestic appliances
ii Transmission and distribution needs
iii Statutory requirements, e.g. hydrogen sulphide content

There is also a requirement that gas delivered shall be free from materials/dust. Solids or liquids might interfere with the operation of lines, metres or regulators. Such gas shall be free from objectionable odours so that a distinctive "gas odour" can be added to meet the requirements of the Gas Safety (Management) Regulations.

This transmission specification is the primary criterion in considering whether a gas needs processing. If the gas lies outside the specification, an evaluation is made of the possibility of mixing with other supplies, although this can have an effect on the security of supplies.

1.1.3.1 Hydrocarbon Dew Point

A temperature that is low enough to ensure that hydrocarbon liquid does not form under all temperature and pressure conditions of distribution must be specified. The dew point temperature specified will be governed by the lowest ambient temperature that the gas will experience while the gas is at pressures between 24 and 38 bar g (350 and 550 psig). This is the pressure range at which retrograde condensation can occur. This phenomenon must be avoided in the transmission and distribution system. If too high a hydrocarbon dew point is allowed, a gas that is single phase at high pressure can become two phase (gas and liquid) at lower pressure even though the gas is not cooled. The extent of the two phase region within temperature and pressure coordinates depends on the gas composition. In particular, the amounts of heavy hydrocarbons can give rise to retrograde condensation. The amount of the various heavy hydrocarbons allowed in the gas in inversely proportional to the carbon number of the hydrocarbon, i.e. in the series C6, C7, Cs, C9, C10, very much less C10 can give rise to retrograde condensation than that of C6. For most natural gas, after separation of the gas and condensate, the gas phase will contain too much C6+ component. This must be removed by gas processing and a chillers plant is usually employed.

1.1.3.2 Water Dew Point

The presence of liquid water in the transmission system must be avoided otherwise hydrate formation and pipeline corrosion can occur. Hydrates are a physical combination of the lower hydrocarbons and water and once formed are stable. A reduction in the pipeline diameter can result causing the flow capacity of the line to reduce. In the extreme, blockage of the line can occur. Hydrates will also interfere with the correct operation of instrumentation, and it is more likely that blockage of supply lines to instruments would occur. Hydrates can only occur if free water is present, i.e. the gas is at 100 % relative humidity. Corrosion, however, can occur at levels below 100 % RH. Protection against corrosion is ensured by operating at not more than 50 % RH. The RH of gas is often conveniently expressed as a water dew point.

1.1.3.3 Carbon Dioxide Content

In early UK gas purchase contracts, it was considered necessary to specify a maximum carbon dioxide content of 2 %. This was to ensure that acid gas corrosion of the 70 bar (1,000 psig) transmission system was avoided. Later, it became apparent that protection against corrosion was being secured with the water dew point limit. It remains necessary to have gas supplies of low carbon dioxide content due to requirements at the LNG plants.

1.1.3.4 Oxygen

The figure of 0.1 mol% was adopted rather than zero, which in practice is the oxygen content of natural gas, to overcome measurement difficulties and false readings which can occur when endeavouring to determine a zero concentration. This 0.1 mol% limit was sufficiently low to cause air ballasting by gas sellers to be generally uneconomic. However, as the molecular sieve purification units at LNG plants can be damaged by small oxygen concentrations, the current NTS (IOYS) specification has been altered to quote a preferred limit of 10 ppm oxygen; however, it should be noted that the GS(M)R limit is 0.2 mol%.

1.1.3.5 Hydrogen Sulphide

A statutory obligation under the Gas Quality Regulations of the numerous Gas Acts required that gas shall contain a maximum of 3.3 ppm (volume/volume) hydrogen sulphide when distributed to customers. However, 3.3 ppm is seen as the limit and not the level for normal operation. Supplies are usually purchased with a much lower level.



Fig. 1.4 Typical flame [4]

1.1.3.6 Sulphur Content

This category is comprised of mercaptans, organic, sulphides and hydrogen sulphide. The UK (NTS) has adopted a limit of 15 ppm (by volume) to control the amount of corrosion which could occur in domestic appliances following gas combustion. In practice, the level of total sulphur in gas delivered to the NTS is governed by the requirement that the gas is free from objectionable odour. The standard odorisation of a smell-free gas introduces 5 ppm (by volume) of total sulphur.

1.1.3.7 Temperature

Too low a delivery temperature may lead to freezing of the soil around a buried pipeline and damage to other pipelines and services can result. Conversely, a temperature too high can be injurious to pipeline wrappings and coatings. The temperature range of the transmission specification reflects these requirements.

1.1.3.8 Natural Gas Interchangeability

When formulating a gas specification, it is essential to take account of how the gas will burn on domestic appliances and the compatibility of the gas and the range of appliances. The international gas union (IGU) recognised the need to categorise gases according to their properties (See Sect. 1.1.4). Within each group, gases that have Wobbe numbers within +5 % and −5 % of the reference gas Wobbe number should be fully interchangeable, i.e. burn safely, cleanly and efficiently without a need for appliance adjustment. Figure 1.4 indicates the flame profile for a given characteristics.

This 5 % guideline holds good while gases have compositions not markedly different from the Reference gas. However, BG has found that for Group H gas the methane content of the gas must not be <80–85 % for the Wobbe index alone to be

a reliable guide to combustion characteristics. Also, a normal distribution limit of +3 % of reference Wobbe number was adopted. However, this aspect is now covered by other standards such as the GS(M)R.

1.1.3.9 Natural Gas Composition Uncertainties

When a gas company such as British Gas negotiates the purchase of a new gas supply, it is necessary to establish "how much and what it is", i.e. quantity and quality. Aspects that require clarification include the following:-

i Is the composition the same for the entire gas field?
ii How reliable is the given composition? What variations are expected?
iii Are any other natural gas supplies possible? If so, when? What is known about the composition?
iv What processing is the seller intending? In particular hydrocarbon dew point.
v What requirements for gas composition does the seller have?
vi What restriction on gas composition is placed by the seller's transmission and compression requirements?
vii Which pressure does the seller intend for the delivery
viii Can the supply be guaranteed for 365 days/year?
ix Can maintenance schedules of the seller influence the availability of gas?
x What plans does the seller have for supplying gas to other markets? What requirements do these markets place on gas composition?
xi If hydrogen sulphide content is quoted as zero is the statement reliable? How, and for how long, was hydrogen sulphide tested for?
xii What is the total sulphur content of the gas?

Answers to these questions give the company confidence that the chemical and physical properties of the gas will conform to the transmission specification and gas quality statutory obligations.

1.1.4 Classification of Gas Families

Natural gas and liquefied petroleum gases are two examples of fuel gases in large scale use in many countries. Gas industries have been in existence since the early 1800s often based on gases produced (manufactured) from fossil fuels such as coal and later petroleum-based feedstock.

The composition, physical and combustion properties such as relative density and CV of the gases were often very different depending on the feedstock, the process and the process conditions. Similarly, natural gases and liquefied petroleum gases also have a range of Wobbe numbers as a consequence of differences in composition and hence CV and relative density. Wobbe number is an important criterion in the interchange ability of gases and in burner design. The Wobbe

numbers of gases that can be distributed in the UK fall into a very narrow range equal to 49.7 MJm^{-3}(st) (1,335 Btu ft^{-3} (dry)) with a +3/−5 % variation.

The international gas union (IGU) recognised the need to categorise gases according to their properties. Three "Families" of gases were identified:

Family 1: gases having a Wobbe number in the range 22.4–30.0 MJ/m^3(st) (dry). This family embraces the manufactured town gases of various types and liquefied petroleum gases (LPG)/air mixtures within the same Wobbe number range.

Family 2: gases having a Wobbe number in the range 39–55 MJ/m^3(st) (dry). This family was intended to include all the various natural gases, including synthetic gases and some LPG/air mixtures.

Family 3: gases having a Wobbe number in the range 73–88 MJ/m^3(st) (dry). This family consists of the LPG's, i.e. essentially propane and butane. The variety of gases within each family is too great; however, for all the gases to be fully interchangeable without appliance adjustment, it has been necessary to subdivide each family into "Groups". For natural gases, the "Second Family" was divided into two "Groups", Group L (based on the Dutch natural gas) consisting of gases in the lower part of the Wobbe number range (39–45) and Group H consisting of gases of higher Wobbe number (45.7–55 British natural gas). For each group, a "reference" or "adjustment" gas was defined, together with limit gases, for use in the testing of appliances (see interchangeability).

1.2 Combustion Properties

1.2.1 Calorific Value

The most important property a fuel gas possesses is the energy liberated when it is burned. This may be expressed as the heats of formation of its combustion products on a molar basis. In the fuel industries, however, this property is much more commonly expressed as the CV which is the quantity of heat released by complete combustion under isothermal conditions at a constant pressure of one atmosphere, and at a specified reference temperature, of unit quantity of the fuel, the water formed during the combustion being in the liquid state, any sulphur in the fuel being converted to sulphur dioxide and any nitrogen remaining as such.

Complete combustion is possible with gaseous fuels at atmospheric pressure, and the CV is measured at constant pressure in a calorimeter. Solid and liquid fuels on the other hand require higher pressure, and determinations of CV are made under constant volume conditions in a bomb calorimeter.

Fuel gases that contain hydrogen or hydrocarbons possess two CV's, the superior (gross) CV and the inferior (net) CV depending upon whether the water formed in combustion is in the liquid or vapour phase.

The superior (gross) CV of the gas (relative to the volume of the dry gas) is defined as the amount of heat given out by the complete combustion of the gas with air, at a constant pressure of 1.01325 bar and at a constant temperature t_H of a

specific volume (V) under specified conditions (t_v, P_v), all the water that is formed during the combustion being condensed at the temperature t_H(6). The superior CV is designated as:

$$H_S\{t_H, V(t_v, p_v)\}. \tag{1.1}$$

The inferior (net) CV of the gas (relative to the volume of dry gas) is defined as the amount of heat given out by the complete combustion of the gas with air, at a constant pressure of 1.01325 bar and at a constant temperature t_H of a specific volume (V) under specified conditions (t_v, P_v), all the water that is formed during the combustion remaining in the gaseous phase at the temperature t_H. The inferior CV is designated as:

$$H_i\{t_H, V(t_v, p_v)\} \tag{1.2}$$

The gross CV provides the basis on which charges are made by the gas industry on their consumers, and the British Gas Industry for example is required by law to declare and maintain (within specific tolerances) this value. The specified conditions referred to above on the international system of units are known as "standard reference conditions" (src) or "metric standard conditions" (MSC) viz. a temperature of 15 °C and pressure of 101,325 Pa (dry). On this system, the (gross) CV is expressed in MJ m^{-3}(st). These are the prime units used by British Gas to "declare" the CV; however, they are now converted for billing purposes. For imperial units, the conditions for gas volume measurement are 60 °F, a pressure of 30 in Hg (at 0 °C and subject to standard gravity 1015.92 mb).Using these units, the conversion from S.I. to imperial is given by 1 MJ/m^3(st) (dry). The adoption of the "dry" cubic foot is a departure from previous gas industry practice resulting from the fact that natural gas, as transmitted at high pressure, is essentially dry.

1.2.2 Wobbe Number

This is termed the Wobbe index or Wobbe number, and it gives a measure of the relative heat input to a burner at a fixed gas pressure of any fuel gas. Consider the flow of a gas through an orifice (which may be the injector orifice of an aerated burner or the burner port of a postaerated burner).

$$W = \frac{\mathrm{CV}}{\sqrt{\rho}} \tag{1.3}$$

Although details of burner design and sizing can be found in Sect. 6.1, when a gas conforms to the specified Wobbe number and CV ranges and the impurity levels, it is suitable for supply to customers. A typical aerated gas burner is shown schematically in Fig. 1.5 indicating the various components of an aerated burner.

This is basically the same design as a BW1Sen burner, but there is no collar to adjust the air supply. Gas enters the burner throat via a jet.

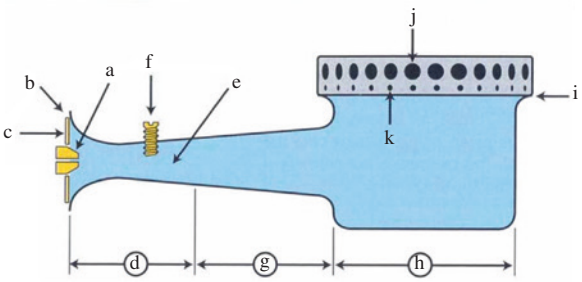


Fig. 1.5 Section through an aerated burner [5]. **a** Injector **b** Primary airport **c** Aeration control shutter (primary air adjustment) **d** Venturi **e** Venturi throat **f** Throat restrictor (alternative primary air adjustment) **g** Mixing tube (burner tube) **h** Burner head (burner body) **i** Alternative flame retention ring **j** Burner ports (flame ports) **k** Retention ports (retention flame ports)

The gas stream entrains air into the burner throat to give a gas/air mixture. Entrainment is by momentum sharing. The momentum of the gas jet is a function of only injector size and gas supply pressure and is independent of gas properties. Therefore, the burner will entrain a fixed amount of air. Typical burners entrain between 40 and 80 % of the air required for complete combustion (referred to as the percentage primary aeration). The gas/air mixture passes from a divergent tube which slows the mixture (the kinetic energy being converted to potential energy in the form of pressure). The gas/air mixture issues at the flame ports at the burner head. Combustion takes place firstly as a premixed (inner cone) and then as a diffusion flame (outer cone). Air diffuses into the partially burnt fuel gas to complete this second-stage combustion.

Heat input is directly proportional to gas CV. The flow of gas into a burner is inversely proportional to the square root of the relative density (RD). Hence, the heat input is dictated by the CV of the gas and the flow characteristic. Extensive research from that time has shown the Wobbe number to be a very good measure of burning character.

What are the combustion consequences if the Wobbe number changes? Consider a gas supply which has high methane content and gives satisfactory combustion, i.e. keen flames and a low carbon monoxide output. Then increase the ethane and propane content in the gas and hence increase the Wobbe number. The primary air entrained remains the same, but the air requirement for complete combustion has now increased, i.e. the percentage primary aeration achieved has fallen. This results in more air being required through diffusion at the flame to complete the combustion. The flame therefore lengthens and is more likely to impinge on cold heat exchanger surfaces. Laboratory work has shown a direct relationship between Wobbe number and CO production.

Raising the Wobbe number by 1.5 MJ/m^3 (40 Btu/ft^3) will double the CO output. Too great a reduction in Wobbe number will lead firstly to a loss of heat service and further reduction will lead to flame stability problems, i.e. the flame can lift away from the burner.

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Chapter 2
Transmission and Distribution Systems and Design

2.1 Transmission Pipelines

2.1.1 Introduction

Once natural gas is first extracted from corresponding well, this is followed by gas treatment plant (downstream process), then via transmission pipeline to distribution network systems, and to the consumers. Through this transmission gas delivery, the concept of pipeline design is established through its location, the type of fluid being carried and its operating pressure and temperature are also of prime importance within this process. Unusual examples of severe location are the trans-Alaska pipeline where long sections are laid above ground to protect the permafrost and the Middle East, where long pipelines are laid both above and below ground and can be subjected to large temperature changes that bring complex design concepts. The operating pressure will determine the grade and wall thickness of the material, and temperature will also affect the requirements for pipe coatings, insulation, expansion joints, anchor blocks, etc. Gas pipelines require special attention over and above the normal design requirements for liquids because of the vast quantity of stored energy in the pipeline steel due to the compressibility of gas. The design should therefore take account of the probable consequences of failure and the parameters that can be adjusted to minimise the possibility of failure in pipeline systems. This chapter therefore attempts to provide the wide guidelines related to issues with the design and integrity of the transmission and distribution pipelines together with safety aspects of the systems.

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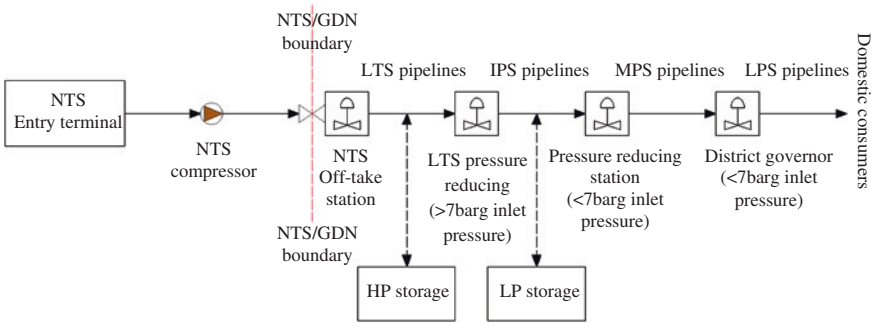


Fig. 2.1 Transmission and distribution system network [21]

2.1.2 Gas Transmission Pipeline Design

The design of gas transmission and distribution systems involves a series of tasks with their interdependencies to one another as shown in Fig. 2.1 below.

2.1.2.1 Design Criteria

Natural gas transmission systems design philosophy has survived revisions of guidelines that specify the detailed criteria for all components design. While standards used in different countries and regions differ, still there is wide overlap of the fundamental guides between them. The major criteria considered below include pressure and temperature ratings, gas specifications, gas velocity, pipeline sizing, stress analysis and location class.

Pressure and Temperature Ratings

Typically, new onshore and offshore gas transmission pipelines are designed and constructed to operate in the pressure range 40–94 barg, although IGEM/TD 1—Edition 5 allows for a maximum operating pressure of 100 barg. The maximum operating pressure of existing onshore pipelines is limited based on a periodic risk assessment of proximity to buildings, crossings and pipe condition. Offshore subsea pipelines are operated at much higher pressures—typically 150 barg or higher—due to the lower consequential impacts of pipeline failure. Normal operating temperatures are relatively constant for buried pipelines, except in extreme conditions, and are generally taken to be 5 °C. IGEM/TD/1 stipulates a design temperature of 0 °C, with some exceptions such as at the exit from

a pressure-reducing installation lower temperatures, at the exit from a compressor station—higher temperatures, at exposed bridge or other overhead pipeline crossings—lower or higher temperatures [1–3]. All pipeline materials should have adequate "fracture toughness" at or below the minimum design temperature. The pipeline must be designed for all possible operating temperatures; extreme examples are the above ground expansion/contraction loops on the trans-Alaska pipeline and similar loops incorporated into above ground desert pipelines in the Middle East. Exceptions to the normally assumed 5 °C operating temperature in the United Kingdom are:

- Low temperatures on the outlet of pressure-reducing regulators
- Above ground pipework affected by ambient temperatures
- High temperatures on the outlet of compressor stations, which can persist for up to 50 km.

Gas Specifications

All pipeline materials should have adequate "fracture toughness" at or below the minimum design temperature. Although Fig. 2.2 has indicated the water dew point, in most cases, the gas that is conveyed in onshore gas transmission pipelines is dry, "sweet" methane, and the key gas quality parameters that are monitored and controlled are:

- Gross Calorific Value
- Relative Density
- Water Dew point
- Hydrocarbon Dew point
- Total Inerts
- Oxygen Content
- Hydrogen Sulphide
- Total Sulphur Content
- Impurities—dust, oil and contaminants

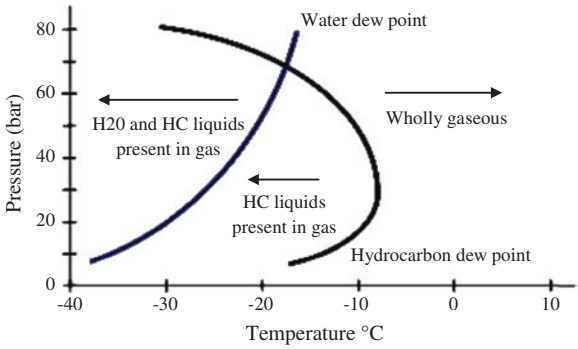


Fig. 2.2 Hydrocarbon dew pointing [21]

Gas Velocity
Theoretically, there is no velocity limit but dust, which is always present in pipeline gas, produces an abrasive effect when carried in the gas stream. A maximum velocity of about 20 m/s is therefore recommended to avoid erosion of the pipe. Particularly at bends, very high flow rates can induce "dust storms" ,and these are often induced during the commissioning of pipelines or equipment resulting in blockages and malfunctions of A.G.I. equipment. With appropriate control of gas quality and suitable gas-filtering arrangements at pipeline entry point, there is generally no need to limit the gas velocity. However, where dust is a particular problem, internal abrasion of the pipe can take place and velocities should be limited accordingly, typically up to 20 m/s. The fundamental equation involves [16] expressed as:

$$V_e = N\frac{c}{\sqrt{\rho}} \tag{2.1}$$

where V_e is the erosional velocity (m/s), N is 1.22 for metric system, c is a constant ranging between 100 and 250 [16], and d is the gas density (kg/m^3).

Pipeline Sizing
The required diameter will depend on economics as well as on the minimum pressures and required flow capacity. The requirement for compressors, which is of strategic and economic importance, will also affect the pipeline size. However, for basic sizing purposes, the steady state general flow equation can be used as described in Sect. 4.3 on Gas Flow in Pipelines. Consideration should always be given to transient analysis (unsteady state) to test the capability of the pipeline to meet emergency conditions and in the case of liquid pipelines to ensure that the pipeline is designed to meet surge conditions.

Pipeline Stress Analysis
Maximum Allowable Stress—Pressure and temperature as well as other operating conditions such as bending can create expansion and flexibility problems, and therefore, stress criteria are specified in all codes, limiting the level of combined stresses allowed in a pipeline. The design factor relates only to hoop stress if other stresses are significant, then these could contribute to the pipeline steel exceeding its yield stress. Design codes vary in the way they calculate the combined or equivalent stress but the following equation is typical:

$$\sigma_{\text{eq}} = \sqrt{\sigma_c^2 + \sigma_1^2 - \sigma_c\sigma_1 + 3\tau^2} \tag{2.2}$$

where, σ_{eq} = von Mises (equivalent) stress, σ_{c} = circumferential stress, σ_{l} = longitudinal stress, τ = shear stress in plane of pipe cross.

Pipeline Fatigue

The fatigue life of a pipeline is usually defined in terms of the allowable pressure (stress) ranges and the associated numbers of pressure cycles. Normally, a 40-year life is assumed, but this could be longer or shorter depending on circumstances. Pressure cycling (i.e hoop stress) can cause small weld defects to grow in time to a critical size and is therefore the major factor in determining the fatigue life of welded steel gas pipelines, particularly those pipelines that are designed to utilise line-pack storage. Fatigue life is not greatly influenced by temperature, providing that the fracture toughness properties are met, because temperature effects are small in comparison with those produced by hoop stress. When a new pipeline is high-level hydrostatically tested, any existing defects will grow under the influence of the high stress level and any that reach a critical length will fail resulting in pipe rupture. Normally, pipelines do not fail and any remaining defects are therefore non-critical. Pressure cycling, however, causes a gradual growth in the remaining defects such that one or more could become critical in time. Restrictions on pressure cycling are required to prevent this, particularly on pipelines operating under line-pack conditions. British Gas has determined that a pipeline life of 40 years, assuming 1 cycle/day, is equivalent to 15,000 cycles of 125 N/mm^2 magnitude. There are two common approaches:

- **Constant daily pressure cycling** where the magnitude of daily pressure cycling is constant, the fatigue life should be determined from: S3 N $=$ 2.93 $\times$ 1,010 where S $=$ constant amplitude stress range (N/mm^2), N $=$ number of cycles.
- **Pressure cycling** where the magnitude of daily pressure cycling is not constant, the fatigue life is normally evaluated on the basis of converting the recorded variable cycles to an equivalent spectrum of constant amplitude stress cycles and the following condition for the damage fraction should be satisfied to obtain an acceptable fatigue life: DF $=$ Σ ni $\leq$ 1.0 Ni

where ni $=$ the actual number of cycles accumulated at stress range Si, DF $=$ damage fraction, Si $=$ stress range (Ni and ni), Ni $=$ number of stress cycles allowed at stress range Si. 15,000 cycles at 125 N/mm^2 has therefore been set as the maximum permissible fatigue life in IGE TD/1. According to Fig. 2.3, if any cycles are <125 N/mm^2, they are multiplied by a factor based on the number of stress cycles at the lower level, as shown in Fig. 2.3, which will be required to cause the same damage as 15,000 cycles at 125 N/mm^2.

Location class

For gaseous fuel such as natural gas in pipelines and other toxic and flammable gas pipelines, the design codes use a system of area classification in order to have an objective view for allocation of design factors and other elements such as minimum proximity. The rationale behind this type of classification is that concentration of people represents an increase in both risk to that population and also an

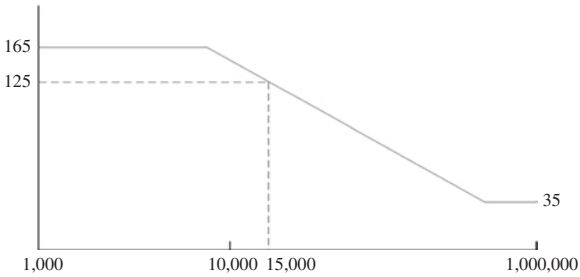


Fig. 2.3 Stress range versus number of cycles [21]

increased chance of third-party interference, i.e. digging up the pipeline, caused by that population. The manner in which the design codes perform this analysis is to establish a corridor centred on the pipeline. For a given length of corridor, the designer counts either the number of "people normally present" or "dwelling units" dependant on the code. The average number of people or dwellings is then calculated for the particular strip and the resultant design factor applied. The codes, however, make the highest classification subjective,—"central areas of towns and cities"—resulting in some confusion from time to time as to exactly what this constitutes.

It should be noted that only ASME B31.8 dictates a regarding of the pipeline on a continuous basis to take account of new developments built near to gas pipelines. Replacement of sections of the pipeline or restrictions in the MAOP can be necessary if a large development is built close to an existing pipeline. It is unlikely in the United Kingdom to occur due to the involvement of the HSE in planning application close to existing high-pressure gas pipelines. Other countries also have a form of building control within certain distances from hazardous installations or pipelines. Design codes BS 8010 and TD/1 also incorporate minimum distance requirements variable on MOP, pipe size, design factor and wall thickness, depicted on a graph.

The different ways of calculating area classification make direct comparison a little complicated; however, if an average of 3 persons/building is used (which is a common assumption used when counting people in the absence of detailed information), then a comparison can be made in the table below. There is a common trend. The general design factor for each area is also given in Table 2.1, to

Table 2.1 Design factor for various standards [16]

ASME B 31.8				ISO 13623		PD8010/IGE TD/1	
	No. of buildings in area	No. of people/km (approx) l km^2	DF	No. of people/km	DF	No. of people/km	DF
1.1	<10	<46	0.8	0–1	0.83	1/R– <250	0.72
1.2	<10	<46	0.72	2– <50	0.77	1/R– <250	0.72
2	10–46	46–215	0.6	3–50 to 250	0.67	1/R– <250	0.72
3	>46	>215	0.5	4– >250	0.55	2/S– >250	0.3
4	Subjective	Subjective	0.4	5–Subjective	0.45	3/T–Subj	<16 bar

show the effect on design factor. There are often exceptions to particular features, e.g. schools and hospitals, which commonly down grade the area by one or more groups.

There are also specific design factors allocated to road, rail and river crossings and other such items. These are too numerous to list here, but are generally one classification lower than the adjacent pipeline.

Material grade

Material type considered as line pipe material specification until relatively recently was limited to American Petroleum Institute (API 5L). This specification provided the basic building blocks but was normally supplemented by additional company standards or specific project ones. In conjunction with the EN and ISO pipeline design codes, ISO line pipe specifications, ISO 3183, have also been issued, which have been incorporated into the EN system as EN 0208. The ISO and EN codes require use of their own line pipe specifications and requirements when the code is used (as well as a number of other "normative" codes listed in each code).

The ISO/EN line pipe specifications are split into three classes, A, B and C. A is essentially a rewrite of API 5L, B adds some general amendments varying such items as the under tolerance and offering numerous additional requirements subject to choice by the designers, and C is for sour service and special pipes. Class B is the one most anticipated to be used for pipelines. The specifications also round up the SMYS in terms of kN/m^2 varying the standard API SMYS ratings by a few per cent. The standard under tolerance for wall thickness for class 8 line pipe is reduced to 5 % from 10–12.5 % on the basic API 5L specification. Pipeline specifications list "standard" thicknesses of pipe that are usually used in order to gross up to the next available thickness. However, for pipelines longer than 50–60 km, there is a growing tendency to obtain pipe of exactly the required thickness at minimal additional cost. This is especially true for pipe of increased thickness where the standard sizes can be several mm apart. For long trunklines, several pipelines are now installed with the 0 D non-standard, with a constant ID, regardless of wall thickness. There are considerable savings to be made by this approach, but it needs to be looked at for each design and is not applicable for all projects, usually those where a fixed throughput is specified and where the distance is long (>100 km). Other non-standard items that have occurred include purchasing line pipe in 18 m lengths. The cost of non-standard trucks to transport them can outweigh the reduction in welds of 50 %. Logistics play an important part in this option as these loads are significantly longer than standard trailers. Line pipe grade is another area where changes in the "standard" have occurred in recent years. The most common grade of line pipe seen is now X60 or X65, certainly for larger diameter and higher pressure pipelines. Pipelines using X70, X80 and X100 are now being assessed or built. They do require greater care during welding, and there are issues related to the narrowing between SMYS and UTS. A minimum ratio of 90 % of UTS to SMYS is commonly applied.

Wall thickness

Wall Thickness (*t*)—Pipeline steels have standard values of nominal wall thickness with tolerances to allow for the manufacturing process. Acceptable tolerances vary depending on the requirements of the pipeline owner, but a typical tolerance for under thickness is 5 % for submerged arc welded (SAW) pipe. The nominal wall thickness therefore is not necessarily the actual wall thickness. In ANSI B31.8 pipelines can be designed using the nominal wall thickness but BS 8010 and IGE/TD/1 require the pipelines to be designed on the basis of "minimum wall thickness", i.e. the nominal wall thickness less the maximum tolerance for under thickness. The minimum wall thickness should be equal to or greater than the design thickness *t* as in Eq. 2.3 where:

$$t = \frac{P \cdot d}{2fs} \tag{2.3}$$

t = design thickness, d pipe diameter, p = internal pressure, S = specified minimum yield stress of the pipe material, f = design factor <1.0.

From the Eq. 2.1, it is clear that the design thickness is dependent on other parameters that will be determined by the required operating conditions. The pressure and diameter will be determined by the required transmission capacity, and the design factor will be dependent on the pipeline route; therefore, it remains to select a material grade that will give a suitable design thickness or vice versa.

An important consideration in the selection of pipe wall thickness, particularly for gas pipelines, is to know what wall thickness is required to resist penetration by mechanical equipment and what depth of defect will result from the different types of machinery likely to be encountered.

Figure 2.4 shows the results of tests carried out by British Gas to find the force required to produce a 0.5 lmm (0.2 inches) dent in X60 pipeline steel of different wall thicknesses. It will be observed that a wall thickness of 12.7 mm (0.5 inches)

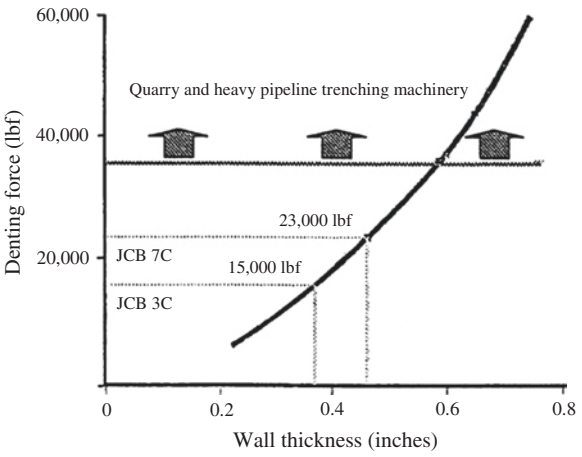


Fig. 2.4 Denting force versus wall thickness [21]

will resist impact by the lighter types of machinery encountered in quarrying and pipeline trenching; however, during normal operations, it is unlikely that such equipment will be in use without the pipeline operator being aware of it. A wall thickness of 12.7 mm (0.5 inches) will resist impact by any of the commercially available excavators likely to be used during normal construction activities and is therefore suitable for use in high-risk areas, and a 9.52 mm (0.375 inch) wall thickness will resist an impact of 15,000 lb_f that makes it suitable for use in the more remote areas. The choice of material grade should therefore reflect the requirement for a wall thickness appropriate to the risk category of the pipeline route. Additionally, the requirement for thicker walled pipe should not be taken as an opportunity to reduce the material grade in any area below that being used on the remainder of the pipeline. High-risk areas require a combination of impact resistance and low operating stress level, which is provided by using a combination of appropriate wall thickness and grade of steel.

Material Properties of the pipe must have sufficient fracture toughness properties to prevent propagating brittle or ductile fractures at the minimum operating temperature of the pipeline. In the United Kingdom, the normal operating temperature is taken as 5 °C and it is required by IGE/TDI that fracture toughness properties are demonstrated at 0 °C to give a margin of safety. In the case of brittle fracture, the requirement is for a minimum of 75 % shear area when subject to a Drop Weight Tear Test (D.W.T.T), i.e. <25 % brittle fracture. However, if the pipeline is to operate permanently at a level below 30 % SMYS, then a D.W.T.T. is not required because propagating fractures do not occur below this level.

For existing pipelines operating above 30 % specified minimum yield stress (SMYS), the maximum design factor should be that corresponding to the Drop Weight Tear Test (D.W.T.T) transition temperature. As previously described, the arrest/propagate boundary is a function of the level of hoop stress and the transition temperature. Samples of the pipeline steel are therefore required for D.W.T. testing. Ductile fracture is avoided by ensuring that the pipeline steel has sufficient energy absorption properties to prevent fracture occurring. The Charpy V-notch impact test is used as previously described, and the energy level required will depend on the grade of steel and diameter of pipe (Fig. 2.5 and Table 2.2).

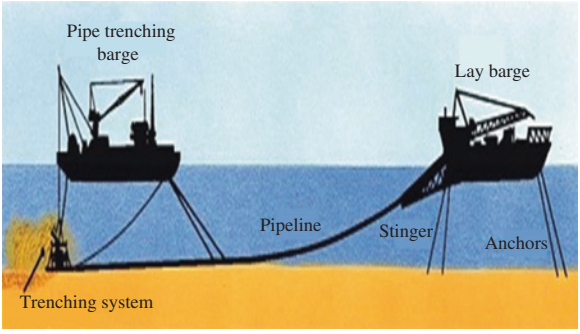


Fig. 2.5 Pipeline trenching system [21]

Table 2.2 Tabulated steel grades [19]

Steel grade (I.S. EN 10208–2)	Minimum yield strength (N/mm^2)	Corresponding steel grade (API 5L)
L 245	245	B
L 290	290	X42
L 360	360	X52
L 415	415	X60
L 450	450	X65
L 485	485	X70
L 555	555	X80

For existing pipelines operating above 30 % SMYS, the maximum design factor should be that corresponding to the D.W.T.T. transition temperature. As previously described, the arrest/propagate boundary is a function of the level of hoop stress and the transition temperature. Samples of the pipeline steel are therefore required for D.W.T. testing before Fig. 2.6 can be used.

High-Density Routes—the pipeline must use pipe with a minimum wall thickness of 11.91 mrn, and the design factor must be <0.3. The thick-wall pipe should extend to a distance equal to the proximity distance on either side of the crossing, i.e. it is treated as though it were a type *S* area.

Other Routes—the pipeline must use a minimum of 9.52 mm wall thickness and have design factor of <0.3. In all locations where a pipe wall thickness of 19.1 mm or more is used, the design factor can be raised to 50 % SMYS because risk assessment has shown the risk of failure to negligible.

Parallel Routes—Pipeline routes parallel to traffic routes of any sort should be avoided. Any pipeline that is parallel and within the proximity distance (from the

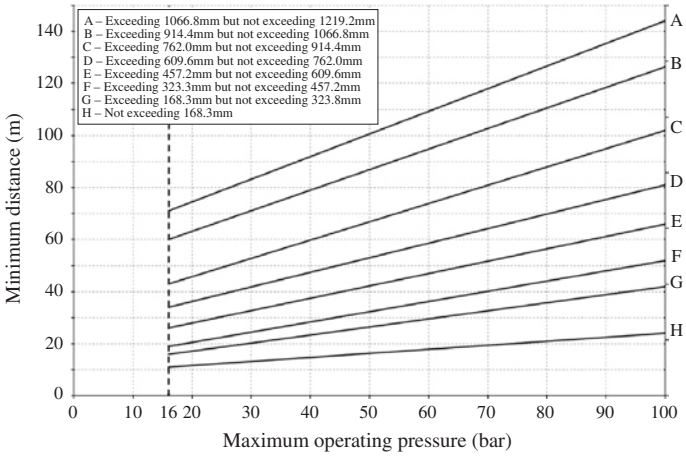


Fig. 2.6 Operating pressure versus minimum distance chart [21]

Table 2.3 Tabulated nominal thickness [19]

Outside diameter (mm)		Least nominal wall thickness (mm)
Exceeding	Not exceeding	
	168.3	4.78
168.3	457.2	6.35
457.2	608.6	7.92
609.6	914.4	9.52
914.4	1066.8	11.91
1066.8	1219.2	12.70

carriageway edge or rail running track) must conform to the traffic route requirements for the whole length of the infringement. Under-exemplified UK codes sleeves are avoided wherever possible, however, ANSI B31.8 still recommends sleeves at crossings as per Table 2.3.

The introduction of "thick-wall" pipe has considerably reduced the need for sleeves, and they should be avoided wherever possible, but may be needed for construction reasons in certain locations. In such cases, the pipeline design requirements should still conform to the appropriate crossing classification. Some existing pipelines were constructed using sleeves and therefore continue to operate under the original design conditions. Sleeves are classed according to their operational use:

Class 1—Sleeves required for (a) protection against external interference and (b) protect the public and property against a failure of the carrier pipe.
This applies where the carrier pipe is operating above 30 % SMYS or has a wall thickness <11.91 mm, i.e. a Class 1 sleeve must contain the line pressure in the event of a carrier pipe failure.

Class 2—Sleeves required for impact protection only, i.e. pipelines operating below 30 % SMYS but where the pipe wall is not thick enough to prevent serious impact damage

Class 3—Sleeves installed for construction purposes only. This class of sleeve can be (a) concrete (b) steel. If a construction sleeve is required, then concrete is better since concrete does not create a shield against CP currents.

2.1.3 Natural Gas Compression

Compressors are used to increase the flow capacity in the system between two points. They increase pressure to help overcome flow-related pressure losses but they also increase flow capacity by making the gas denser as shown in the cross section of Fig. 2.7; denser gas occupies less space and it moves with lower velocity so it loses less pressure along the way. Issues with legacy compressors are as follows: requirements of the industrial emission directive for emission limit values on No_x and CO, best available technique (BAT), low-efficiency operation, NTS reconfiguration requirements for changing flow patterns. Electric variable speed

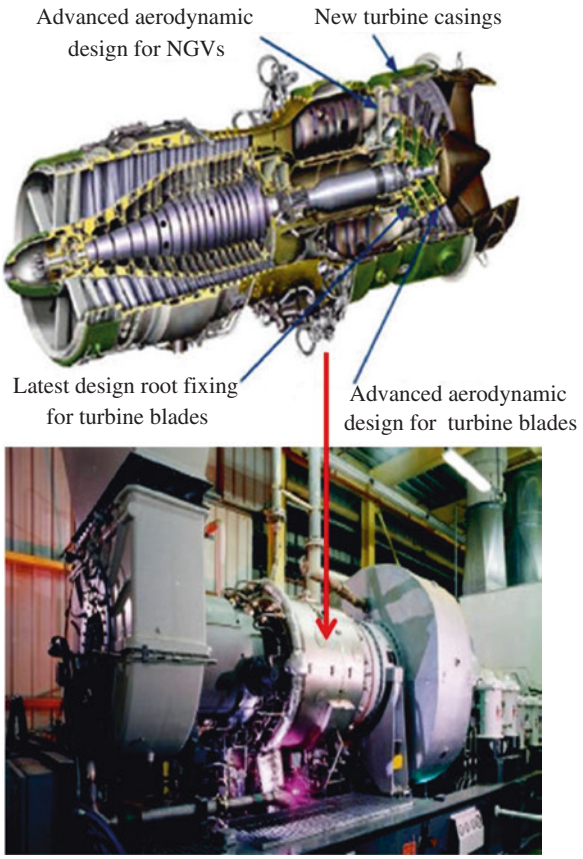


Fig. 2.7 Compressor units [22]

motor drives and compressors are gradually replacing gas turbines and gas engines because of the following reasons: (i) There are no "local" air quality emissions, (ii) the electric motor has a high efficiency at its rated power, i.e. the full speed range of the compressor can be used without being compromised by adverse compressor drive issues, (iii) the GB transmission system operator has installed electric variable speed drive units on about six sites nationally.

2.1.4 Testing and Commissioning

Hydrostatic testing is carried out for all transmission pipelines operating at a design factor $\geq$0.3, and the required test pressure is calculated from the expression in Eq. 2.3 and can be expressed as [3–6]:

$$P_t = 20 t_n f_s D^{-1} \tag{2.4}$$

where P_t = test pressure (bar), t_n = nominal wall thickness (mm), s = specified minimum yield strength (N/mm^2), f = design factor, D = outside diameter of the pipe (mm)

In order to provide some test that the final completed pipeline with its various components and variations possible in materials, jointing, construction and back-filling are fit-for-purpose, a pressure test with or using a harmless fluid prior to commissioning is required in most design codes and safety legislation as being necessary. The accepted means is for a hydrostatic pressure test, whereby the pipeline is filled . completely with water, pressurised up to a set pressure and the pressure held for a minimum period of time, commonly 24 h, but sometimes Jess. Some of the design codes allow pneumatic testing, but this is generally limited to low pressure and low design factors due to the vastly increased energy in a pneumatic test versus hydrostatic (>300 times) and the fact that any small leak takes much longer to be apparent on a pneumatic test versus hydrostatic.

Pressure testing is designed to achieve the following considerations: (i) demonstrate the integrity of the pipeline, (ii) locate the presence of small leaks and pinholes, (iii) remove defects, (iv) work-hardening small defects increasing fatigue life. The setting of the pressure test level needs to follow the methodology of the design code, but care also needs to be taken not to overstress the pipeline either at its lowest point or, when design calculations have used the empty weight of the pipe, as in gas pipes, and not the temporary weight of the hydrotest water. There have been occasions, often not publicised, where supports, pipe bridges and other spans have failed when a gas pipe is hydrostatically tested because insufficient thought had been given to the weight of the pipe when full of water instead of gas. The case for high-level or yield testing of pipelines relates to fracture mechanics, fatigue analysis and crack growth. The testing procedure normally recognises that if the pressure *I* volume graph for the water being used as the pressurising medium starts to get to a slope half of that found during the elastic portion of the test or that the pressure increase per unit volume is half that previously encountered, pressurisation should cease, regardless of the pressure as the pipeline has started to yield.

Acceptance criteria for pressure tests are normally blithely stated as “no pressure drop that cannot be accounted for by variations of temperature or other factor”. It is very rare for a pressure test to achieve the test pressure and to remain unchanged within the limits of accuracy of the test instruments. The pipeline temperature is critical to understanding and computing the variations in pressure that will occur. Temperature . probes should be buried alongside the pipeline in at least three locations along the pipeline route. BS 8010 Sect. 8.6 and associated graphs provide a practical guide to computing the variations due to temperature. In the end, a decision is made based on observations, calculations and acceptance by one of the parties, usually the client or the testing company that the pipeline is free from defects. The pipeline is filled with water using pipeline “pigs” and pressurised to the required test pressure as a test of mechanical strength and integrity. The hydrostatic test also “work-hardens” and stress relieves the pipeline. A pressure–volume (PV) plot in Fig. 2.8, of the water added during the pressurisation process, is constructed, and this allows for the calculation of air content, which must be <0.2 % of the fill volume under test (or 0.5 % for short pipeline lengths).

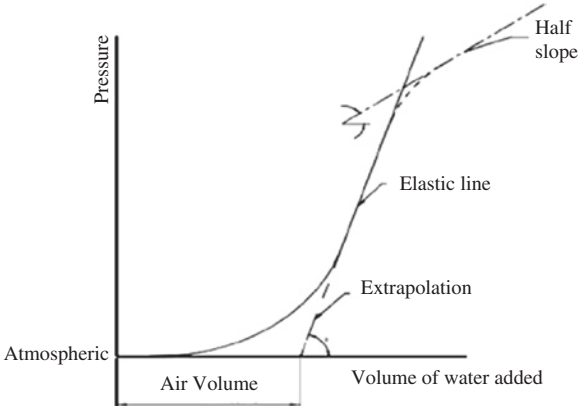


Fig. 2.8 Typical PV representations

Commissioning of pipelines is carried out in the following order of preference:

(i) Super-dry air/nitrogen followed by gas—an air-drying unit and compressors are used to propel foam pigs through the pipeline, which absorb liquid water and distribute residual water as a thin film on the internal pipe wall to facilitate faster evaporation.
(ii) Vacuum drying followed by purging and gassing up—by reducing the pressure of air to 10 mbar, water will evaporate at 70 °C (the "saturation vapour temperature") and the vapour is extracted via the vacuum. Methanol swabbing using a commissioning pig train is also conducted to ensure drying the pipe for water condensates, and it follows a sequence as outlined in Fig. 2.9 below.

In general, the key threats/risks to a gas transmission pipeline during its operating life arise from the following considerations (i) third-party interference, (ii) corrosion, (iii) ground movement, (iv) flooding, (v) internal/external stress corrosion cracking, (vi) fatigue and (vii) human error. A suitable, risk-based approach to Operations and Maintenance (O&M) is required to ensure that the pipeline has adequate integrity and remains fit-for-purpose. This involves assessing pipeline threats and risks and mitigating these risks through appropriate O&M and training [7].

(i) **Third-party interference**: Typical O&M measures that are used to mitigate these risks include, pipeline surveillance in the form of dial-before-you-dig programmes, foot surveys, aerial surveys and landowner liaison, increased depth of cover

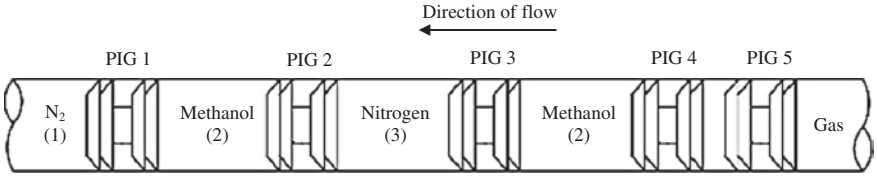


Fig. 2.9 Methanol swabbing process



Fig. 2.10 Typical third-party interferences [21]

This has historically been not only the most common loss of integrity as exemplified in Fig. 2.10, but also provides the largest loss of integrity and the highest level of damage. The root cause of many incidents is lack of knowledge of a pipelines existence or depth. The most obvious place to start with in terms of information is visual markers on the ground. However, operating companies often find themselves in a dilemma as to whether increased knowledge of the location of the pipelines makes them vulnerable to deliberate attack and whether this out weighs the decrease in third-party activity. Marker posts also require constant maintenance to keep up to date telephone numbers, etc., and to repair damage. Marker tape below ground has been shown to be of little effect unless combined with some other form of protection. Regular contact and notification with the landowner and occupier usually reap large dividends in preventing damage. Many pipelines have a right to walk *I* inspect the pipeline annually, which affords time to contact each landowner. Loss of this right through inactivity and not keeping on up to date list of owners and occupiers has often been shown to be a false economy for the pipeline operator. A further means of preventing damage that has received much attention in the USA is the use of a "one-call" system whereby a contractor can phone one (free) number giving the location of where he is planning to excavate and a central record provides details on any pipeline or cable buried in the vicinity. This normally requires legislation to require all operators to provide the information and contribute towards the running costs. The incorporation in the United Kingdom of the New Roads and Street Works Act may perform a similar function, but the register of services is still limited and may take some time to build up to a comprehensive record. The majority of pipeline companies utilise an aerial observation of their pipelines, which can provide warning of works being undertaken on or close to the pipeline. The normal frequency adopted is bi-weekly, although this can have the effect of activities being kept on hold until the helicopter or plane has been over with the knowledge that there is two weeks grace before another inspection.

As in most activities, the hardest information to gather is near misses or in the case of pipelines dents or gouges. There is a human tendency not to report damage

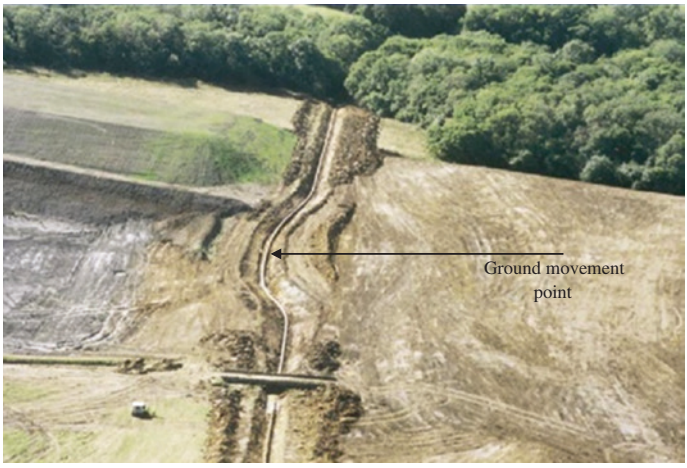


Fig. 2.11 Ground movement [21]

to someone else's property if it is not apparent and forget about it. Pipelines are vulnerable to this type of damage, especially gouges that concentrate stress levels and accelerate corrosion at the locality. The only practical way to get information about these types of incidents is for the pipeline operator to accept that repair costs will not fall onto the person inflicting the damage. In the long run, this will produce effective results, but is sometimes difficult to accept.

(ii) **Ground movement**

The concept of ground movement as shown in Fig. 2.11 due to movement of ground occurring as a result of sand displacement due to ground movement

(iii) **Corrosion**

There are very few pipelines that are buried without some form of corrosion protection incorporated on them in the form of coating(s) and cathodic protection as shown in Fig. 2.12. The various types of coatings and their advantages are covered in depth elsewhere, but two main points can be highlighted.

The first is that the long-term success of a coating has been shown many times to be related to the surface preparation. There is no real substitute for grit blasting to an acceptable standard. The second is that a cathodic protection system needs regular inspection and monitoring to ensure its continued success. Too low a voltage does not provide protection and too high a voltage can damage coatings quite severely [8]. Internal corrosion can be prevented by the use of either the correct material to resist attack, internal coating or additives to prevent corrosion. Internal coatings have not had a good track record in the past, but advances in application technologies now mean that they should be at least as good as external coatings now are. Corrosion inhibitors have a good track record when used correctly, but require constant injection, albeit of concentrations in the order of 10 parts per

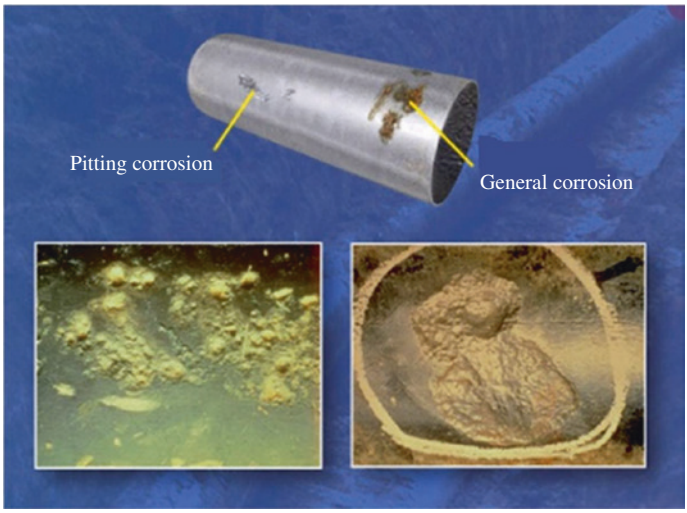


Fig. 2.12 Corrosion [21]

million. The inclusion of water within oil and product pipelines can cause considerable problems when a pipeline is left dormant for long periods of time when full or partly full of product. The only satisfactory way to mothball a pipeline is to clear the contents and replace them with an inert dry liquid or gas. This is, however, often not feasible for short durations, which over a period of a few years, can lead to substantial internal corrosion, commonly located about the 6 o'clock position.

(iv) **Flooding**

Flooding is one of the serious issues along the route of gas transmissions as shown in Fig. 2.13 typical scenario of flooding.

2.1.5 Safety in Pipelines Design and Operations

Safety is of paramount importance and in terms of protection of the asset a pipeline with a low risk also has a low probability of failure, thus keeping it in service for longer. Any design cannot make a pipeline absolutely safe and will not reduce the inherent risk involved in transporting a hazardous fluid. What it can do is make that risk sufficiently low so that it meets legal and project requirements. It also provides the basis for any assessment made as to the effort and cost involved in reducing a risk even further than the initial design. Most UK safety legislation and general practice world wide is for the risk to be to as low as reasonably practicable (ALARP). This means that items that would only contribute marginally to the overall risk level but cost a significant quantity of money to install (e.g.



Fig. 2.13 Typical Flooding in transmission pipeline [21]

block valves every 500 m) can be discounted in a rational and reasoned manner. Quantified risk assessments, structural reliability assessments and the data used to make such calculations are covered elsewhere in the text. Activities that design needs to cover includes (i) identification of all the possible failure modes and actions taken to eliminate or minimise their effect (HAZID), (ii) identification of hazards during construction (HAZCON), (iii) discussion of potential hazards and all modes of operating safely (HAZOP), (iv) compilation of potential hazards and reports relating to construction, operation and maintenance under the COM regulations [9], (v) production of the major accident prevention document (MAPD) required under the PSR 1996 regulations. Having introduced very briefly the environments in which gas pipes are laid, the materials and fittings used, the influence of pressure on the mode of failure, it is important now to consider how all of these elements can be drawn together to provide a general framework whereby safe operation of gas systems can be established. This has been done in Britain. The documents described as IGE/TD/1 and IGE/TD/3 are a result of lengthy deliberations on the assessment of safety with regard to operation of piping systems at pressures above and below 7 barg. Some of the steps, based on data presented so far and other items that describe the industry's experience, are now discussed in relation to distribution systems operating at pressures below 7 barg and will affirm the approach used in IGE/TD/3. Whichever guideline is being used, it must be seen to be logical and consistent with the objective of realising safe, cost effective, gas systems.

The separation between transmission and distribution is not just a division conveniently created for reasons of operational responsibility but because of the difference in the initial premise from which hazard is assessed. For transmission systems, it is implicit in IGE/TD/1 that the objective of transmission decisions

is to prevent incidents which involve the public. As the majority of transmission pipelines pass through environments with population densities of <2.5 persons/ha, this is a technically reasonable and economically realisable objective. This is not so for distribution pipes. The majority are in environments with population densities well above 50 persons/ha. Also, there is a difference in historical development. Many distribution systems have a long history, and it would seem unreasonable to develop standards for new pipes that did not take into account some of this heritage (like the patterns, loading, strain and consequences of failure). The first priority, therefore, is to establish from the history of these older mains a level of risk acceptable to the public.

Over the past 15 years, an average of 15 incidents/annum, involving damage to property or injury to people, occurred in Britain as a result of leaks from the distribution system. (Incidents are normally described as events that incur a severe explosion, i.e. fire, damage to property of an extent > £100, personal injury or death.) The incidents are not uniformly distributed. It is known that 80 % have occurred on 20 % of the older cast iron and PE system. Accordingly, the policy is now to give this section priority for replacement. The rest it has been determined, is operating at an acceptable level of safety. It is reasonable, therefore, to use this level as the base level for the design and construction of new pipelines operating at pressures below 7 barg.

At this stage, for the purposes of analysis, the distribution mains are divided into two sets: higher risk and lower (acceptable) risk. There is a variation in risk level within these sets but a useful general guide for comparison can be established by considering the difference in the average risk levels within these sets. As indicated, the more hazardous set comprising 20 % of mains covers 80 % of incidents, then conveniently the reverse is true of the rest of the mains that comprise the other set. This gives a difference in relative average risk between the sets of mains of 160. To continue this approach, it is now important to define a further set of attributes that help to resolve each of the above sets of mains into subsets. This will then enable the examination of the relative risks of these mains and, as a result of this information, to suggest the improvement in the performance of the materials, which is required to achieve a level of risk equivalent to the remaining 80 % of the cast-iron system.

2.2 Natural Gas Distribution Networks

The planning and design of gas distribution systems is an iterative process; there is no single absolute solution. The ideal design will minimise cost while at the same time retain sufficient flexibility to allow for future changes in the pattern of gas consumption. Figure 2.14 exemplifies the UK distribution system, although it varies from country to country around the world.

Systems should be designed to meet the maximum demands placed upon them. In low-pressure systems, experience has shown that this is likely to be the

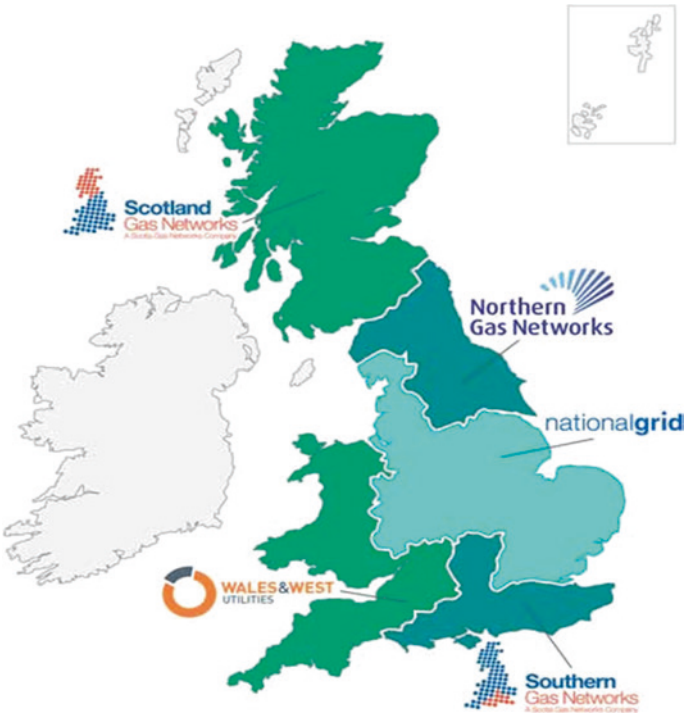


Fig. 2.14 UK distribution network [23]

maximum demand that will occur in any period of not <6 min, expressed as an hourly rate. However, it should be noted that flow rates may exceed this level for shorter durations and that instantaneous values could be up to 12.5 % higher. Where pressure differentials in excess of 25 mbar are to be used across a system; the design should be reassessed to ensure that, under peak instantaneous conditions, the minimum pressure in the system will ensure the continued safe operation of gas appliances.

For supplies to domestic estates, the design flow rate should be estimated from predicted and annual consumption. The estimate should be based upon the expected space-heating load and the rated load of any other appliances, taking into account the size and type of property. For supplies to groups of domestic consumers, diversity can be taken into account and diversity curves, which relate the peak flow for groups of consumers to the annual consumption of individual consumers within the group, are useful for this purpose. For supplies to individual industrial or commercial consumers, the design flow rate should be assessed. In the absence of specific information, it may be possible to estimate the design flow rate from the predicted annual consumption, appliance details or floor area. For supplies to groups of commercial and industrial premises and buildings and for the purposes of designing the overall system, allowance should be made for the

effects of diversity if such information is available. In the absence of information on diversity, it should be assumed that the maximum flow will occur at the peak of the system design flow rate. For systems where interruptible supplies are included, special consideration should be given to the effect of these supplies on the system design flow rate. Where a system contains a significant proportion of interruptible load, the maximum system demand may not occur under the peak 6-min conditions with interruptible supplies off, but rather at the point just prior to interruption when firm demands will be at a lower level. Under these circumstances, this higher demand level should be used as a basis for the system design flow rate. Any system designed using this revised criteria should be tested for robustness against the theoretical peak.

2.2.1 Distribution Network Design Consideration

The size of pipework and associated equipment should be determined, either by applying a suitable flow equation to a simple pipe system, or by using a sophisticated network analysis computer programme to model a more complex integrated system. Some computer programmes have cost functions, which allow the design to be optimised for least cost as well as capacity

2.2.1.1 Gas Demand

Gas demand will depend upon: (i) season (outside ambient temps); (ii) day of the week; (iii) time of day; and (iv) the nature of gas use (heating and/or cooking)—and must take into account "diversity of demand". The peak gas demand for one consumer is always less than the aggregate of all consumers' individual demands because as the number of consumers in a group increases, the probability of the coincident use of gas burning appliances decreases. In simple terms, diversity of demand (D) is defined as:

$$D = \frac{\text{Maximum potential gas demand}}{\text{Maximum actual gas demand}} \tag{2.5}$$

Historical design diversity curves have been developed for various categories of consumer for a maximum 6-min demand period [10].

2.2.1.2 Source Pressure

Systems should, generally, be designed to operate with the optimum pressure differential in order to make the best use of the available pressure without creating such a large pressure loss that the system is vulnerable to increases in gas demand. Where a suitable source of pressure is available, consideration should be given to

designing an intermediate-/medium-pressure supply system. In general, the same steps and principles apply as are used for low-pressure system design. Maximum use should be made of the pressure available, but due regard should be given to any effect on the upstream supply system. For example, its capacity to meet existing and future commitments (including interruptible loads), operating constraints, ongoing repair and maintenance requirements and possible abandonment as a result of a replacement policy [17, 18]. For demands taken from medium- and intermediate-pressure systems, source pressure will depend upon the available outlet pressure from the regulator supplying that part of the system concerned when operating at the system design flow rate. In such cases, the source pressure supplying a low-pressure system should not exceed the maximum normal operating pressure of the system. Source pressure should be based on the highest available pressure to optimise pipe sizing but subject to proximity and routing constraints. For discrete, new PE networks, the source pressure can be 75 mbar (or 2 barg for medium-pressure networks), however, where new PE networks are interconnected with older, leaking metallic pipes, the source pressure is normally limited to around 50 mbar or less [11].

2.2.1.3 System Pressure and Installations

The minimum design pressure for a gas distribution system should be that pressure, at the extremity of the system, which will provide the minimum pressure for the safe operation of customer appliances, allowing also for the pressure loss in the customer's plant pipework. This requirement also applies particularly to the operation of service pressure regulators. Supplies to consumers from a medium-pressure distribution main should provide adequate pressure to ensure proper control by the service regulator which should, normally, be located outside of any building. The pressure at which gas is supplied inside domestic buildings should not, normally, exceed 75 mbar but [12], where a higher pressure is considered, account should be taken of any risks involved. Where it is known that a gas compressor or booster is to be used, to increase pressure at a customer's premises, protection of the gas distribution system may be required by way of antifluctuators or valves. Pressure-regulating installations should normally be designed to pass the anticipated peak flow rate at the minimum expected inlet pressure and the maximum likely outlet pressure. The design should be based upon the peak flow rate likely to be experienced under normal supply conditions. However, the effects of any instantaneous flow rates that could be imposed by the downstream configuration should be considered. The design of installations should use a medium-term planning horizon, 10 years, based on network analysis predictions. Consideration should be given to any peculiarities of the system, for example the presence of interruptible consumers where the local load before interruption may be higher than at peak, or the effect of nearby non-domestic consumers on the normal load pattern. If any regulator installation is expected to provide security for another installation, adequate capacity should be provided within the installation.

Such additional capacity will, normally, be contained within the standby stream of a dual stream installation. The configuration of equipment within a regulator installation will affect the pressure loss across the station and therefore the capacity. Manufacturer's data apply only to their individual items of equipment, it is the responsibility of the design engineer to ensure that combinations of equipment will still deliver the required quantities of gas.

2.2.1.4 Distribution Velocity

Gas Velocity should be limited to 20 m/s [20] in systems where there is a dust problem, particularly where the new network is interconnected to an older, metallic system. Otherwise, gas velocities up to 40 ms^{-1} are acceptable. The velocity equations used for low-pressure and higher (than LP)-pressure pipelines are described in Sect. 4.1 [13].

2.2.1.5 Pipe Materials

Pipe material is invariably polyethylene (PE) for new pipelines and may be PE80 or PE100 grades of pipe. Long-term testing using linear regression analysis has established that PE pipes will have a lifetime of at least 50 years when subject to a constant hoop stress of 80 bar for PE80 and 100 bar for PE100 as shown typically in Fig. 2.15, at a constant temperature of 20 °C. PE100 pipes can operate at significantly higher operating pressures than PE80 pipes for the same wall thickness. Similar calculations are performed for other type of pipe materials.

The relationship between hoop stress and internal operating pressure is expressed in terms of hoop stress and thickness as

$$\text{Hoop stress} \tag{2.6}$$

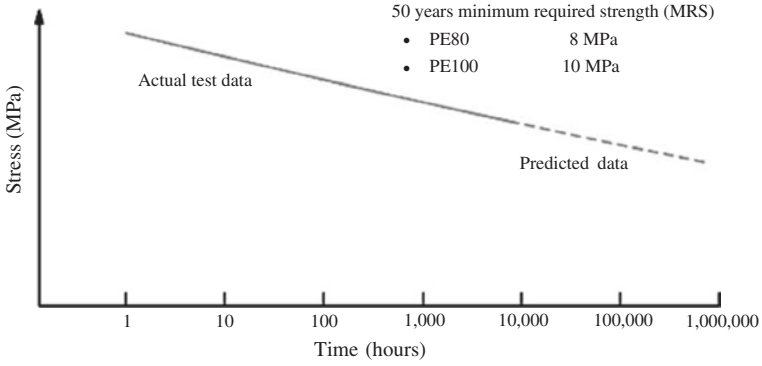


Fig. 2.15 Stress per hourly distribution [21]

The maximum operating pressure (MOP) for the PE pipe is then made on the basis that it does not exceed 10 barg and that the overall design coefficient ("C") shall be greater than or equal to 2, such that [14, 15]:

$$C = \frac{20 \times \text{MRS}}{\text{MOP} \times (\text{SDR} - 1) \times D_f} \tag{2.7}$$

where MRS is the minimum required strength (long-term hydrostatic strength) and D_f is a de-rating factor to allow for the influence of increased operating temperatures—i.e. 1.1 at 30 °C and 1.3 at 40 °C. The MOP in GB has typically been limited to 4 barg for PE80 (5.5 barg for pipes of diameter ≤140 mm) and to 7 barg for PE100 within the temperature range. PE pipe has good "rapid crack propagation" (RCP). The critical RCP level is the pressure at which a crack can rapidly propagate through a pipeline (at a reference temp. of 0 °C). Sinusoidal crack at speed of sound over long distances.

Pipeline failure may result in catastrophic explosion of pipeline due to wrong pipe thickness or overpressure as shown in Fig. 2.16 for a typical failed section prior to pipe laying process (Fig. 2.17) from the stock PE pipes shown in Fig. 2.18.



Fig. 2.16 Typified pipeline failure in PE pipes [21]



Fig. 2.17 Typical PE pipeline used in construction [21]

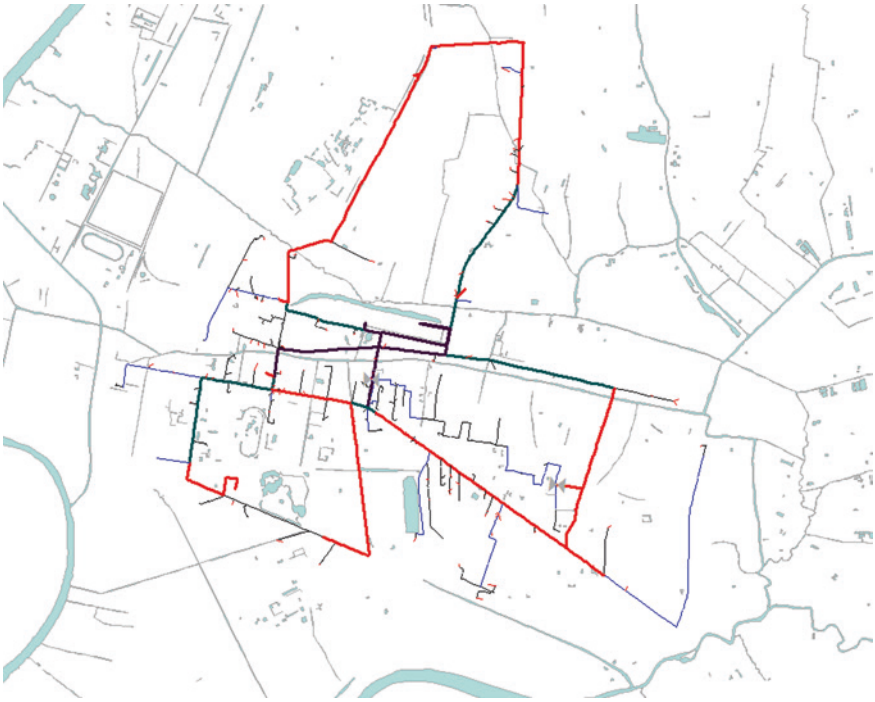


Fig. 2.18 Computer-aided network design [24]

2.2.2 Computer-Aided Design

There are a variety of computer programmes available that can be used to calculate pipe diameters for a specific pipe or system configuration. Any computer-aided system design should be checked by a competent person, before implementation, to ensure its practicability. In some cases, special features are incorporated, including, (i) fixing certain pipe sizes, for example to specify the route of distribution mains supplying a new housing development or industrial estate, (ii) restricting regulator and/or other source flows where one limited capacity is available, (iii) security runs to determine whether looping 1 s necessary and/or to test the effect of regulator failure, and (iv) diversity runs, where the number of consumers supplied from each node is used in a diversity model to calculate the varying diversified flow throughout the system. Specifying pressures at particular points to allow for future extensions, further analysis of simulation-aided design is discussed in Sect. 4.2. Figure 2.18 shows a typical computer-aided network design.

Modern, sophisticated network design and modelling tools can develop a distribution network design very easily. They also allow simulation analysis and validation from actual pressure data in the field.

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Chapter 3
Liquefied Natural Gas

3.1 Liquefied Natural Gas

3.1.1 Introduction

In the late 1920s, transportation of liquefied gases in bulk started. In the very beginning, it was transportation of propane and butane in fully pressurised tanks. When the steel quality became better and the knowledge about propane and butane was better, they started to carry those liquefied gases under temperature control. From the mid-1960s, we have carried fully refrigerated liquefied gases, and now the biggest gas carriers are more than 125,000 m^3. The biggest gas carriers in the world are Q-max of Cargo capacity: 265,977.7 m^3 at 100 %. Liquefied gas is divided into different groups based on boiling point, chemical bindings, toxicity and flammability. The different groups of gases have led to different types of gas carriers and cargo containment systems for gas carriers.

Economical pipeline supply is limited to around 2,500 km. Methane gas volume is reduced by 600 times and can be stockpiled, and therefore, longer distances make delivery by the liquefied natural gas (LNG) carriers economical. The LNG value chain represents $7–$14 billion dollars of investment from start to finish. It begins with natural gas extracted from underground reservoirs and is sent through a pipeline to a liquefaction facility.

At the liquefaction facility, impurities are removed from the gas, and it is sent through three cooling processes until it reaches a final temperature of −160 °C.

The chilled gas, now LNG, is then loaded onto specially designed tanker ships where it will be kept chilled for the duration of the voyage, which may last anywhere from 4 to 30 days, depending on the destination port. Once the ship arrives at a regasification terminal, the LNG is offloaded into large storage tanks, built with full-containment walls and systems to keep the LNG cold until it is turned

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Fig. 3.1 LNG value chain [1]

back into a gaseous state. A schematic pattern of natural gas processing trends is shown in Fig. 3.1 as typical example for most gas treatment and delivery processes.

3.1.2 Physical Properties and Composition of LNG

LNG results from a liquefaction process which consists in removing, by means of a condensation–vaporisation cycle, the sensible and latent heats of natural gas. Compared to natural gas, the volume of LNG is reduced by a factor of approximately 600. Except for nitrogen, the components of LNG are hydrocarbons whose molecules contain only hydrogen and carbon atoms. Their general formula is C_nH_{2n+2}; they are "saturated" since the hydrogen atoms are singly united with carbon atoms.

LNG is chemically non-reactive. It does not present compatibility problems with air, water and materials commonly used in the cargo tanks and handling installations. In the presence of moisture, however, LNG can form hydrates. Liquefied natural gas (LNG) is a clear, colourless and odourless liquid. The LNG is usually transported and stored at a temperature very close to its boiling point at atmospheric pressure which is approximately −160 °C. Its density (less than half as dense as water), the heating value and other properties depend on its composition.

The main concern is accordingly with the properties of methane. The properties of the other constituents must not be neglected, however, particularly with regard to possible increases in concentration of higher hydrocarbons after evaporation of LNG from insulated containers.

Some of the combustion properties that determine the likelihood of a flammable gas or liquid catching fire are given in Table 3.1. It is seen that methane diffuses more rapidly and is more buoyant than all other gaseous fuels except hydrogen.

The actual composition of the LNG will vary depending on its source and on the liquefaction process, but the main constituent will always be methane. Other

Table 3.1 Physical properties of LNG [1]

	Methane CH_4	Ethane C_2H_6	Propane C_3H_8	Butane C_4H_{10}	Pentane C_5H_{12}	Nitrogen N_2
Molecular weight	16.042	30.068	44.096	58.120	72.150	28.016
Boiling point at 1 bar absolute (°C)	−161	−88.6	−42.1	−0.5	36.1	−195.8
Liquid density at boiling point (kg/m^3)	426	544.1	580.7	601.8	610.2	808.6
Vapour RD at 15 C and 1 bar absolute	0.553	1.04	1.55	2.00	2.49	0.97
Gas volume liquid ratio at boiling point and 1 bar abs.	619	431	311	222	205	694
Flammable limits in air by volume (%)	5.3–14	3.1–12.5	2.1–9.5	1.8–8.5	3–12.4	Non-flammable
Auto-ignition temperature (°C)	595	510	468	365,500	-	-
Gross heating value at 15 °C (kJ/kg)	55,550	51,870	50,360	49,520	49,010	

constituents will be small percentages of heavier hydrocarbons such as ethane, propane, butane, pentane and possibly a small percentage of nitrogen. The physical properties of the major constituent gases are given in Table 3.1. During a normal sea voyage, heat is transferred to the LNG cargo through the cargo tank insulation, causing part of the cargo to vaporise, i.e. "boil-off". The composition of the LNG is changed by this boil-off because the lighter components which have lower boiling points at atmospheric pressure vaporise first. The discharged LNG therefore has a lower percentage content of nitrogen and methane than the LNG that was loaded and a slightly higher percentage of ethane, propane and butane, due to methane and nitrogen boiling off in preference to the heavier gases.

The flammability range of methane in air (21 % oxygen) is approximately 5.3–14 % (by volume). To reduce this range, the oxygen content is reduced to 2 %, using inert gas from the inert gas generators, prior to loading after dry dock. In theory, an explosion cannot occur if the O_2 content of the mixture is below 13 % regardless of the percentage of methane, but for practical safety reasons, purging is continued until the O_2 content is below 2 %. This safety aspect is explained in detail later in this section. The boil-off vapour from LNG is lighter than air at vapour temperatures above −100 °C or higher depending on the LNG composition; therefore, when vapour is vented to atmosphere, the vapour will tend to rise above the vent outlet and will be rapidly dispersed. When cold vapour is mixed with ambient air, the vapour air mixture will appear as a readily visible white cloud due to the condensation of the moisture in the air. It is normally safe to assume that the flammable range of vapour–air mixture does not extend significantly beyond the perimeter of the white cloud. The auto-ignition temperature of methane, i.e. the lowest temperature to which the gas needs to be heated to cause self-sustained combustion without ignition by a spark or flame is 595 °C.

At atmospheric pressure, LNG, depending on its composition, will boil at a temperature in the range −157 to −166 °C. At an absolute pressure of 13.8 bar, LNG can be stored at temperatures up to −ll8 °C. The presence of nitrogen in LNG depresses the boiling point by about 2.5 °C per volume per cent of nitrogen. The density of LNG is determined by its composition, temperature and pressure. Nitrogen increases the density by 5 kg/m^3 per volume per cent. If the temperature rises such that the vapour pressure increases by about 70 mbar. The density decreases by about 1.3 kg/m^3.

3.1.2.1 Evaporation of Liquefied Natural Gas

When LNG is poured on to the ground, there is an initial period of film boiling, after which the rate of evaporation slows rapidly at a rate determined by the characteristics of the ground and the liquid. Rates of evaporation are of the order of 1.5 m^3 (st) of gas per m^2 of liquefied surface per minute after the first few minutes. This must be related to the rates of mixing and pocketing of the gas produced. If a combustible mixture is ignited, the flame will travel through the flammable zone and will tend to heat the liquid by radiation. This will increase the rate of

evaporation and hence the burning rate to values that could be as high as 8 m^3 (st)/m2 surface per min.

The volumes of gas in m^3 produced per tonne of LNG vaporised for the gases are Algerian 1,260, Libyan 1,030 and North Sea 1,440. The gross C.V of LNG is approximately 24 MJ/m^3, and about 610 KJ are required to vaporise enough LNG to form 1 m^3 of gas at ambient conditions. Immediately after evaporation, the natural gas vapour is heavier than air and remains so until the temperature has risen above −ll3 °C. As the cold natural gas mixes with the warmer air and the temperature rises towards ambient, its natural buoyancy and rapid dispersion ensure that the flammable zone extends only a few pool diameters from the edge of a pool of liquid once the initial rush of cold gas has passed. Experiments with an open flame near ground level have failed to ignite the vapour cloud from a 6-m (20 ft) pool of LNG at a distance greater than 2.1 m (7 ft) when a fairly strong breeze was blowing. The visible cloud formed by the condensation of water vapour usually extends beyond the boundaries of the flammable zone.

3.1.2.2 Variation of Boiling Point of Methane with Pressure

The Fig. 3.2 shows the vapour pressure diagram of liquid cargoes. The boiling point of methane increases with pressure, and this variation is shown in the diagram for pure methane over the normal range of pressures on board the vessel. The presence of the heavier components in LNG increases the boiling point of the cargo for a given pressure. The relationship between boiling point and pressure of LNG will approximately follow a line parallel to that shown for 100 % methane.

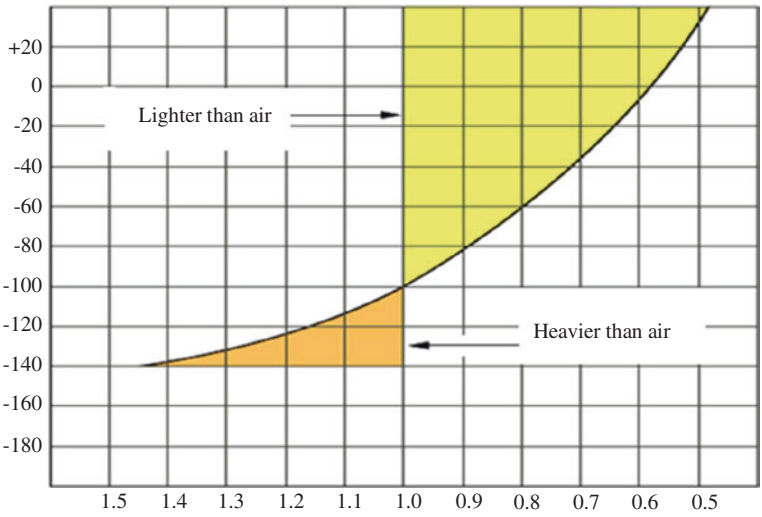


Fig. 3.2 Dependency of density on temperature for LNG [1]

3.2 Characteristics of LNG

3.2.1 Flammability of Methane, Oxygen and Nitrogen Mixtures

The ship must be operated in such a way that a flammable mixture of methane and air is avoided at all times. The relationship between the gas/air composition and flammability for all possible mixtures of methane, air and nitrogen are shown in Fig. 3.3. The vertical axis A–B represents oxygen–nitrogen mixtures with no methane present, ranging from 0 % oxygen (100 % nitrogen) at point A to 21 % oxygen (79 % nitrogen) at point B. The latter point represents the composition of atmospheric air. The horizontal axis A–C represents methane–nitrogen mixtures with no oxygen present, ranging from 0 % methane (100 % nitrogen) at point A to 100 % methane (0 % nitrogen) at point C. Any single point on the diagram within the triangle ABC represents a mixture of all three components, methane, oxygen and nitrogen, each present in specific proportion of the total volume. The proportions of the three components represented by a single point can be read off the diagram. For example, at point D: methane: 6.0 % (read on axis A–C), oxygen: 12.2 % (read on axis A–B) and nitrogen: 81.8 % (remainder).

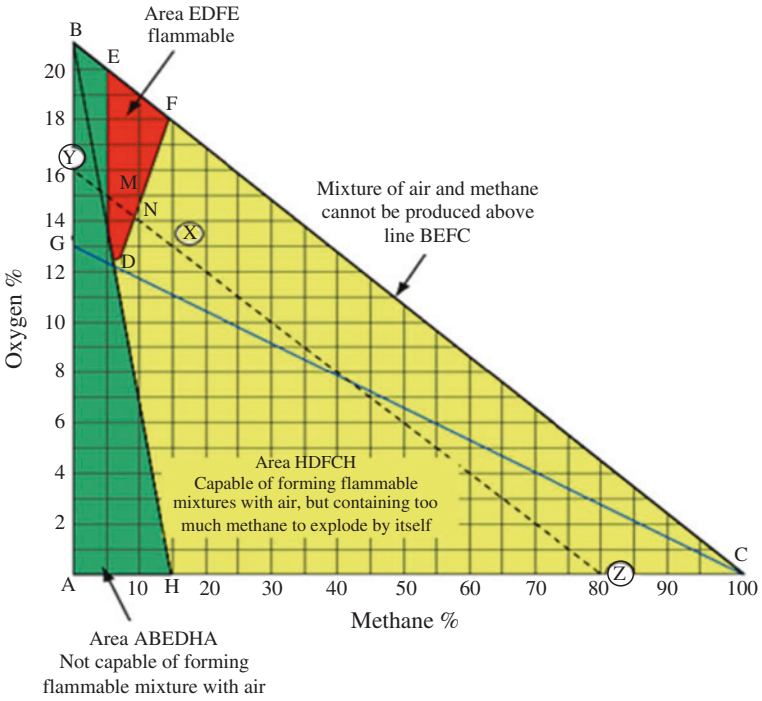


Fig. 3.3 Flammability of methane, oxygen and nitrogen mixtures [1]

The diagram consists of three major sectors as shown in Fig. 3.3:

1. The Flammable Zone Area EDF. Any mixture whose composition is represented by a point which lies within this area is flammable.
2. Area HDFC. Any mixture whose composition is represented by a point which lies within this area is capable of forming a flammable mixture when mixed with air, but contains too much methane to ignite.
3. Area ABEDH. Any mixture whose composition is represented by a point which lies within this area is not capable of forming a flammable mixture when mixed with air.

Assume that point Y on the oxygen–nitrogen axis is joined by a straight line to point Z on the methane–nitrogen axis. If an oxygen–nitrogen mixture of composition Y is mixed with a methane–nitrogen mixture of composition Z, the composition of the resulting mixture will, at all times, be represented by point X, which will move from Y to Z as increasing quantities of mixture Z are added. In this example, point X, representing changing composition, passes through the flammable zone EDF, that is, when the methane content of the mixture is between 5.5 % at point M and 9.0 % at point N. Applying this to the process of inerting a cargo tank prior to cool down, assume that the tank is initially full of air at point B. Nitrogen is added until the oxygen content is reduced to 13 % at point G. The addition of methane will cause the mixture composition to change along the line GDC which, it will be noted, does not pass through the flammable zone, but is tangential to it at point D. If the oxygen content is reduced further, before the addition of methane, to any point between 0 and 13 %, that is, between points A and G, the change in composition with the addition of methane will not pass through the flammable zone.

Theoretically, therefore, it is only necessary to add nitrogen to air when inerting until the oxygen content is reduced to 13 %. However, the oxygen content is reduced to 2 % during inerting because, in practice, complete mixing of air and nitrogen may not occur. When a tank full of methane gas is to be inerted with nitrogen prior to aeration, a similar procedure is followed. Assume that nitrogen is added to the tank containing methane at point C until the methane content is reduced to about 14 % at point H. As air is added, the mixture composition will change along line HDB, which, as before, is tangential at D to the flammable zone, but does not pass through it. For the same reasons as when inerting from a tank containing air, when inerting a tank full of methane, it is necessary to go well below the theoretical figure to a methane content of 5 % because complete mixing of methane and nitrogen may not occur in practice.

The procedures for avoiding flammable mixtures in cargo tanks and piping are summarised as follows:

1. In practice, the tanks and piping containing air are to be inerted with inert gas provided by the inert gas generator, as the nitrogen generator capacity is far too small to inert the cargo tanks, before admitting methane until all sampling points indicate 5 % or less oxygen content.

2. Tanks and piping containing methane are to be inerted with inert gas provided by the inert gas generator before admitting air until all sampling points indicate 5 % methane. It should be noted that some portable instruments for measuring methane content are based on oxidising the sample over a heated platinum wire and measuring the increased temperature from this combustion. This type of analyser will not work with methane–nitrogen mixtures that do not contain oxygen. For this reason, special portable instruments of the infrared type have been developed and are supplied to the ship for this purpose.

3.2.2 Supplementary Characteristics

3.2.2.1 When Spilt on Water

1. The boiling of LNG is rapid due to the large temperature difference between the product and the water.
2. LNG continuously spreads over an indefinitely large area, and it results in a magnification of its rate of evaporation until vaporisation is complete.
3. No coherent ice layer forms on the water.
4. Under particular circumstances, with a methane concentration below 40 %, flameless explosions are possible when the LNG strikes the water. It results from an interfacial phenomenon in which LNG becomes locally superheated at a maximum limit until a rapid boiling occurs. However, commercial LNG is far richer in methane than 40 % and would require lengthy storage before ageing to that concentration.
5. The flammable cloud of LNG and air may extend for large distances downwind (only methane when warmer than −100 °C is lighter than air) because of the absence of topographic features which normally promote turbulent mixing.

3.2.2.2 When Agitated By Water

When a flange drip tray becomes filled with LNG as a result of a leaking flange, under no circumstances should a water jet be directed into the drip tray. Such action will cause a severe eruption and a rapid expansion/boiling of the LNG within the tray, resulting in LNG and ice particles being blasted outwards. The LNG should be allowed to boil-off naturally or the drip tray warmed with water spray on the sides or base.

3.2.2.3 Vapour Clouds

1. If there is no immediate ignition of an LNG spill, a vapour cloud may form. The vapour cloud is long, thin and cigar shaped and, under certain

meteorological conditions, may travel a considerable distance before its concentration falls below the lower flammable limit. This concentration is important, for the cloud could ignite and burn, with the flame travelling back towards the originating pool. The cold vapour has a higher density than air and thus, at least initially, hugs the surface. Weather conditions largely determine the cloud dilution rate, with a thermal inversion greatly lengthening the distance travelled before the cloud becomes non-flammable.

2. The major danger from an LNG vapour cloud occurs when it is ignited. The heat from such a fire is a major problem. A deflagration (simple burning) is probably fatal to those within the cloud and outside buildings but is not a major threat to those beyond the cloud, though there will be burns from thermal radiation.

3.2.2.4 Reactivity

Methane is an asphyxiant in high concentrations because it dilutes the amount of oxygen in the air below that necessary to support life. Due to its inactivity, methane is not a significant air pollutant, and due to its insolubility, inactivity and volatility, it is not considered to be a water pollutant.

3.2.2.5 Cryogenic Temperatures

Contact with LNG or with materials chilled to its temperature of about −160 °C will damage living tissue. Most metals lose their ductility at these temperatures; LNG may cause the brittle fracture of many materials. In the case of LNG spillage onto the ship's deck, the high thermal stresses generated from the restricted possibilities of contraction of the plating will result in the fracture of the steel.

3.2.2.6 Behaviour of LNG in the Cargo Tanks

When loaded into the cargo tanks, the pressure of the vapour phase is maintained substantially constant at slightly above atmospheric pressure. The external heat passing through the tank insulation helps generate convection currents within the LNG tank causing warmer LNG to rise to the surface where it then boils-off. The heat necessary for vaporisation comes from the LNG. As long as the vapour is continuously removed by maintaining the pressure as substantially constant, the LNG remains at its boiling temperature. If the vapour pressure is reduced by removing more vapour than is generated, the LNG temperature will decrease. In order to make up the equilibrium pressure corresponding to its temperature, the vaporisation of LNG is accelerated, resulting in an increased heat transfer from LNG to vapour.

LNG is a mixture of several components with different physical properties, particularly the vaporisation rates; the more volatile fraction of the cargo vaporises at

a greater rate than the less volatile fraction. The vapour generated by the boiling of the cargo contains a higher concentration of the more volatile fraction than the LNG. The properties of the LNG, i.e. the boiling point and density, have a tendency to increase during the voyage.

Nitrogen is used on board for the pressurisation of the cargo tank wedge and insulation spaces, the purging of cargo pipelines and heaters, boiler gas lines and Whessoe gauges and for the sealing of the LNG compressors. It is produced by the nitrogen generators whose principle is based on hollow fibre membranes to separate air into nitrogen and oxygen.

3.3 Natural Gas Liquefaction Plants

3.3.1 Introduction

The refrigeration and liquefaction sections of any LNG project are very costly items of equipment. A number of cycles have been developed to achieve the low temperatures required. Brief descriptions of a small number of process routes will be given; fuller details can be obtained from Sect. 3.4.

Some twelve countries were operating by 2003. The cycles employed are the classical cascade, the mixed refrigerant and the propane pre-cooled mixed refrigerant. The process selected must ensure high onstream factors, reliability of equipment, flexibility and ease of operations and guaranteed capacities. Small differences in projected thermodynamic efficiencies are usually less than the uncertainties in equipment performance. A proper choice must be made between the use of well-proven technology and the acceptance of innovations designed to improve efficiencies and reliabilities.

In addition to the liquefaction equipment, a facility must have extensive gas purification equipment, designed to operate at high pressure, together with large compression machinery, special materials of construction, complex heat exchangers and large tankage.

3.3.2 LNG Purification

Gas piped to a liquefaction plant from fields around 100 miles away is usually only given the minimum purification necessary at the well head. The feed gas may therefore contain water, carbon dioxide, hydrogen sulphide, higher hydrocarbons and other impurities. The line must be pigged regularly to prevent blockage and irregularities due to two-phase flow. The first-stage treatment will therefore comprises traps and facilities to collect liquids. Depending on ambient temperature, water content and pressure drop, glycol or methanol can be injected to prevent hydrate formation. In this case, glycol/methanol recovery involving fractionation will be required on the aqueous layer in the gas/liquid separator.

After reduction of liquid water, glycol and heavier hydrocarbons by simple gas/liquid separation, the gas is cooled by heat exchange to a few degrees below freezing. At pipeline pressure, this results in further condensation, and more water and heavy hydrocarbons separate out in a knockout drum.

The next stage, gas sweetening, removes H_2S and CO_2. The reagents used can operate either by a reversible chemical reaction between the acid gas and the solvent or by the acid gas dissolving in the absorber liquid in preference to the other gas components.

In the case of physical absorption, equilibrium concentrations of H_2S and CO_2 in the liquid are strictly proportional to the partial pressures of the gases. Reactive solvents, however, have absorption equilibria independent of the gas partial pressure.

Most solvents used for absorption of acid gases are non-or only partially selective. CO_2, H_2S and mercaptans are removed roughly in proportion to their original concentration in the gas. Typical chemically reactive solvents include aqueous solutions of most alkanolamines such as monoethanolamine (MEA), diethanolamine (DEA), diglycolamine (DGA), di-isopropanolamine (Adip), triethanolamine (TEA) and anthraquinone disulphonic acid (Stretford solution). Apart from the latter, the acid gases are absorbed at near ambient temperature by the alkaline compound and are released by heating to near its boiling point. The Stretford solution, which also contains sodium vanadate, sodium carbonate and a trace of chelated iron, when blown with air, oxidises HS to elemental S which can be removed by filtration.

Another series of absorption solvents are based on potassium carbonate and act similarly to the alkanolamines. In the Benfield, Vetrocoke and Catacarb processes, the $C0_2$ reacts to form bicarbonate, which decomposes at elevated temperatures. A similar reaction occurs with H_2S, and here, various additives, often arsenates, assist H_2S removal by forming thioarsenates which decompose into arsenates and elemental sulphur as in the Giammarco Vetrocoke process.

Physical absorbents for acid gases include anhydrous propylene carbonate (Fluor solvent), *N*-methyl-2-pyrrolidone (Purisol), the dimethyl ether of polyethylene glycol (Trigol). A disadvantage compared with chemical absorbents is a tendency to remove higher hydrocarbons from the gas. This can cause problems in Claus plant. The main disadvantage is the corrosive nature of absorbents and absorbent–acid compounds. Hybrid processes have been developed to alleviate the problem. An example is the sulfinol process which uses a mixture of the physical solvent sulpholane and chemical absorbents of the alkanolamine type.

The choice of process for a particular application depends on:

1. The original pressure of the gas—high pressure assists physical absorbtion
2. The original concentration of the acid gases—in general, chemical absorbents have a higher absorption capacity
3. The relative concentrations of the acid impurities—some solvents are more selective than others
4. The permissible residual concentration of the impurities
5. The presence or absence of COS and CS_2.

After sweetening, the gas is generally saturated with water which obviously has to be removed before liquefaction. Drying can be carried out by

1. Simple refrigeration: for instance in a turbo expander.
2. Glycol dehydration: in a counter-current scrubber using di-, tri- or tetraethylene glycol at gas temperatures from 15 to 65 °C. The saturated glycol is continuously regenerated by stripping at 200 °C. Gas dew points of −70 °C can be achieved. Higher hydrocarbons are removed separately.
3. Solid desiccant adsorption: using desiccants such as silica gel, alumina or molecular sieves. The gas is passed at pressure through a packed adsorption vessel usually in a two-tower system where both hydrocarbons and water are removed. The towers are switched for regeneration by hot gas flow through the bed. The water and hydrocarbons can be recovered by cooling and condensation. Hydrocarbon and water dew points of −50 to −70 °C can be achieved.

The use of certain molecular sieves permits selective separation of water, hydrocarbons sulphur compounds and carbon dioxide. However, the adsorption capacities of the sieves limit these processes to final gas clean-up.

The level of impurity that can be tolerated in liquefaction plants depends on the actual process chosen. Generally, the water content should be less than 1 ppm, carbon dioxide concentration 50–150 ppm, hydrogen sulphide could be as high as 30–50 ppm, but considerations of odour, corrosion and toxicity restrict it to a maximum of 3 ppm. Higher hydrocarbons could be removed by separation facilities in the liquefaction section of the plant, but removal prior to liquefaction is better.

3.4 LNG Liquefaction Processes

Refrigeration is based on the conversion of internal energy of a fluid into external work, and the second law of thermodynamics imposes a limit on the efficiency with which such a conversion can be carried out. To operate between two temperatures, e.g. the boiling point of a fluid at compressor exits pressure T_B and ambient temperature T_A. A fluid undergoing a closed cycle would absorb a minimum amount of mechanical energy as expressed as:

$$W = Q = \frac{T_A - T_B}{T_B} \tag{3.1}$$

In order to transfer energy equivalent to Q from T_B to T_A, the efficiency term $(T_A - T_B)/T_B$ defines the ideal (reversible) energy required for refrigeration. In practice, the efficiency of actual processes falls seriously short of such values. To convert the internal heat of refrigeration into mechanical energy, the compressed gas can be expanded through an orifice and its temperature lowered by the Joule–Thompson effect, or energy can be recovered by extracting work from the expanding gas in an engine.

Table 3.2 Approximate dew points °C

	Pressure (bar)				
	1.0	3.4	6.85	17.1	34.2
Methane	−159	−144	−133	−92	−71
Ethane	−91	−63	−44	−15.8	
Propane	−46	−12	+12	–	
Ethylene	−104	−80	−62	–	
Propylene	−49	−18	−5.6	−8.3	
Nitrogen	–	–	–	–	

To liquefy a boiling gas, it has to be cooled below its dew point (the temperature at which condensation starts taking place at a given pressure). Table 3.2 lists dew points for a number of gas pressures. To cool a gas, heat energy has to be removed from the compressed gas either by means of cooling water, if this is possible, or by means of an evaporating refrigerant if the temperature of heat removal is lower than ambient. Table 3.2 shows that only propane can be liquefied by heat exchange with cooling water at moderately high pressure. All lower boiling gases require refrigeration by means of a refrigerant prior to a final compression step which results in liquefaction.

Since sensible heats of gases are lower than their latent heats, latent heats are used to transfer the bulk of the refrigeration energy. Any intermediate refrigerants will boil at atmospheric or slightly lower pressures, at some intermediate temperature between the boiling point of the gas which is to be liquefied and ambient temperature. A single refrigerant will as a rule permit cooling by 60–90 °C; if an even lower temperature is required, a number of refrigerants may be needed to act as intermediaries in the overall transfer of heat from the cold gas to cooling water or air.

3.4.1 Classical Cascade

This cycle uses three separate refrigerants, propane, ethylene and methane, in three compression refrigeration cycles operating at successively lower temperatures with the lower temperature cycles each rejecting heat to the next warmer cycle, i.e. cascading on to it. Operating conditions for a cascade process are therefore largely defined once the number and type of refrigerants have been selected. Appropriate temperatures and pressures are given to Fig. 3.4 which slows a simplified flow sheet for the process. Since cooling water or ambient air is the cheapest means of refrigeration, each gas is first cooled to ambient temperature before being heat exchanged with the condensed refrigerant. Also, each of the refrigerants is cooled by heat exchanger with its own vapour (not shown) before expansion through the throttling valve. Finally, each of the higher boiling refrigerants can be expanded at more than one pressure, e.g. propane liquid at medium pressure pre-cools both ethylene and methane; this is followed by a second expansion at lower pressure to cool methane only. Ethylene is also used at two pressures to

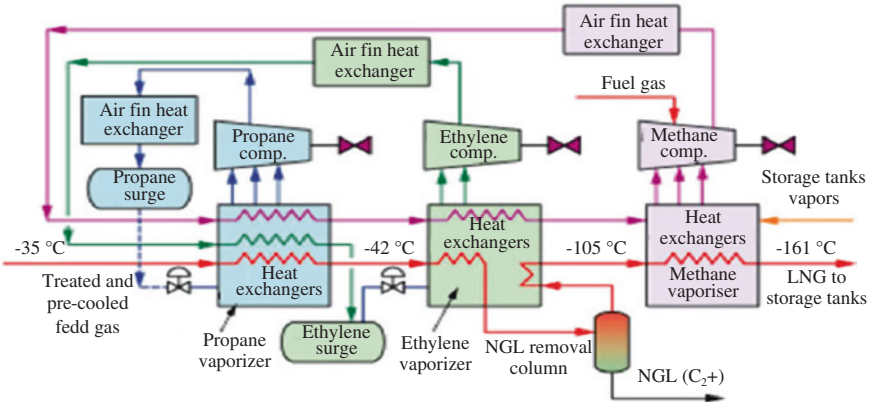


Fig. 3.4 Modified cascade [2]

cool the methane in two stages before its final expansion (not shown). The first two base load plants built in Algeria and Alaska used cycles of this type.

3.4.2 Modified Cascade Cycles (Mixed Refrigerant Cycles)

In these cycles, refrigerant circulation is confined to a single stream, and by operating a number of stages of heat exchange at different pressures, the mixed refrigerant produced using the heavier hydrocarbons from the natural gas itself can be pumped as a single fluid. Figure 3.5 shows a simplified flow diagram for a closed cycle mixed refrigerant cycle plant. The refrigerant stream consists of methane, ethane, propane and butane compressed to about 40 bar in a two stage compressor, cooled with cooling water and split in a knockout drum into a condensed propane/butane

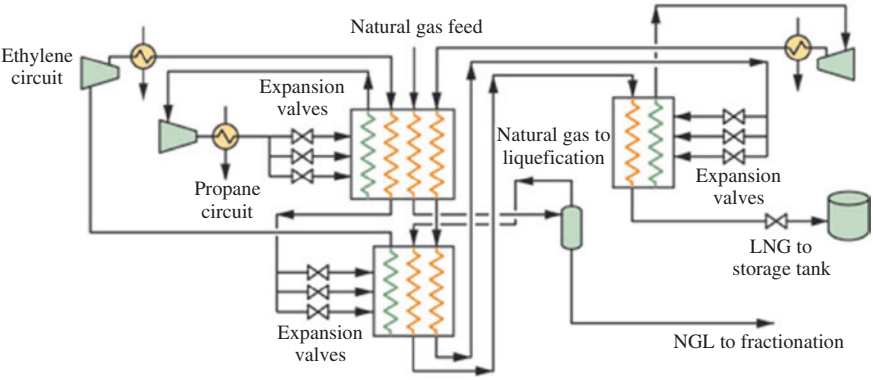


Fig. 3.5 Modified cascade [2]

stream and an overhead consisting of the lighter components. Both streams, together with the natural gas feed, pass to the first heat exchanger (E.l) at the inlet the heavy stream is expanded to compressor inlet pressure and returned through the third passage of the exchanger. The overhead after pre-cooling in E.l is split in a second knockout vessel, the ethane separator into condensate which is expanded at the inlet of the second exchanger (E.2) through which it passes in counter-current with the light overhead, which also passes through E.2 and goes into E.3. After expansion to compressor inlet pressure, it joins the low-pressure propane/butane vapour from E.l having first exchanged its cold with the liquid ethane and gaseous feed streams in E.2. E.3 serves to further cool the natural gas feed and the uncondensed methane, which is expanded at its outlet to compressor inlet pressure, its flow is reversed, and it is mixed with the other expanded gases after heat exchange in E.3. Liquefaction of the natural gas feed is effected by passing it as a separate medium pressure stream through the three heat exchangers in series and expanding the gas through a throttling valve after the last exchanger.

The main advantages of such modified cycles are the smaller number of compressors and heat exchangers required; they also require less space, are simpler to operate and cost less to build. There is no need for a supply of or storage for refrigerant, and losses are easily made up. By careful selection of operating pressure, refrigerant composition and heat exchanger arrangement, efficient processes can be designed. The first base load plant of this type entered service in Libya in 1970.

3.4.3 Pre-cooled Mixed Refrigerant Cycle (C3 = MR Cycle)

This process was developed later and has been selected for all the most recent LNG plants. It combines the simplicity of a mixed refrigerant cycle with the efficiency of the cascade cycle and is now used at more than one half of the base load LNG plants. The process is now the preferred choice in terms of costs flexibility and efficiency; it consists of a mixed liquefaction refrigerant cycle and a separate cycle for pre-cooling the natural gas feed and the liquefaction refrigerant. The mixed refrigerant consisting of nitrogen, methane, ethane, propane and possibly some butane is compressed and cooled against an ambient temperature cooling medium and pre-cooling refrigerant. The mixture that is split into a light and heavy stream is liquefied in the bundles of a spool-wound heat exchanger and subsequently expanded into the exchanger shell to serve as refrigerant.

The refrigerant is gradually vaporised and warmed up against natural gas and high-pressure refrigerant in the bundles (auto-cooling). Refrigerant vapour from the bottom of the exchanger shell is recompressed and the loop is closed. A simplified flow diagram is shown in Fig. 3.6. Normally, pre-cooling of liquefaction refrigerant and natural gas is provided by a propane cycle consisting of kettle-type heat exchangers and three or four pressure stages. The plant built years ago in Nigeria is a double mixed type.

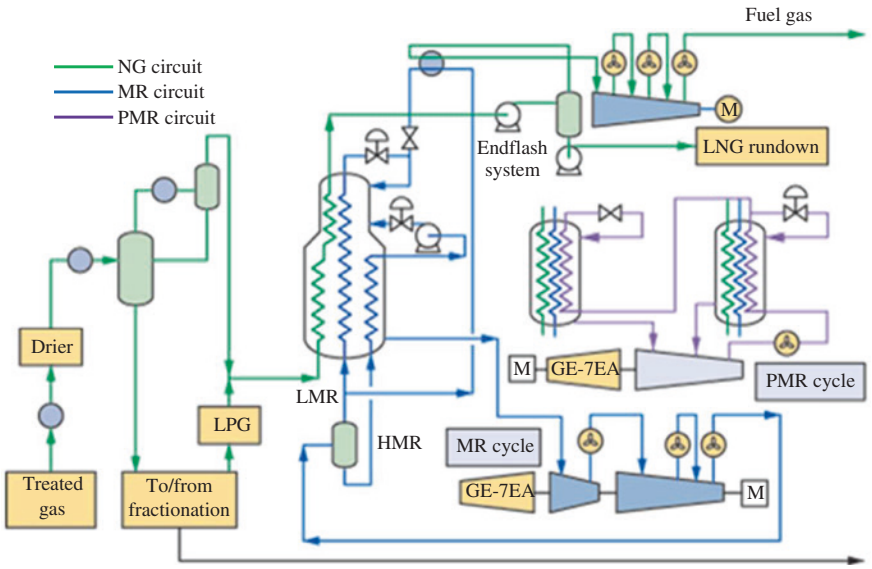


Fig. 3.6 Pre-cooled mixed refrigerant cycle [2]

3.5 LNG Import Terminal Storage Tanks and Regasifaction

An LNG import terminal must have good port access to locate the dedicated facilities to unload LNG carriers. For example, a 265,977 m^3 LNG tanker would need a water depth of at least 13 m. Each unloading berth is equipped with several unloading arms that connect the LNG tanker's manifolds with the unloading line. They incorporate articulated joints to allow for different sizes of LNG tankers and for the LNG tanker's movements as the tide rises and falls. The arms are often equipped with automatic emergency release couplings, which ensure rapid disconnection in the event of an emergency situation. Unloading line(s) runs from the arms to the storage tanks, where the LNG is stored until it is needed for vaporisation. These dedicated lines are usually made of stainless steel or some other material, such as aluminium, that can safely withstand the extremely low temperatures of LNG. They are insulated to reduce heat transfer from the atmosphere to the LNG as much as possible. Unloading lines also include expansion loops, which permit the lines to expand and contract as they warm up and cool down without overstressing the lines.

Most terminals have more than one unloading line, so that the lines can be kept cold between cargo deliveries by continuously circulating LNG through the unloading lines from storage to the berth and back again. If only one unloading line is installed, then a separate circulation line is required to feed LNG into the unloading line to keep it cold. LNG vapour is carried back to the LNG tanker's cargo system by separate vapour return line(s), to ensure that the vapour pressure

in both the shore and the LNG tanker's storage tanks is maintained within the design operating parameters for these tanks. The marine facilities may include a vapour return blower to maintain the pressure in the LNG tanker's cargo tanks. LNG is unloaded from the LNG tankers into the storage tanks using the LNG pumps installed in the LNG tanker's cargo tanks. Where the unloading line is very long, additional booster pumps may be installed between the LNG tanker and the shore tanks to assist in the cargo-pumping operation.

LNG off-loading typically requires 24 h. The purpose of an LNG storage tank is to maintain the liquid at −163 °C before it is vaporised and delivered into the local gas market. When it comes to choosing a design for the LNG import terminal's storage facilities, two main decisions have to be taken. First, decide whether to build storage tanks that are above ground or that are partially or fully buried. Second, must decide whether to build single- or full-containment storage tanks. Most terminals opt for a full-containment storage tank.

3.5.1 Largest Above-Ground Full Containment LNG Storage Tanks

Since the first introduction of liquefied natural gas (LNG), the consumption of natural gas in domestic and industrial use has been dramatically increased. The largest above-ground full-containment LNG storage tank in the world has a gross capacity of 200,000 m^3. The main objective of the development of the large capacity LNG storage tank is to reduce the construction cost and to use the construction site more effectively.

For storage of large volumes, upwards of 10,000 m^3, fully refrigerated storage in cylindrical metal tanks with a flat bottom has been the most common method for liquefied natural gas for nearly 50 years now.

Large quantities of gas can be liquefied and stored in such a tank, provided that the temperature of the stored product is maintained near its normal boiling point so that the product vapour pressure is just above atmospheric pressure.

The industry norm is a design pressure inside the storage tank of somewhere around 1–2 psi above atmospheric pressure.

3.5.2 Self-Supporting Classification

3.5.2.1 Single Containment

Single-wall tank, where only the liquid container is required to store the refrigerated product and only this primary container is required to meet low-temperature ductility requirements as part of the schematic shown in Fig. 3.7.

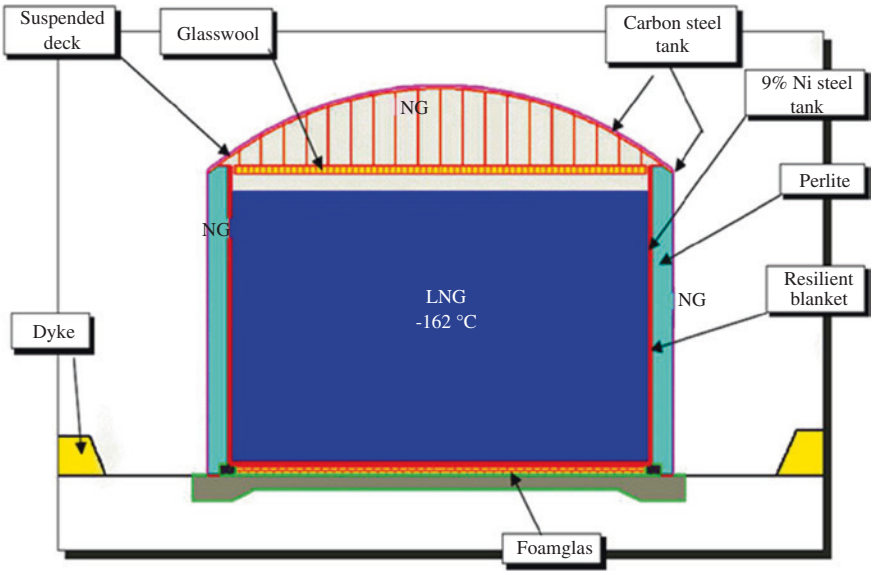


Fig. 3.7 Single containment [1]

3.5.2.2 Full Containment

Double-wall tank designed and constructed so that both inner and outer containers are capable of independently containing the stored refrigerated product as shown in Fig. 3.8 and additionally allowing for controlled release of product vapour resulting from a product leakage from the primary (inner) container resulting from a credible event.

There are two standards normally used to build LNG facilities in the world: the North American standard of NFPA 59A (Incorporates an LNG tank standard of API 620) and the European standard of EN 1473 (Incorporates an LNG tank standard of BS 7777).

According to BS7777, LNG tanks are classified into three different types. Single-containment tank is either a single tank or a tank comprising an inner tank and outer container designed and constructed so that only the inner tank is required to meet the low-temperature ductility requirements for storage of the product. Full-containment tank (Fig. 3.8) is described as a double tank designed and constructed so that both the inner tank and the outer tank are capable of independently containing the refrigerated liquid stored. The difference between the double containment and full containment is that the outer tank of a full-containment tank is intended to be capable of both containing the refrigerated liquid and of controlled venting of the vapour resulting from product leakage after a credible event.

Among these three types of LNG tanks, the full-containment type is regarded as the most advanced type. The tanks are also classified by the elevations from the ground level: above-ground type, in-ground type and under-ground type.

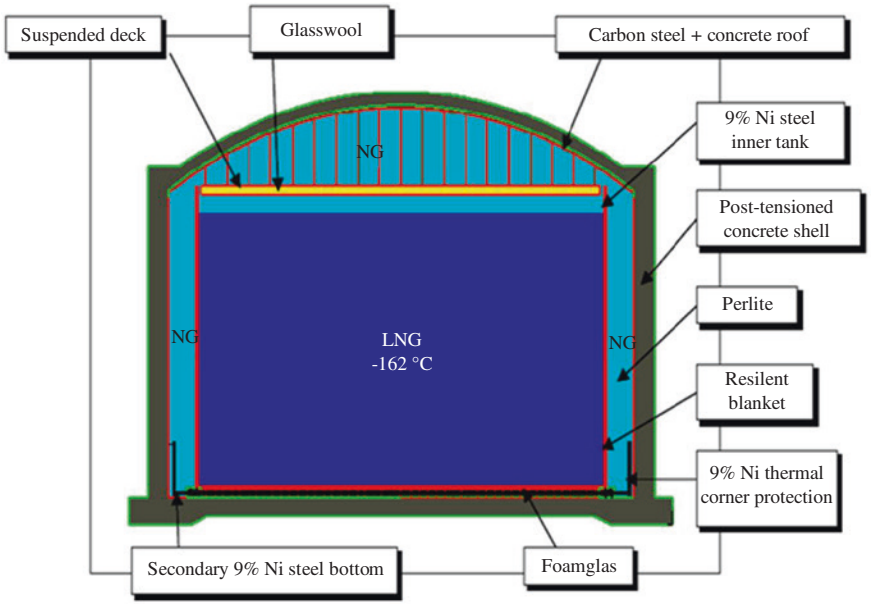


Fig. 3.8 Full containment [1]

The largest tank developed in the world is an above-ground, full-containment type diagrammatically shown in Fig. 3.9, which consists of inner and outer tank. The inner tank is manufactured with 9 % nickel steel and the outer tank is composed of reinforced concrete and pre-stressed concrete. The 9 % nickel steel is widely used as a material for the inner tank since it has the strength and toughness enough for the cryogenic uses. The inner tank also has a function of preventing the LNG from leakage. Meanwhile, the concrete outer tank is designed to resist all the external loads including seismic load. Insulating materials are placed between the inner and outer tank to preserve the stored LNG.

The tank is of an above-ground, 9 % nickel steel full-containment type with a pre-stressed concrete outer tank and a corner protection system of 9 % nickel steel as a secondary barrier. The base of the tank has a bottom heating system (BHS) using ethylene glycol as brine fluid. The roof has a suspended ceiling deck and a steel-lined concrete dome.

First of all, different hydraulic calculations and process simulations from ship to tanks were carried out for determining the maximum elevation of the tank top. With the results of the calculations and simulations and the technical requirements above, the size of the tank to be developed was determined. Typical cross section of the full containment is shown in Fig. 3.10 indicating the different layers of the tank.

The LNG is stored in the 9 % nickel steel inner tank. The steel liner installed on the inside surface of the outer concrete tank provides the gas tightness. The boil-off rate is determined by the insulation system.

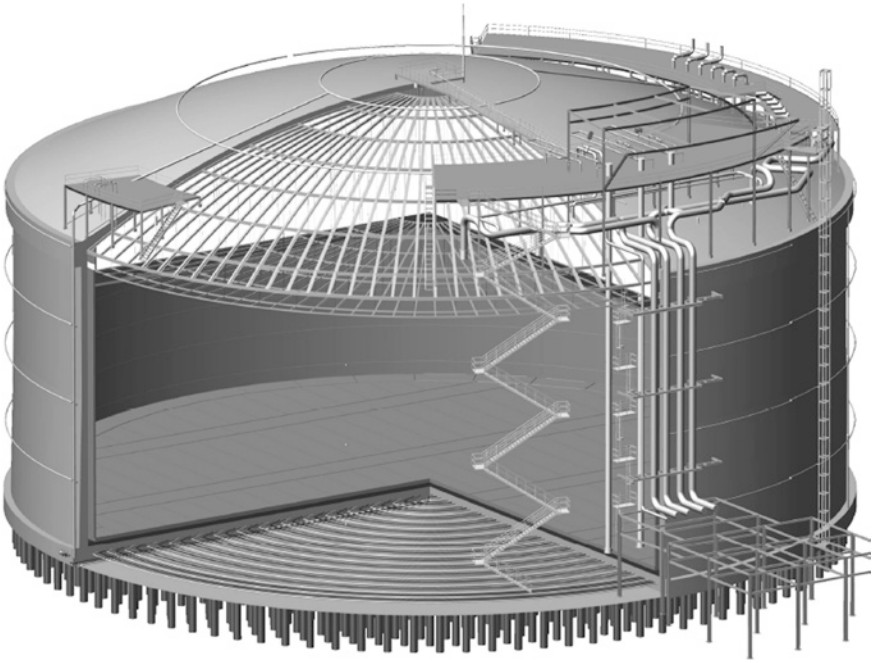


Fig. 3.9 Full-containment LNG storage tank [1]

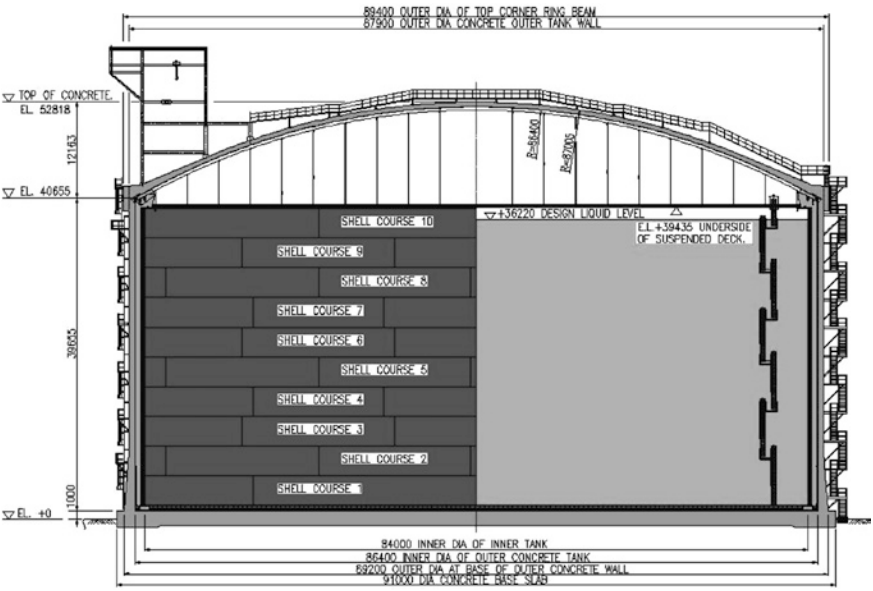


Fig. 3.10 Sectional view of the 200,000 m^3 full-containment LNG storage tank [1]

In case of an LNG leakage, liquid may impact on the outer tank. Accordingly, the liquid tightness must be guaranteed by the corner protection system as well as the polyurethane foam coating installed on the inside surface of the concrete wall. The concrete outer tank protects the inner tank in case of an emergency coming from the outside. The BHS is installed in order to avoid frost heave.

The roof liner consisting of a 5-mm-thick steel membrane stiffened with rafters in radial and tangential directions acts as formwork for the concrete sphere. The steel structure is fabricated on the bottom slab and lifted by air pressure to its final position. Rafters and roof liner plates are connected with a steel compression ring anchored in the concrete roof ring beam by welding.

3.5.3 Tank Design

The principal design codes applied to the design of the inner tank are API 620 and NFPA 59A. Figure 3.11 shows a general procedure for the design of 9 % nickel steel inner tank and the basic design data applied in the design.

3.5.3.1 Static Design

The inner tank is sized to contain 200,000 m^3 of LNG at a specific gravity of 0.48 in cold condition. An additional dead space for the in-tank pump NPSH of 2.0 m has been given at the bottom of the inner tank, and the allowance for seismic slosh plus free board has been made above the design liquid level to determine the height of the inner tank. Figure 3.11 shows stages involved in the design.

The thickness of 9 % nickel steel plates of the inner tank shell is designed:

- For the full head of LNG corresponding to the design liquid level
- For a partial hydrostatic test load corresponding to 1.25 times the full head of LNG.

The operating vapour pressure is equalised on both sides of the tank shell since the inner tank has an open top.

3.5.3.2 Seismic Design

The inner tank is designed as an unanchored and unstrained tank to resist the OBE and SSE level design accelerations given in basic design data.

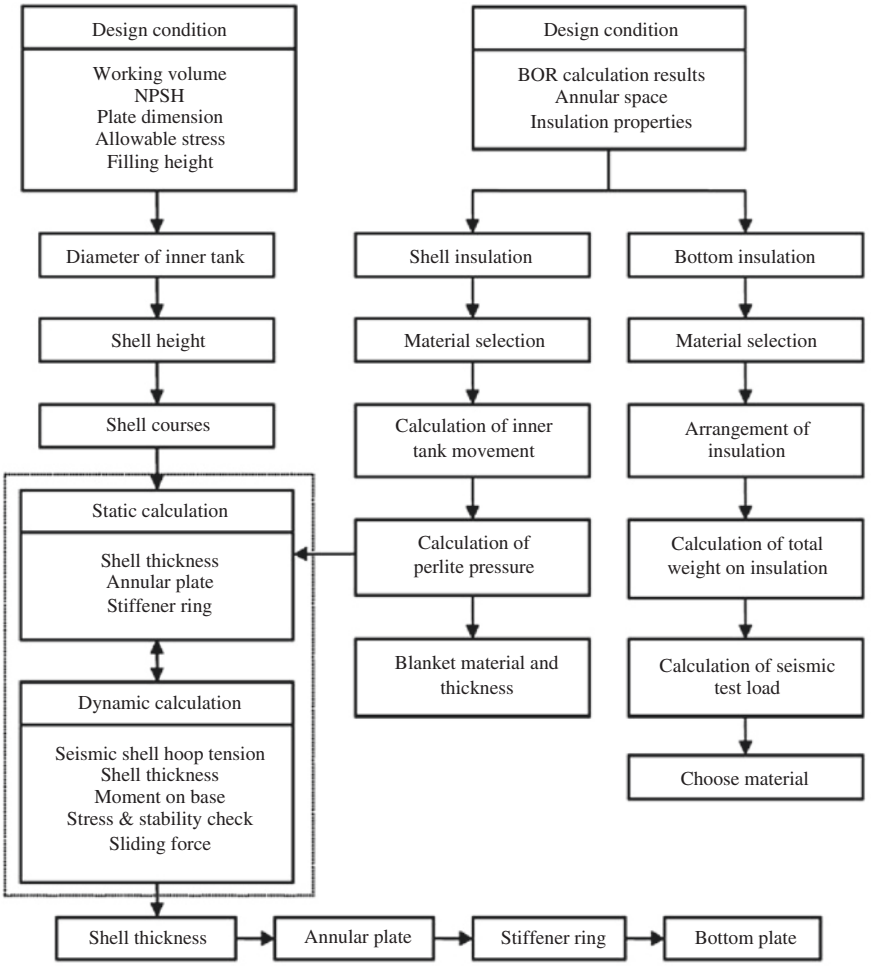


Fig. 3.11 General design procedure for 9 % nickel inner tank [1]

Hoop Loadings

The hoop forces on the inner tank shell are a combination of the static pressure due to the product liquid, the impulsive pressure due to the action of the impulsive component of the liquid, the convective pressure due to the action of the convective component of the liquid, and the vertical (or barrelling) pressure caused by the vertical component of the seismic loading.

Resistance to Overturning

The calculation of the resistance to overturning caused by contained liquid action on the annular plate and the width and thickness of the annular plate are based on the requirements of API 620.

Shell Compression
The shell compression is calculated using the procedure in API 620 for unanchored tanks. The maximum allowable shell compression is in accordance with API.

Seismic Sloshing
The seismic slosh height is calculated using the formula from API 620 L.8. An addition for shell run-up is made in accordance with the requirement of API 620 paragraph L.8.1.

Secondary Bottom
If the secondary bottom is located at the top of the base insulation, it will be almost at the same temperature as the inner tank bottom in service. In the event of an inner bottom leak, no thermal shock to the secondary bottom will be given.
If the secondary bottom is located at the middle of the base insulation, it will be at a higher temperature than the inner tank bottom in service. Leakage from the inner tank bottom will give a local thermal shock to the secondary bottom. More severe considerations than the code requirements for allowable stress and joint factors must be taken in this circumstance.

Corner Protection System
In case of a leakage from the inner tank, LNG may accumulate in the annular space between the inner and outer tanks and cool the wall/bottom corner. In order to prevent the LNG from cracking of the lower concrete wall section, a liquid tight protection system thermally isolated with cellular glass insulation and shielded with 9 % nickel steel plates will be provided as shown on Fig. 3.12. The design will be carried out using finite element analysis taking into account pressure loads and thermal stresses/movements.
The outer tank shall be capable of containing the full inner tank contents. Minor and major leak cases shall be checked by finite element analyses and combined with the maximum pressure in the analyses.

Suspended Deck
The suspended deck is designed to resist the loadings of self-weight, insulation weight, differential vapour pressure, construction traffic and seismic loadings. The design code applied in the design is British Standard Code of Practice CP 118.
An aluminium-suspended deck with concentric ring stiffeners is used, and the thickness of the deck plate is 5 mm. The deck plate is thermally isolated by glass fibre blankets of six layers as shown in Fig. 3.13 for the top corner arrangement.

Roof Sheeting
Roof sheeting design is in accordance with the rules of API 620 or RPLAS. The design pressure for the roof sheeting will be maximum air pressure for either air-lift or during concreting of the roof.

Roof Fittings
The reinforcement of the roof plates for the roof fittings is designed in accordance with the rules of API 620. If heat breaks are fitted, they shall be designed

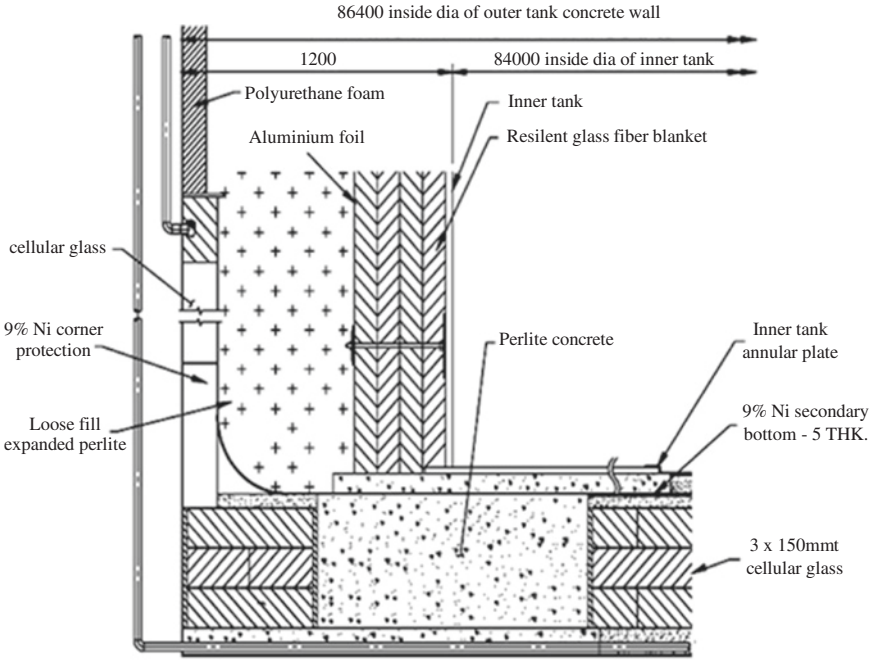


Fig. 3.12 Insulation and corner protection system [1]

in accordance with BS5500. The loadings to be considered include self-weight, product weight, seismic loadings, loadings arising from the action of seismically induced liquid motions and wind loading on associated external and internal pipe work.

3.5.3.3 Thermal Design

The thermal insulation system of the tank is designed to have a boil-off rate less than 0.05 % of the full tank contents per day due to atmospheric heat leakage. The heat leak calculation is based on a maximum ambient air temperature and the product design metal temperature of −170 °C. The concrete base slab temperature is assumed to be +10 °C due to the bottom heating, and solar radiation effect is taken into consideration.

3.5.3.4 Design of Outer Tank

The outer container consists of an entirely closed monolithic concrete structure—bottom slab, wall ring beam and roof. The concrete containment is designed for

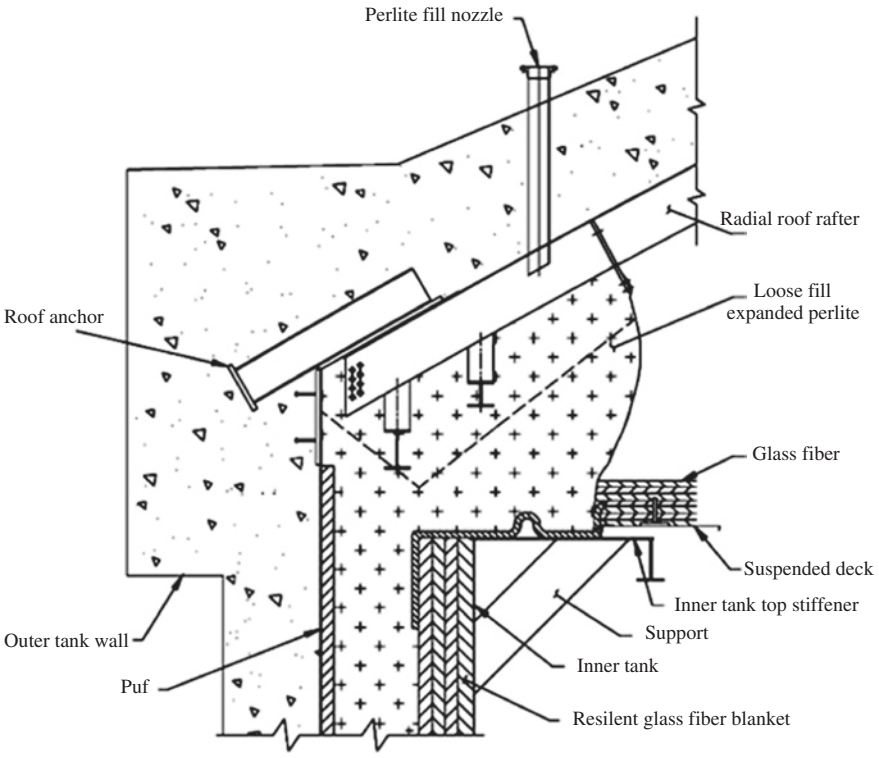


Fig. 3.13 Top corner arrangements [1]

all possible combinations of normal and emergency loads which may occur during construction, testing, commissioning, operating and maintenance of the tank. Design load combinations include the most severe combinations of loads. The general procedure for the design of outer tank is shown in Fig. 3.14, and the principal design code applied in this design is the British Standard BS8110: Structural use of concrete.

Design Criteria

The concrete containment is designed for the conditions of four stages: construction, operation, test and emergency as shown in Fig. 3.14. The concrete structure is designed on the ultimate limit state (ULS) and the serviceability limit state (SLS) for all normal situations and for the emergency situations of liquid spill in order to ensure containment for the liquid. For all other emergency situations, the concrete structure is designed on the ULS. The design was carried out according to the partial safety-factored design method of BS 8110. For the reasons of durability, the cracks at the concrete surface are limited.

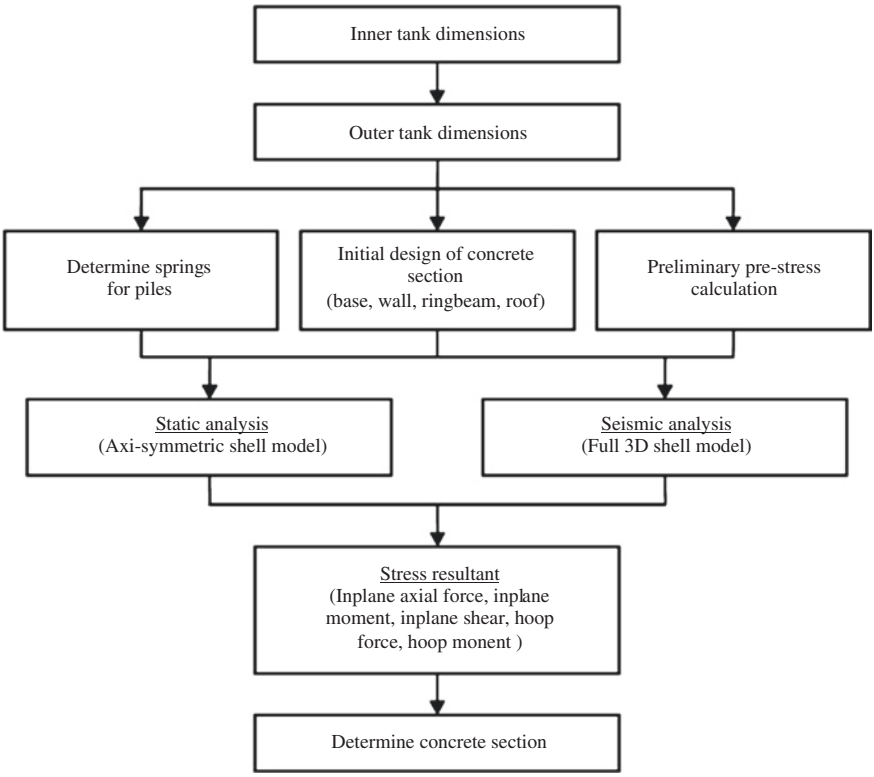


Fig. 3.14 Design procedure for concrete outer tank [1]

Loadings Considered in Design

The loadings to be considered in the design of outer tank are as follows:
For operation and test:

- Dead weight of the tank structure: concrete, inner steel tank, steel liners, suspended deck
- Pre-stressing including creep and shrinkage
- LNG filling inner tank
- Operating design gas pressure
- Hydrostatic test of inner tank
- Pneumatic test pressure
- Piping and installation on roof
- Live load on the dome
- Wind
- Snow
- Temperature operation
- Temperature difference slab–wall–roof

For emergency:

- Earthquake OBE
- Earthquake SSE
- LNG leakage
- Temperature leakage
- Increased overpressure
- Thermal radiation from an adjacent area
- Missile impact
- Cold spot

Static Analysis
For the static analysis of linear sectional forces, the main concrete tank structure is modelled using a semi-finite element computer program for axis-symmetric systems. The model comprises the bottom slab, the wall, the roof ring beam and roof. Spring elements were used to model the pile foundation in order to take into account the interaction between the tank structure and the subsoil.

Nonlinear Analysis
In emergency cases in which the concrete is exposed to extreme temperatures due to heat radiation, nonlinear effects of the concrete behaviour caused by the reduced stiffness due to crack formation are taken into account. The linear analysis where the individual single load cases can be superposed, a nonlinear analysis must be carried out for load combinations determined previously.
For nonlinear calculations, the reinforcement is an input data.

Dynamic Analysis
The seismic calculations for OBE and SSE were done in two stages:

- **Tank structure:** modal analysis of an axis-symmetric model using acceleration response spectra as excitation input. The whole system is required to remain in the elastic limit; no physical or geometrical nonlinearities shall occur.
- **Roof:** additional modal analysis of a three-dimensional finite element model (roof) using response spectra as excitation input.

Figure 3.15 shows an example of full seismic analysis model including the outer tank, internal structures and liquid.
Modelling of a burnout scenario involves 2D axisymmetric solid field elements and transient thermal analyses of the LNG tank outer walls. The tank roof and insulation layers (except any PUF layer) are assumed to have been destroyed and are not included in the analysis. Steady state conditions are initially applied for a specified time. To model the burnout situation, a temperature load of a specified peak temperature reducing to −170 °C over a distance of 1.5 m is moved down the inside of the tank at a constant speed for the burnout time under consideration. The constructed full containment is shown in Fig. 3.16 indicating the resulted structure.

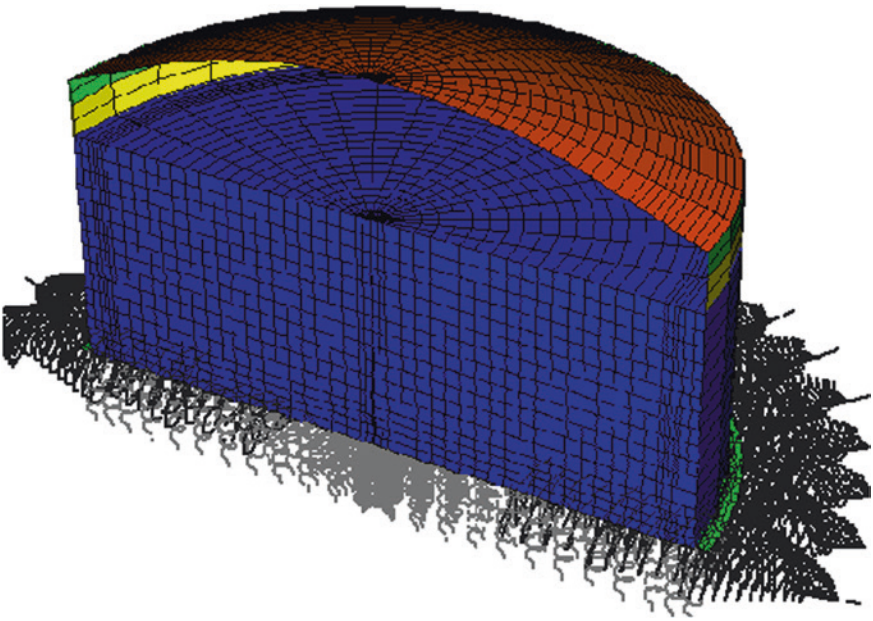


Fig. 3.15 Full model for 3D seismic analysis [1]

Relief Valve Heat Flux Modelling
With relief valve heat flux modelling, the tank bases are normally excluded from an analysis because they are considered to be remote from the heat flux loading. 3D solid field and continuum elements are used for semi-coupled transient thermal analyses of a segment of an LNG tank. An initial stress-free temperature is applied to all elements, and steady state conditions are established for an internal temperature of −170 °C and a specified mean annual external temperature. A heat flux is then be applied to a specified region on the top of the roof for the number of time steps under consideration.

Over The Top Piping and Tank Internal Pumps
Codes require that shell/bottom penetrations be avoided, so all piping has to be over the top of the tank.
Withdrawal is through in-tank pumps located in pump columns. Pump columns are provided with foot valves. In-tank pump systems require considerable tank top piping plus provision for removing the pump out of the tank for maintenance.

Anchors
Single-containment tanks need to be anchored to the foundation to resist uplifting loads due to tank internal pressure. This may not be necessary in



Fig. 3.16 200,000 m^3 full-containment LNG storage tank [1]

full-containment tanks, due to the weight of the concrete wall. This must be reviewed in high seismic locations.

Construction Sequences

These are shown in Figs. 3.17, 3.18, 3.19, 3.20, 3.21 and 3.22. The roof that has been built inside the tank is slowly lifted as shown in Fig. 3.23 using a shift in air pressure to raise the roof from floor to ceiling, when it reaches the top a team of welders secure the roof to the rim of the tank. The tank roof raising requires a focused effort to ensure that the work is completed safely. The wall profile is shown in Fig. 3.24.

Roof air raising and then welding to compression ring, erection of outer concrete wall liner. The roof that has been built inside the tank is slowly lifted using a shift in air pressure to raise the roof from floor to the ceiling, when it reaches the top, a team of welders secures the roof to the rim of the tank. The tank roof raising requires a focused effort to ensure that the work is completed safely and that the all the people involved in this important task go home safely at the end of the day. The constructed wall profile is shown in Fig. 3.24.

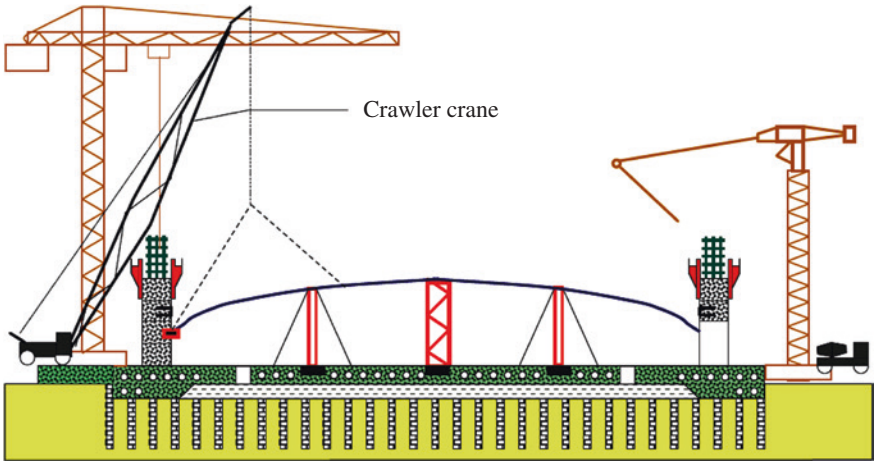


Fig. 3.17 Roof structure erection concrete wall construction in progress [1]



Fig. 3.18 Construction of base of the LNG storage tank in progress [1]



Fig. 3.19 Roof structure erection concrete wall construction in progress [1]



Fig. 3.20 Construction of LNG storage tank side wall in progress [1]

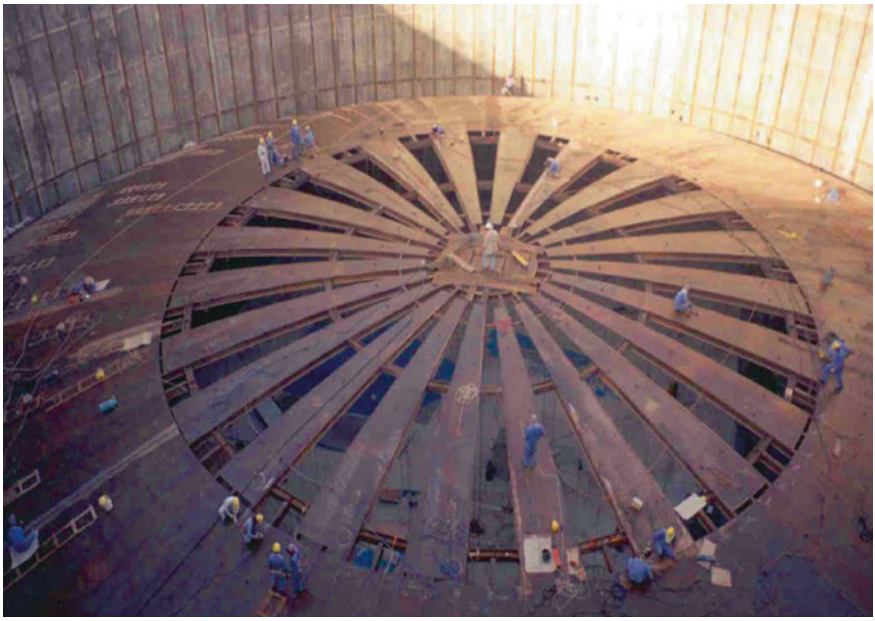


Fig. 3.21 Construction of roof of LNG storage tank in progress [1]



Fig. 3.22 Completion of roof construction [1]

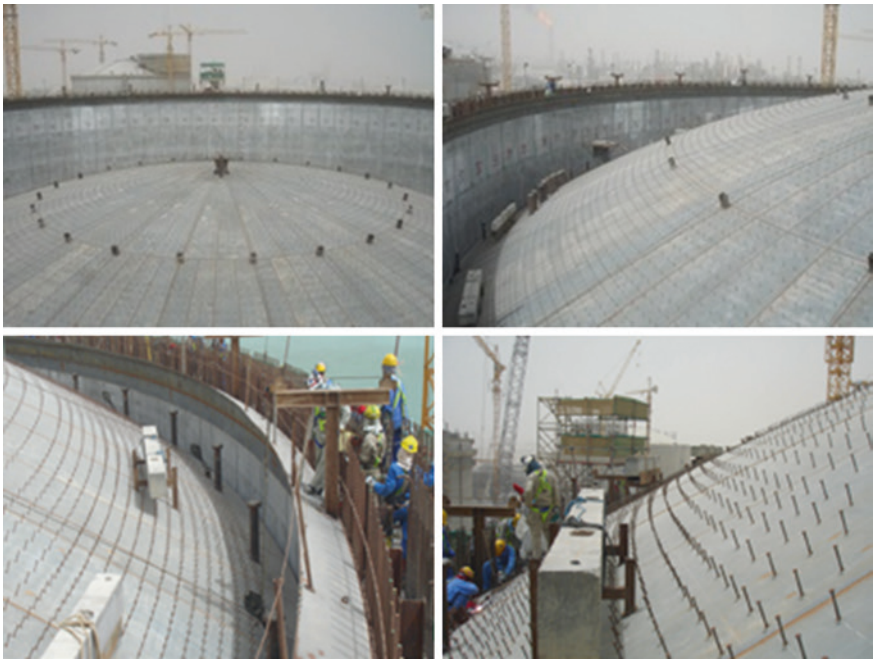


Fig. 3.23 Roof air raising in progress [1]



Fig. 3.24 Wall profile [1]

3.6 Regasification of LNG

The vaporisers at an LNG receiving facility transform LNG back into its gaseous state by warming the liquid so that the gas is at or above 5 °C (41 °F). There are several types of vaporisers in common use, but they all employ the same general principle—extracting heat from water to warm the LNG.

To return LNG to a gaseous state, it is fed into a regasification plant as shown in Fig. 3.25. On arrival at the receiving terminal in its liquid state, LNG is first pumped to a storage tank. When it is required for consumption, it is pumped at high pressure through various terminal components where it is warmed by passing it through pipes heated by direct-fired heaters, sea water or through pipes that are in heated water.

This illustration represents the process of taking liquefied natural gas (LNG) off of the ship as shown previously in Fig. 3.25, moving it into cold storage tanks and then regasifying it back into natural gas to send to market. The grey lines represent natural gas, and the yellow lines represent LNG. Naturally, there is a small amount of LNG which vaporises through contact with warmer air in the top of the ship containers and in the LNG storage tanks. This natural gas is syphoned out of the tanks and moved into the pipeline.

3.6.1 LNG Terminal Regasification Technology

The process of returning LNG to a gaseous state requires the introduction of heat energy.

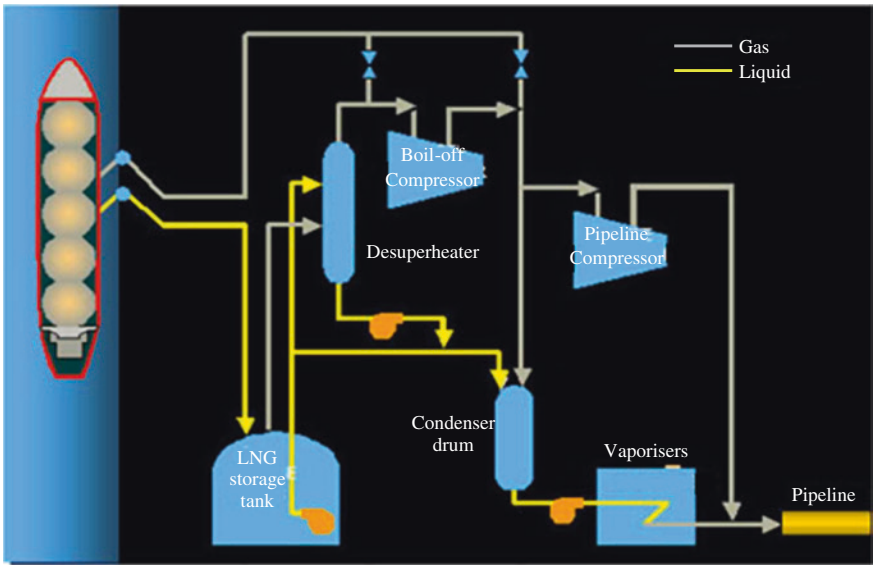


Fig. 3.25 LNG import terminal gas flow [1]

The various vaporisation technologies considered include the following:

- Open rack vaporisers (ORVs)
- Submerged combustion vaporisers (SCVs),
- Shell and tube heat exchanger with heat transfer fluid (HTF)
- Ambient air vaporisation systems, including:
- Heat Integrated Ambient Air Vaporiser (HIAAV)
- Direct Ambient Air Vaporisers (AAVs),
  - Direct Natural Draft Ambient Air Vaporiser
  - Direct Forced Draft Ambient Air Vaporiser
- Indirect Ambient Air Vaporisers (IAAVs)
  - Ambient Air Heat Exchanger with Heat Transfer Fluid (AAV-HTF)

The following describes each vaporisation system in greater detail.

3.6.1.1 Open Rack Vaporisers (ORV)

The ORV is commonly considered in the design of LNG import terminals as shown in Figs. 3.26, 3.27, 3.28 and 3.29. The relatively low mechanical, electrical and process complexity and reduced air emissions present good engineering arguments in its favour. However, life cycle operating costs must also be considered. The ORV uses sea water as the sole heat source to vaporise LNG. The vaporiser consists of a heat conductor panel with multiple tubes through which the LNG passes. A typical ORV arrangement is illustrated in Fig. 3.26.

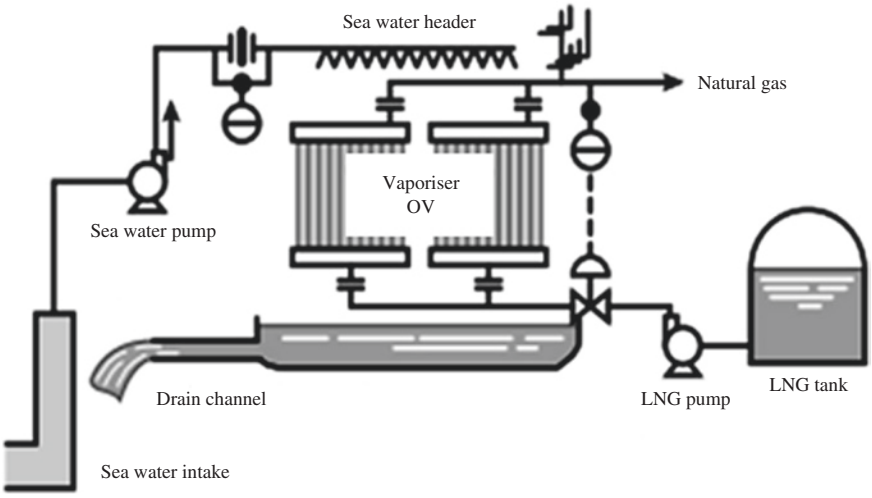


Fig. 3.26 Open rack vaporiser (ORV) [1]

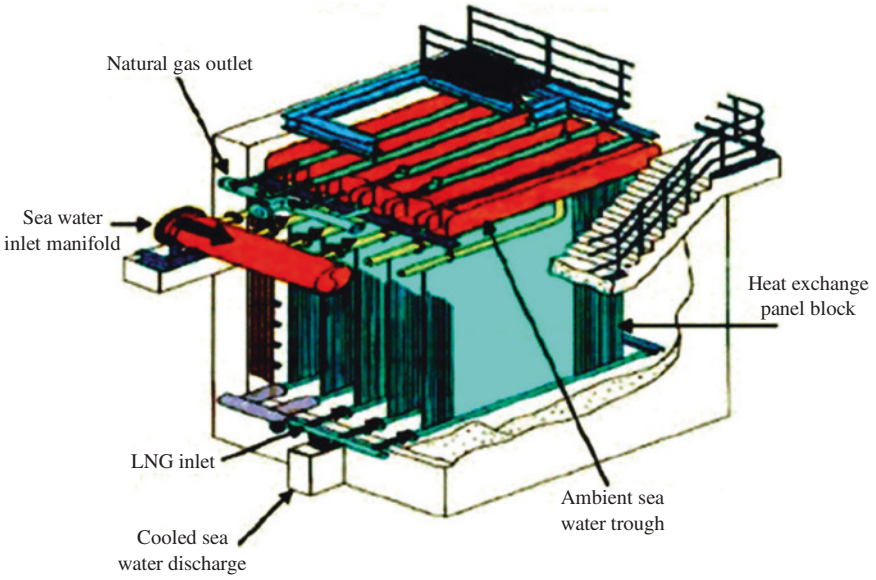


Fig. 3.27 A typical open rack vaporiser (ORV) arrangement [1]

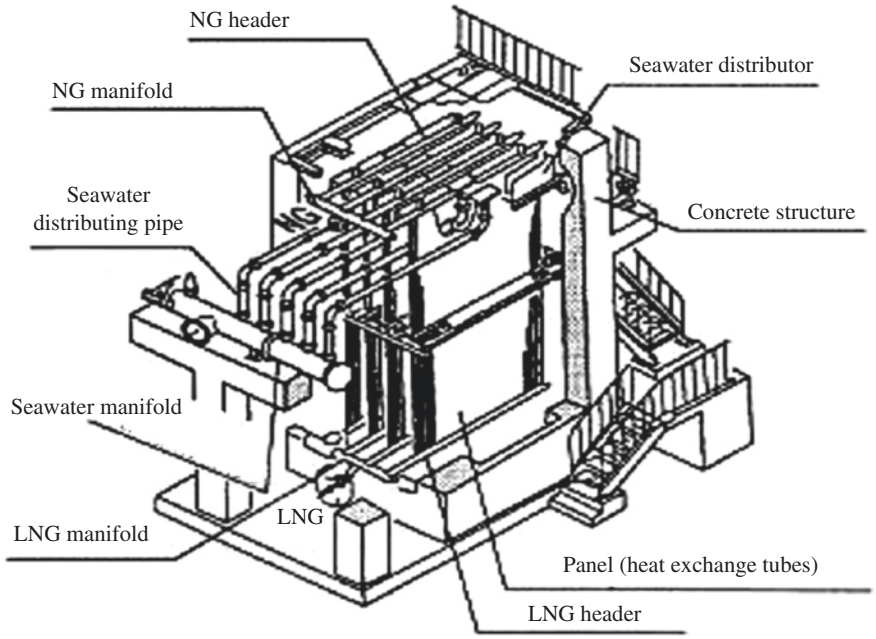


Fig. 3.28 A typical open rack vaporiser (ORV) arrangement [1]



Fig. 3.29 Typical open rack vaporisation arrangement [1]

LNG enters at the bottom of the vaporiser through a distribution header and moves up through the tubes while sea water flows down along the outer surface of the tube panels. Vaporised natural gas is removed from the top of the vaporiser and is sent to the distribution pipeline. The cooled sea water collects in a trough at the bottom of the vaporiser and is discharged to an outfall.

Chlorination of the sea water is used to prevent bio-fouling. Typically, sodium hypochlorite would be injected continuously to maintain a concentration of 0.2 ppm. In order to shock the system, elevated concentrations of 2.0 ppm would be injected for 20 min every 8 h, during ORV operation. De-chlorination of the effluent may also be required to meet environmental standards.

Further examples of open reach vapouriser arrangements and details are shown in Figs. 3.27, 3.28 and 3.29.

3.6.1.2 Submerged Combustion Vaporiser (SCV)

SCV systems are also commonly considered in the design of LNG import terminals as shown in Figs. 3.30 and 3.31. Study has shown the operational history, low capital cost, simplicity in design and operational flexibility combine to make this an attractive option. The SCV system uses natural gas as its heat source and requires electrical power to operate combustion air blowers and circulating water pumps. LNG is routed to a stainless steel tube bundle that is submerged in a water bath heated with flue gases generated by a submerged combustion burner. A schematic of typical SCV operation is presented in Fig. 3.30.

The forced air draft combustion burner is fuelled by low-pressure gas from either the boil-off gas (BOG) header or the natural gas send out pipeline. Hot

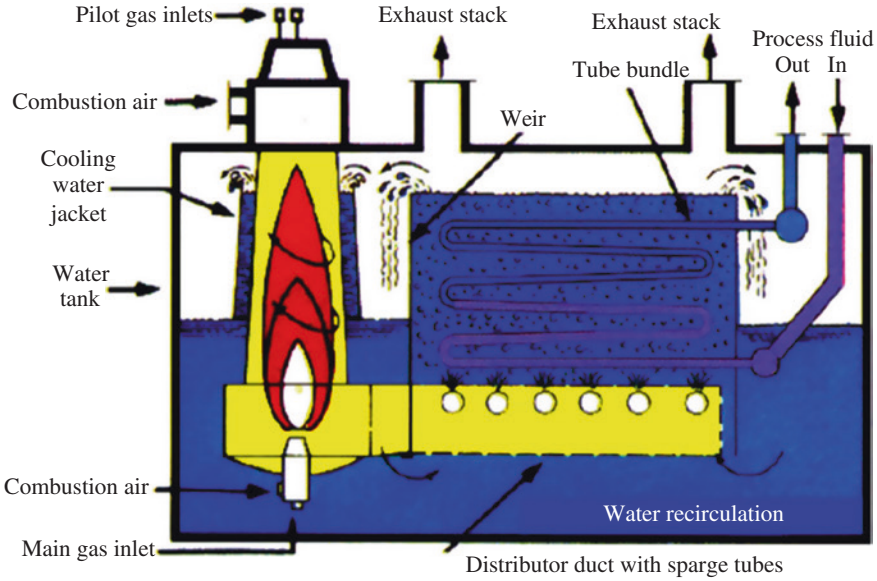


Fig. 3.30 Schematic of submerged combustion vaporiser [1]

exhaust gases from combustion are sparged into the water bath and create a relatively low-temperature (typically in the range of 12.8–32.2 °C) thermally stable heat source for the vaporisation of LNG flowing through the coil bundle. Natural gas exits the coils at pipeline pressure and temperature for pipeline distribution.

Since the SCV system does not use sea water, there is no impact on marine systems. However, since the heat source is natural gas, consideration has to be given to emissions to the atmosphere from combustion. The capital cost of an SCV system is low when compared to other vaporisation systems. Since the SCV system uses natural gas as its heat source, operating costs are comparatively high. This reduces the attractiveness of SCVs as a sole means of vaporisation at the terminal. Figure 3.31 shows a typical SCV installation.

3.6.1.3 Shell and Tube Vaporiser with HTF

There are many configurations of shell and tube vaporiser technologies that are available for LNG applications. One such system uses a closed-loop heated water–glycol system to provide heat to vaporise the LNG using U-tube heat exchangers and a superheater. This is shown typically in Fig. 3.32. The HTF is warmed from an external heat source, such as a fired heater.

Since these types of vaporisers can be stacked, this arrangement requires relatively small amounts of space when compared to other available vaporisation technologies. However, the inclusion of fired heaters requires additional space that can



Fig. 3.31 SCV installation [1]

negate the effects of any space saving advantages of this technology. The capital cost of the shell and tube arrangement with fired heaters is similar to that for SCV units.

3.6.1.4 Heat Integrated Ambient Air Vaporisation System with Waste Heat Recovery

This technology utilises the heat available in ambient air to vaporise LNG in a shell and tube heat exchanger as shown typically in Fig. 3.33. The design is generally such that the required outlet temperature of the natural gas cannot be achieved from ambient heating alone, and therefore, supplementary heating is required. Supplementary heating is usually obtained from the transfer of heat from the exhaust of power generation prime movers such as gas engines or turbines to the natural gas, using a recirculating quench water system. Provisions for additional heat for vaporisation, should it be needed, can be made by trim heating the natural gas through a supplementary duct burner. The trim heater can be a conventional shell and tube exchanger or a printed circuit heat exchanger (PCHE).

Exhaust gases from the power generation gas turbine/engine are cooled in the waste heat recovery unit prior to entering the quench column, from where they are released to the atmosphere at approximately 90 °F. The quench column condenses water vapour from the flue gas, which results in a small blowdown stream that

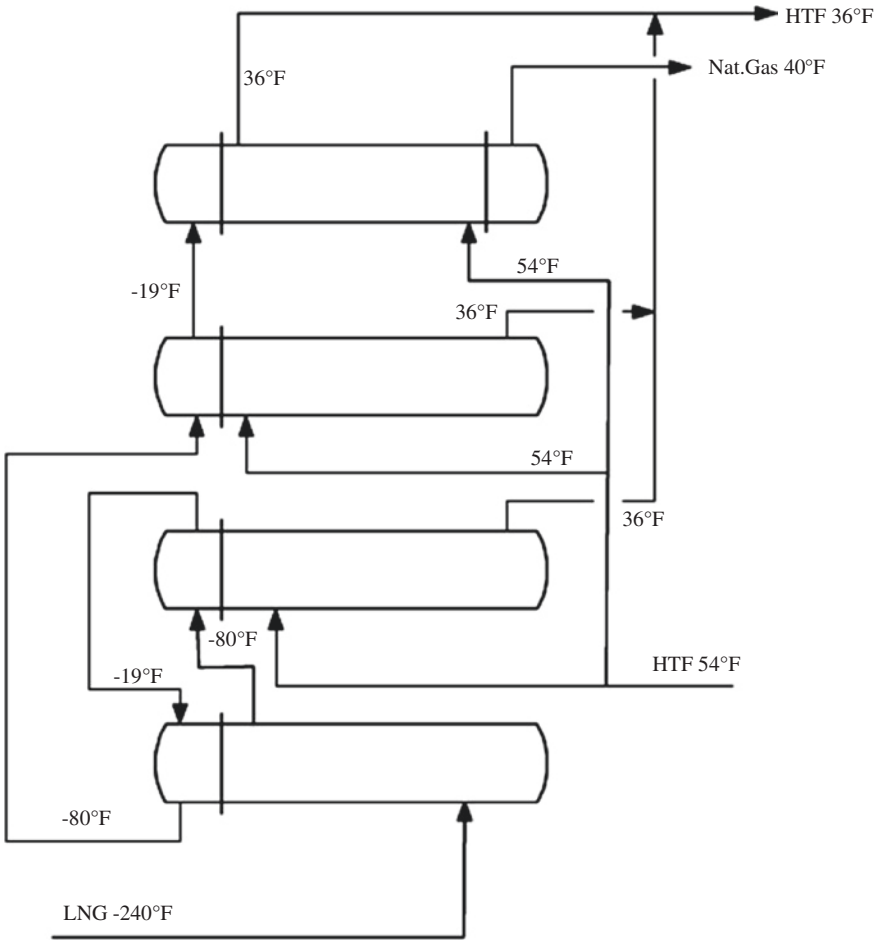


Fig. 3.32 Shell and tube vaporiser arrangement [1]

must be disposed of after treatment. Condensing the combustion products results in a high overall thermal efficiency of the system.

3.6.1.5 Direct Ambient Air Vaporisers (AAVs)

Direct AAVs transfer heat from the ambient air directly into the LNG through a heat exchanger heat transfer surface. In typical direct AAVs shown in Fig. 3.34, the LNG is passed through a manifold that divides the flow into a number of vaporiser units where a series of smaller flows are directed through individual heat transfer tubes. Each tube has aluminium fins for increased heat exchange area and is in direct contact with the ambient air.

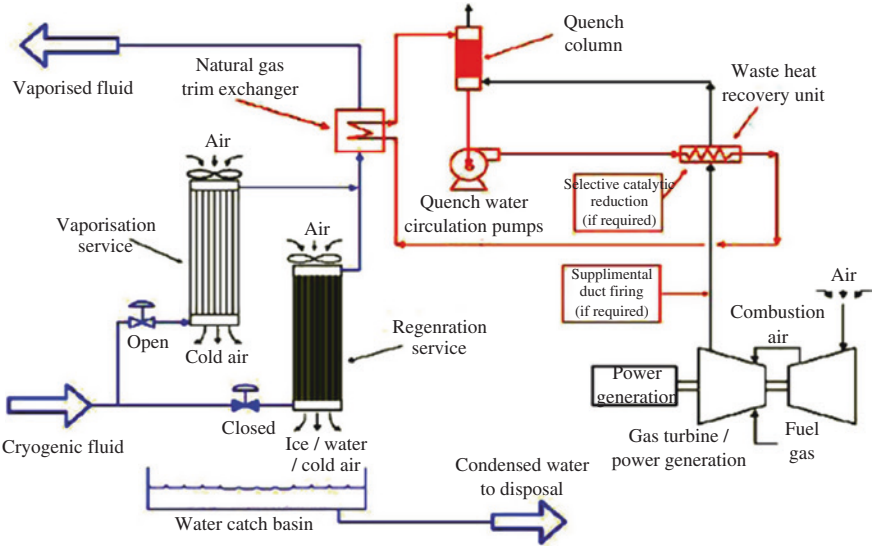


Fig. 3.33 Heat integrated ambient air vaporisation [1]

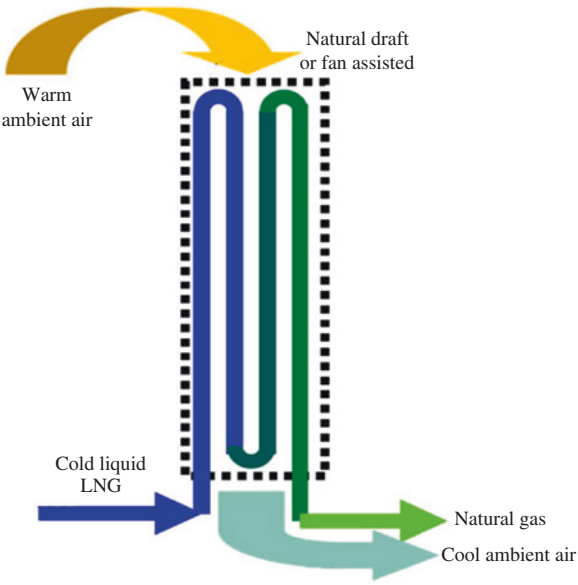


Fig. 3.34 Schematic of direct ambient air vaporiser [1]

Although there is no significant operational experience for direct AAVs in high-volume LNG applications, Direct AAVs have been used in liquid nitrogen service (a fluid colder than LNG) for over 50 years.

While Direct AAVs are in operation, frost and/or ice will build up on the units due to the proximity of the LNG to the ambient air. The longer a unit runs, the



Fig. 3.35 Direct natural draft ambient air vaporiser in operation [1]

more frost and/or ice builds, which gradually reduces the performance of the unit due to a reduction in the heat transfer surface. Hence, Direct AAV units need to be periodically shut down and de-iced.

1. **Direct Natural Draft Ambient Air Vaporisers**

   Direct Natural Draft AAVs shown in Fig. 3.35 rely on wind and natural convective currents to move air over the tubes and fins of the vaporiser unit. As warm air contacts the tubes containing LNG, the air cools and becomes dense, causing it to flow downwards to the bottom of the vaporiser unit. This causes warm ambient air from the surroundings to be drawn through the top of the unit.

   The outlet gas temperature of a Direct AAV is dependent not only on ambient conditions but also on other factors such as run-time and number of units in operation. Consider a single AAV unit, at the onset of operation, the temperature of the outlet stream is nearly the same as the ambient temperature. As the run-time increases, the unit begins to frost and/or ice up, thus reducing the amount of heat being transferred to the LNG. A significant amount of heat transfer is still achieved, but the temperature of the outlet stream reduces as time passes and ice accumulates.
2. **Direct Forced Draft Ambient Air Vaporisers**

   In Direct Forced Draft AAVs shown typically in Figs. 3.36 and 3.37, airflow into the unit is controlled by fans installed on top of the vaporiser. Each unit can be equipped with shrouds on each side to direct airflow through the vaporiser. Direct Forced Draft Vaporisers are approximately 1.7 times more effective than Natural Draft AAVs.



Fig. 3.36 Direct forced draft ambient air vaporisers without shrouds [1]



Fig. 3.37 Direct forced draft ambient air vaporisers with shrouds [1]

3.6.1.6 Indirect Ambient Air Heat Exchanger with Heat Transfer Fluid (AAVHTF)

This type of vaporisation system consists of shell and tube heat exchangers, fin-fan air heaters or reverse cooling towers and a heat transfer fluid (HTF)

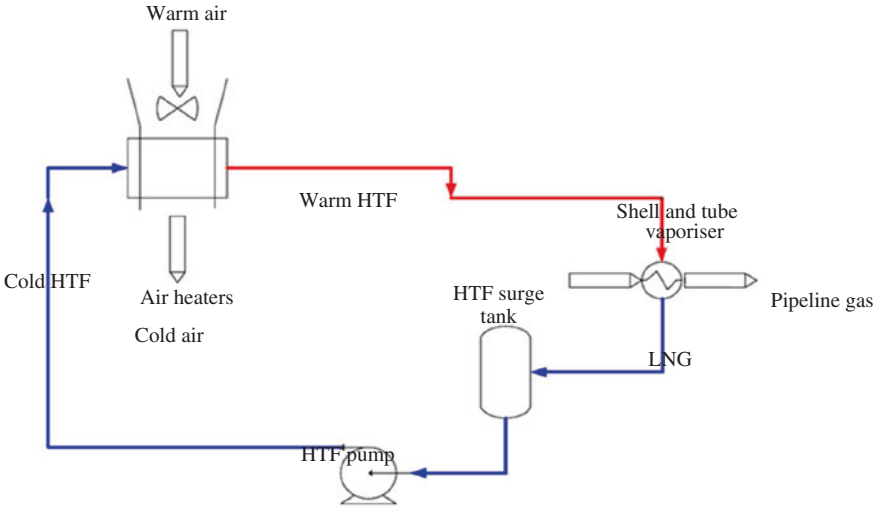


Fig. 3.38 Schematic of AAV-HTF system using air heat exchangers [1]

loop. Fin-fan air heaters are used to transfer heat from ambient air into the HTF, which is then sent to the LNG shell and tube vaporiser. The cooled HTF flows into a surge tank and is then pumped back to the air heaters as specified in Fig. 3.38.

The vaporised gas is then regulated for pressure and enters a pipeline system as natural gas. Once the LNG is regasified and leaves the regasification terminal, it ceases to be "LNG" and is indistinguishable from the conventional piped natural gas. Finally, residential and commercial consumers receive natural gas for daily use from local gas utilities or in the form of gas-powered electricity.

Although the main heat source for this vaporisation arrangement is the heat contained in the warm ambient air, supplemental heating is also required to maintain capacity and delivery temperature when the ambient air temperature drops below a nominal value.

3.6.1.7 Supplementary Heating

When the ambient temperature is not warm enough to raise the temperature of the vaporised LNG to 4.4 °C, supplementary heating must be provided to provide natural gas to the pipeline at the required 4.4 °C. The minimum amount of heat needed from the supplemental system has been determined to be 133 MMBtu/h. To be conservative, the supplemental vaporisation system will be sized for 180 MMBtu/h which provides 35 % extra capacity. Figure 3.39 shows the schematic of vaporisation system.

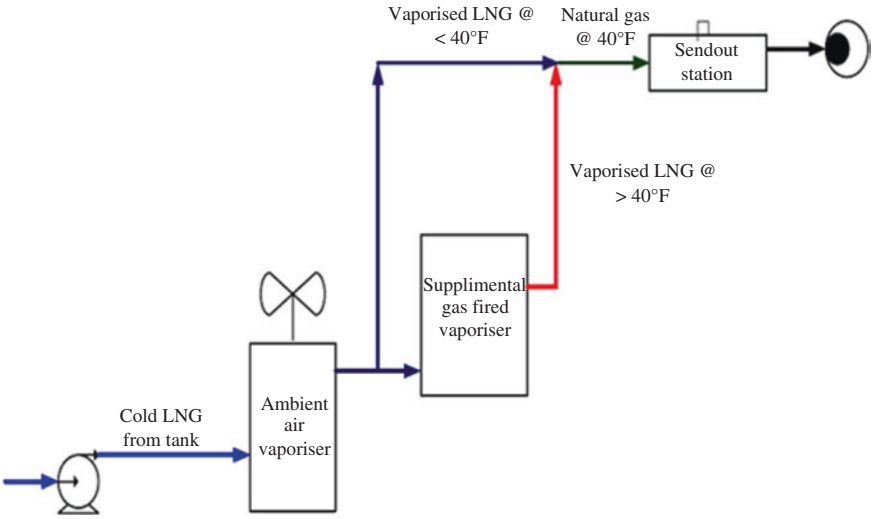


Fig. 3.39 Schematic of vaporisation system [1]

3.7 Safety on LNG Carriers

3.7.1 Hazards on LNG Ships

3.7.1.1 Cryogenic Hazard

Since liquefied gases are carried at very low temperatures, they pose a very serious threat in terms of cold burns, frost bite and damage to structures by making them brittle and susceptible to fracture. For temperature below −55°, fully killed fine grain carbon manganese steels are required. For temperature below 104°, high nickel content steels, stainless steel or aluminium may be used.

Due to the thermal stresses caused by changes in temperatures, the cryogenic property of LNG makes it necessary to carry out operations such as cooling down at a very regulated and controlled rate.

3.7.1.2 Flammability Hazard

LNG is a flammable substance. Methane vapour is easily ignited. Much effort is put into ship design to ensure effective cargo containment so as to limit vapour escaping to atmosphere and adjacent spaces. In addition, ships and terminals have design specifications for electrical equipment so as to ensure that, within well-defined operating zones, such sources of ignition are eliminated. Furthermore, in ship and terminal working environments, stringent operational procedures apply that limit other possible sources of ignition. Because of the very rapid vaporisation of spilled liquefied gases, the spread of flammable vapour will be far more

extensive than in the case of a similar spillage of oil. The chances of ignition following a spill of liquefied gas are, therefore, much greater.

The amount of vapour produced depends upon the volatility of the liquid.

Volatility can be defined as the tendency of a liquid to evaporate. Volatility of a liquid is measured using the following:

- Flashpoint
- Vapour pressure
- Boiling Temperature at atmospheric pressure.

3.7.1.3 Oxygen Deficiency Hazard

Oxygen deficiency in an enclosed space can occur due to any of the following conditions:

- When large quantities of cargo vapour are present.
- When large quantities of inert gas or nitrogen are present and
- Where rusting of internal tanks surfaces has taken place leading to oxygen depletion below 21 %

In the case of LNG cargoes, the presence of large quantities of LNG vapour would act as an asphyxiant and hence cause oxygen deficiency.

Normal content of oxygen in the atmosphere for survival is 21 %. Susceptibility of persons to reduced oxygen levels varies, but at levels below 19 %, impaired mobility and mental confusion rapidly occur. At levels below 16 %, unconsciousness takes place rapidly and may lead to permanent brain damage and will result in death.

3.7.1.4 Reactivity Hazard

Hydrate Formation: with moisture and water, the cargo can form ice and slushy material which can lead to clogging of pipes and impellers etc. Since CO_2 also forms dry ice at temperatures below $-70°$, this would also manifest as a powder in the tanks.

Self-Reaction: this risk does not exist in the case of LNG Reactions with other materials: for LNG, this hazard is not significant; however, due the varying composition of the gas, the specific material safety data sheets should be referred to.

3.7.1.5 Hazard Due to Low Flashpoint

The lowest temperature at which a liquid gives off sufficient vapour to form a flammable mixture with air near the surface of the liquid.

Flashpoints of some gases are given below in degrees celsius:

- Methane: −175 °C
- Propane: −105 °C
- Butane: −60 °C

3.7.1.6 Auto-Ignition Hazard

The auto-ignition temperature of a substance is the temperature to which its vapour-in-air mixture must be heated to ignite spontaneously. The auto-ignition temperature is not related to the vapour pressure or to the flashpoint of the substance, and since the most likely ignition sources are external flames or sparks, it is the flashpoint rather than the auto-ignition temperature which is used for the flammability classification of the hazardous materials. Nevertheless, when vapour escapes are considered in relation to the adjacent steam pipes or other hot surfaces, the auto-ignition temperature is worthy of note. Listed below is auto-ignition temperature of some gases:

- Methane: 595 °C
- Propane: 468 °C
- Butane: 365 °C

3.7.1.7 Rollover Hazard

This is more common in shore tanks and is primarily caused by the layering caused by density differences. When lighter fractions evaporate from the surface, the heavier fractions that remain form a dense layer at the top of the tank trapping the less dense layer below. As the cargo continues to warm, the layers at the bottom get less dense, and at a certain stage, this becomes unstable and the dense layer rolls over to the bottom causing the lighter fractions to rise to the top with a heavy release of vapour. This phenomenon is less likely to occur on ships due to the constant motion of the vessel which would not allow the layering process to occur.

To summarise the hazards that can be associated with LNG, ships would fall under one of the following categories.

- LNG—hazards associated are frost bite, asphyxiation and flammability
- Inert gas—toxic, skin irritation
- Nitrogen—asphyxiant

3.7.1.8 Pressure Hazard

Due to the high evaporation rate of liquefied natural gases, there is a tendency to produce very high pressures in closed containment systems. Hence, all cargo systems are provided with relief valve arrangements so as to mitigate the risk of damage to the containment systems.

Pressure surges also can cause significant damage to pipelines which are caused by static and dynamic forces. Hence, to deal with this risk, it is necessary to regulate the flow in manner so as to avoid the kinetic energy being converted into a strain energy that traverses the length of the pipeline at the speed of sound.

The size of the pressure surge is dependent upon the density of the liquid, the rate of deceleration and the length of the pipe. Pressure surges commonly can occur when filing a dead segment of a pipeline too fast, slamming of valves or even opening and closing a pipe segment to relieve pressure which would cause a sudden acceleration followed by a rapid deceleration.

3.8 First Aid Action

In the event of accidental leakage, or during maintenance tasks, personnel may be exposed to liquid or gaseous products being carried. The overall approach in the avoidance of hazards to personnel should always be prioritized as follows:

- Hazard removal
- Hazard control, and then only on
- Reliance on personal protection

3.8.1 Skin Contact

Since liquefied gases are carried at cryogenic temperatures, they constitute a significant hazard in terms of sustaining frost bite. If the skin is exposed to severe cold, the tissue becomes frozen. The symptoms of frostbite are extreme pain in the affected area (after thawing), confusion, agitation and possibly fainting. If the affected area is large, severe shock will develop initial symptoms are the skin initially becomes red, but then turns white

- The affected area is usually painless and
- The affected area is hard to touch if the area is left untreated, and the tissue will die and gangrene may occur.

Treatment

- Warm the area quickly by placing in warm water (~42 °C) until it is thawed.
- Keep the patient in a warm room.
- Do not massage the affected area
- Severe pain may occur on thawing: give pain killer or morphine if serious
- Blisters should never be cut or clothing removed if it is adhering firmly
- Dress the area with sterile dry gauze
- If the area does not regain normal colour and sensation, obtain medical advice.

3.8.2 Inhalation

The hazard with LNG inhalation is asphyxiation as the LNG vapour can result in the deprivation of oxygen supply to the blood. Although no toxic properties are present, a condition of asphyxia can lead to unconsciousness and can be fatal if the condition is not reversed.

Some cargoes are toxic because their chemical composition/properties can result in temporary or permanent health hazards causing symptoms such as irritation, tissue damage or impairment of faculties.

Other gases may effect the nervous system (state of anaesthesia). The unconscious person may react to sensory stimuli, but can only be roused with great difficulty.

3.8.3 Ingestion

Cases of ingestion on board will be very rare. MFAG and MSDS for the concerned cargo to be referred for the exact procedure to be followed.

3.8.4 LNG Fire Fighting Techniques and Equipment

Fire fighting on-board ships is an inherently hazardous and dangerous operation. The effects of smoke and heat are magnified by the steel structure which confines and controls the direction of the spreading fire. The steel structure, with ventilation shafts and multiple openings, also gives a fire an opportunity to spread in a three-dimensional fashion which requires ship's staff to have an intimate knowledge of both the ship's structure and fire fighting techniques and procedures.

In the case of a fire, it is vital that the response time is minimised as specifically in the case of LNG, spillage would result in the formation of a large flammable vapour cloud which could spread to other areas/sources of ignition.

Any LNG spillage would immediately start to boil-off as mentioned above. If the escaping liquid has a temperature below ambient, then the gas will form a visible white cloud since it condenses the water vapour in the atmosphere. Should ignition take place the cloud will be rapidly consumed and flash back to the area above the leak. The rate of vaporisation will be increased by the heat generation.

The most effective way of stopping such fire is by cutting off the source and extinguishing the fire by starvation. The emergency shutdown system should be activated immediately and as many cargo system valves as possible closed. This will isolate most of the cargo system. In taking this action, it is important to know that suitable arrangements should be made to relieve the pressure in the pipelines as warmed up LNG inside these lines may expand and cause further leaks and damage to the system.

3.8.4.1 Use of Dry Powder

The chemical reaction of the small particles of the extinguishing material when exposed to flame extinguishes the flame. Various types of powders are marketed; potassium-based powders are more effective than the sodium-based or multi-purpose powders.

When using the dry powder, the following points are worth noting:

- Because of the danger of reignition, dry powder should not normally be used unless the source of gas can be isolated.
- Maximum possible rate of application is desirable and as many high velocity jets as are available should be brought to bear at once so that reserves are not wasted and successful extinguishment obtained.
- Coordinated simultaneous attacks with all available applicators, if possible from windward, will have maximum effect.
- Jets should be aimed with the objective of reducing boil-off rate by sweeping the whole fire area, but on no account must the pool be agitated.

The powder jets should be slightly depressed below the horizontal sweeping rapidly back and forth over the entire fire area. Direct impact of the powder jets on the liquid pools should be avoided. The application should be such that the entire surface area of the burning pool is covered. LNG ships are provided with dry powder hose stations along the cargo decks. The hoses are 15–20 m long and are located along the cargo deck so that they can cover the cargo tanks, manifolds and the machinery spaces. Figure 3.40 summarises the several steps taken to contain such incidents.

Process Isolation and Draining

Isolating valves and pipelines also require careful consideration as trapped LNG in a pipeline would expand rapidly when heated and can result in ruptures in the pipelines causing new leaks and extensive damage. This could cause a greater spread of the fire. Hence, when isolating, it is essential to drain the contents back into the tank or safely depressurize the system. Relief via the vent masts is provided with nitrogen snuffing systems while draining back to tanks would have to be done with great care to prevent flash back to the source.

Water Spray Protection

Water spray systems are designed to apply water at a predetermined application rate to protect specific equipment or areas. Water is cheap and at sea is readily available. It is non-toxic and does not give rise to toxic products. It has high heat absorption capability in terms of its specific heat and latent heat of vaporisation. When water turns to steam, the volume increase is in the ratio of 1:1,700.

Steam forms a protective cover around the fire cutting off the supply of air, so that the combustion cannot be supported. Water spray system can be used to protect tanks, ship structure and fire fighters from radiated heat. Fixed water curtains and spray systems should be backed up by portable spray hoses. Fire fighters should

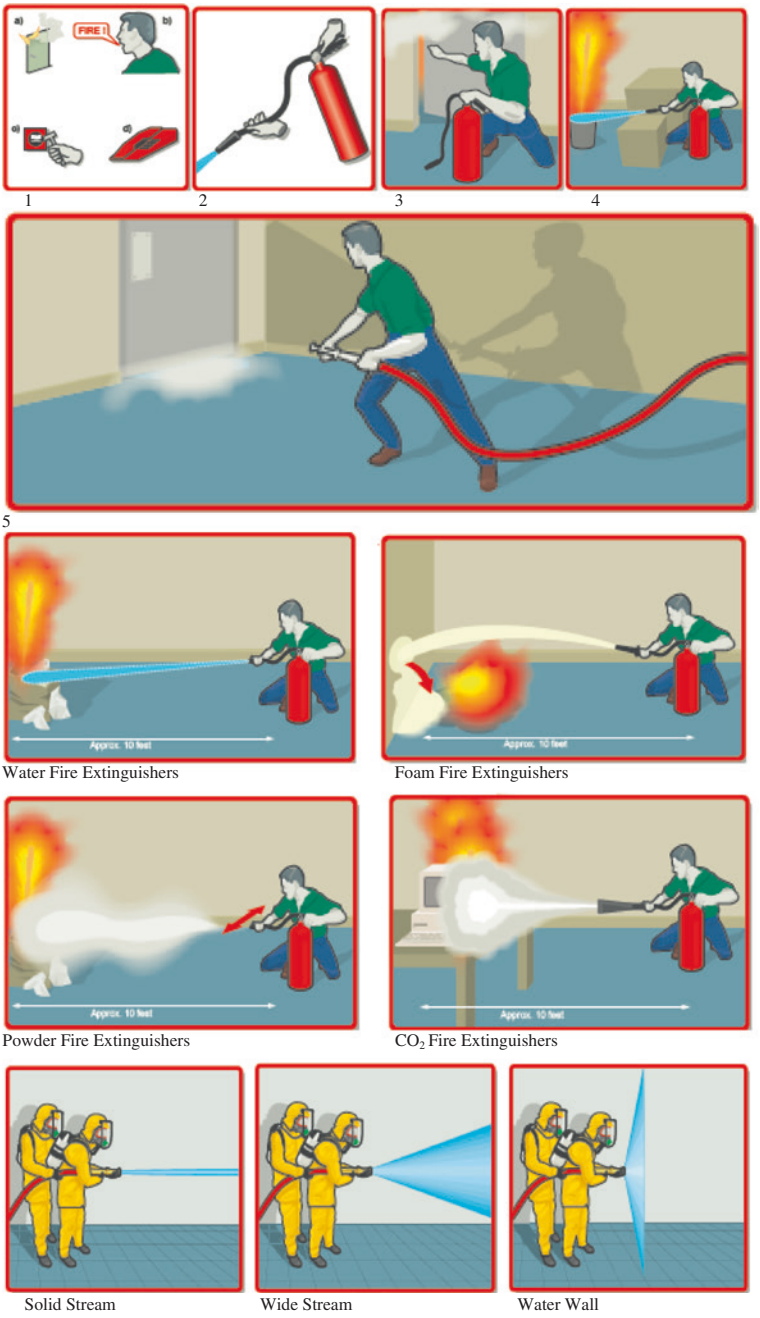


Fig. 3.40 Ideal use of water hoses [1]

approach from windward protected from possible flash burns by water sprays and/or protective clothing.
Water can also be used to prevent contact of liquid LNG with deck and metallic structures thereby reducing the chances of brittle fracture.

Use of High-ex Foam (Limited Use Only in Machinery Spaces)
Application of foam would result in forming a barrier which would prevent the radiation from the flame to the liquid below and will thereby reduce the vaporisation rate and consequently the extent and radiation of the fire. Foam should be continuously applied and a depth of 1–2 m should be achieved. High expansion foam of about 500 to 1 expansion ratio has been found to be the most rapidly effective for this purpose. Portable foam tanks would generally be found in the E/R and other machinery spaces; however, their use in the cargo spaces is limited.

Action to be Taken in the Event of LNG Vapour Release
In the event of unintentional vapour release, the following actions should be taken:
Spaces and areas where leakages or spillages have occurred should be evacuated upwind immediately. Take care as flames may be invisible. Leaking gas may be extremely cold.
Measures should be taken to prevent leaking gases from penetrating into any other part of the ship. Bear in mind that some gas are heavier than air or may otherwise accumulate in lower or non-ventilated parts of the ship. Particular attention should be taken in order to prevent gases drifting into occupied areas of the ship, e.g. living quarters, machinery spaces and working areas.
Wear protective clothing suitable for gas protection and self-contained breathing apparatus.
Avoid all sources of ignition (e.g. naked lights, unprotected light bulbs, electric hand tools, friction). Wear non-sparking footwear.
Even short inhalation of small quantities of gas can cause breathing difficulties.
Keep clear of evolving gases. Avoid all skin contact.
Let spilt liquefied gas evaporate. When in contact with cold liquefied gases, most materials become brittle and are likely to break without warning. Avoid all contact even when wearing protective clothing. If practicable, protect ships superstructure with copious quantities of water. Do not direct water jet into the centre of the spill.

Spillage On Deck
Small spillage: Let gas dissipate. Keep clear.
Large spillage: Let gas dissipate. Keep bridge and living quarters upwind. Otherwise, protect crew and living quarters against flammable or toxic gases using water spray to drive gases away (water curtain). Spilt liquefied gas: Use water jets from as far as practicable to accelerate evaporation, not directing them straight onto the spill.

Spillage Under Deck
Small Spillage: Do not enter space. Provide adequate ventilation. Where a ventilation system is used, particular attention should be taken in order to prevent gas penetrating into other areas of the ship. Let gas evaporate. Keep clear.

Check atmosphere before entering (toxicity and explosion hazard). Do not enter space without self-contained breathing apparatus.
Large spillage: Do not enter space. Provide adequate ventilation. Where a ventilation system is used, particular attention should be taken in order to prevent gas drifting into other areas of the ship. Bridge and living quarters must be kept upwind. Otherwise, protect crew and living quarters against flammable or toxic gases using water spray to drive gases away (water curtain). If practicable, use water spray to avoid ignition of flammable gases in the space. Check atmosphere before entering (toxicity and explosion hazard). Do not enter without self-contained breathing apparatus.

LNG Jettison
There are occasions where it is necessary to jettison cargo from a cargo tank.
A containment or insulation failure in one or more of the cargo tanks may necessitate the jettisoning of cargo from that particular cargo tank to the sea.
Jettisoning of cargo from a particular cargo tank to the sea is carried out using a single main cargo pump, discharging LNG through a portable nozzle fitted at the ship's manifold. The jettison pipe is stored on the deck forward of the compressor house.
As jettisoning of LNG will create hazardous conditions:

- All relevant fire fighting equipment must be manned and maintained in a state of readiness during the entire operation.
- All accommodation vents, other openings and all vent fans must be secured with the exception of that ventilation to the engine room necessary to maintain it in an operational status.
- The NO SMOKING rule must be rigidly enforced.
- The fixed installation water curtain is to be placed in operation on the side of the jettison to protect the ship's structure.

The portable nozzle is approximately 6.5 m long and when connected extends 1.5 m from the ship's side. Weather conditions, and the heading and speed of the vessel relative to the wind, must be considered so that the jettisoned liquid and resultant vapour cloud will be carried away from the vessel. In addition, if possible, avoid blanketing the vapour with exhaust gases from the funnel as shown in Fig. 3.41. The discharge rate must be limited to the capacity of one cargo pump only and, if necessary, reduced to allow acceptable dispersal within the limits of the prevailing weather conditions.

Too rapid a flow of LNG will result in rapid phase transfer (RPT) when the liquid hits the sea water. Violent increase in gas vapour as liquid hits the sea water, producing a very cold cloud of vapour which does not dissipate readily from the immediate vicinity of the vessel.

Vent Mast on Fire
It is likely that occasionally one of the four LNG vent masts will light-off when the ship is travelling through a thunder and lightning storm.

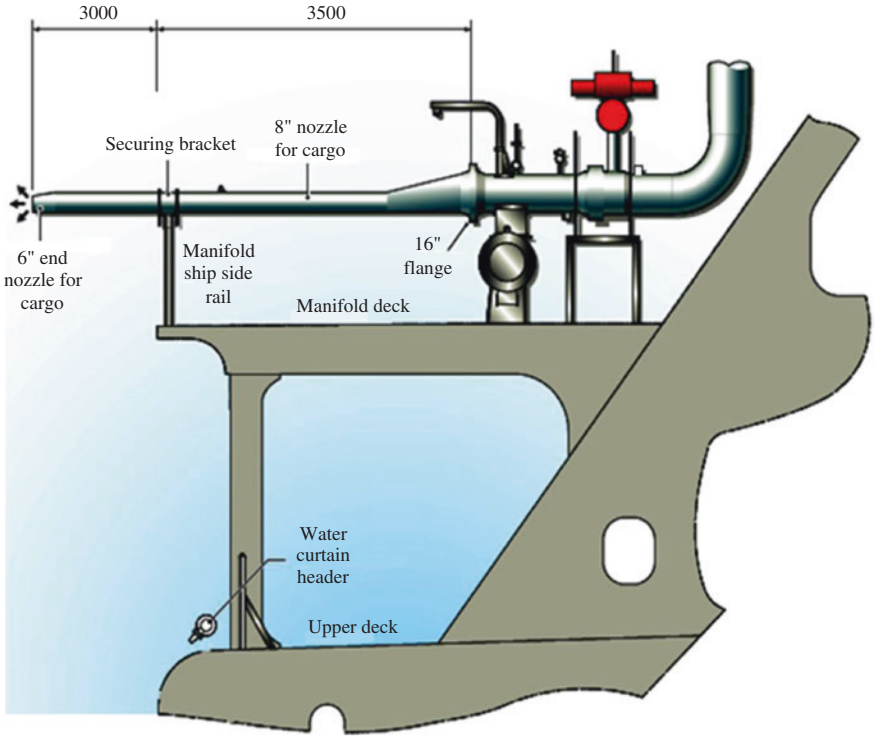


Fig. 3.41 LNG jettison [1]

The possible causes of the fire are as follows:

- Cargo tank pressure above 19kPa can result in one or more of the cargo tank safety valves temporarily lifting off their seats due to the surging motion of the ship in extremely rough sea conditions
- Leaking vapour header forward vent control valve CG702
- Leaking cargo tank safety valve
- Manually controlled venting through control valve CG702

The fire should be extinguished, but there is not any real danger to the ship or its personnel while the fire is burning because all of the LNG vent masts incorporate the following safety features:

- Hood is located 12 m above the trunk deck
- Hood and vent mast are 316 grade stainless steel
- Hood is fitted with a regulatory body approved flame screen to prevent any flame from passing across the screen and down into the vent mast
- A nitrogen fire smothering valve is piped to each LNG vent mast

Corrective Action to Extinguish the Fire on the Top of the LNG Vent Mast
If carrying out controlled venting from No.1 LNG vent mast:

1. Stop venting by closing the control valve CG702.
2. Manually open the nitrogen fire smothering valve piped to No.1 LNG vent mast.

If non-controlled venting:

1. Manually open the nitrogen fire smothering valve piped to the LNG vent mast that has the flame.
2. If that does not extinguish, the flame and other means of reducing the cargo tank pressure fail:

- Insert a 1-mm-thick blank flange into the outlet flange of the cargo tank relief valve that is suspected of leaking and reopen the nitrogen fire smothering valve.
- If that does not extinguish the fire, remove the blank flange from the first tank relief valve and place it in the outlet flange of the other tank relief valve protecting the tank.

Reopen the fire smothering valve on the vent mast. If that extinguishes the flame, leave the fire smothering valve partially open to that mast and remove the 1-mm-thick blank from the outlet flange on the cargo relief valve.

Boil-Off Rate Measurement Purpose
The purpose of this note is to define the method of boil-off rate (BOR) measurement on-board G.T.T. membrane-type LNG carriers designed for QGTC and having a total cargo capacity of 266,000 m^3 each.
The design daily BOR of cargo during laden voyage shall be less than 0.135 % of the full loaded cargo when assuming the following conditions:

- Ambient air temperature = +45 °C
- Sea water temperature = +32 °C
- Cofferdams temperature maintained at +5 °C
- Cargo considered as pure methane (density 425 kg/m^3, latent heat of vaporisation 511 kJ/kg)
- Cargo piping empty.

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2. http://www.worldoil.com/February-2006-Process-selection-is-critical-to-onshore-LNG-economics.html (05/03/2013).

Chapter 4
Gas Flow and Network Analysis

4.1 Gas Flow in Circular Pipes

4.1.1 Introduction

Since natural gas almost always flows through circular pipes, it is thus important to understand how the gas behaves in the pipes. What are the flow parameters and the properties of the flowing fluids and what are the gas flow governing equations in distribution and transmission systems? For gas flow in pipes, the parameters such as velocity, pressure and density, which describe the behaviour and state of a fluid, are not constant. They vary from one point in the pipe to another and from one instant in time to another. The concepts of steady and *unsteady* state flow in the context of some fundamental flow parameters should assist in better understanding of how gas behaves in practice. In order to simplify many of the practical problems encountered in the gas industry, we make a distinction between flow where the parameters vary with time (*unsteady* or *transient flow*) and flow that can be considered steady (although strictly speaking no gas flow is truly *steady*).

Transient flow is dealt with later in this module; for the time being we will only consider steady state

Flow Not surprisingly, steady state flow is much simpler to analyse than *transient* flow, and fortunately, it can be applied widely in gas transportation, particularly for low-pressure distribution.

Steady state flow is defined as that in which the various parameters at any point do not change with time. Uniform flow is defined as that in which the various parameters at any particular instant do not change with position.

Often, we can assume steady state uniform flow, for example flow at a constant rate through a long straight pipe of constant cross section. However, the assumption of uniform flow is not valid near to the solid boundaries along which the fluid is flowing. This is because all fluids have *viscosity*, (a property that determines the

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resistance to flow of a particular fluid), which reduces the relative velocity to zero at the pipe wall. Gases have very low viscosities compared to liquids and will flow more easily through the pipes.

The interrelationship between the properties of the pipe wall and the fluid viscosity is very important in determining the overall resistance to flow. Also, important is the relationship between the properties.

4.1.2 Pressure Drop Along the Pipeline

The flow capacity of the pipe is directly related to the pipe diameter since the pipe diameter determines the cross-sectional area available for fluid flow. Figure 4.1 shows the general profile of pressure along the length of a pipeline at different flow rates in which for zero flow, there is no pressure loss so the outlet pressure is the same as the inlet pressure [profile (1)]. The highest flow rate produces the greatest pressure loss giving the lowest outlet pressure Pout (4) [profile (4)] and the lowest flow rate [profile (2)], results in a higher outlet pressure Pout (2).

The parameters that affect the pressure drop of gas flowing between two points in a pipe are (i) rate of gas flow, (ii) internal pipe diameter, (iii) pipe length gas temperature, (iv) gas properties (i.e. (iv) viscosity, density), (v) properties of the pipe wall and (vi) initial pressure of the gas. We may list most of these parameters, but possibly not the initial pressure of the gas. Unlike liquid (incompressible) flow, gas flow is affected by the compressible nature of the gas, as represented by the equation of state:

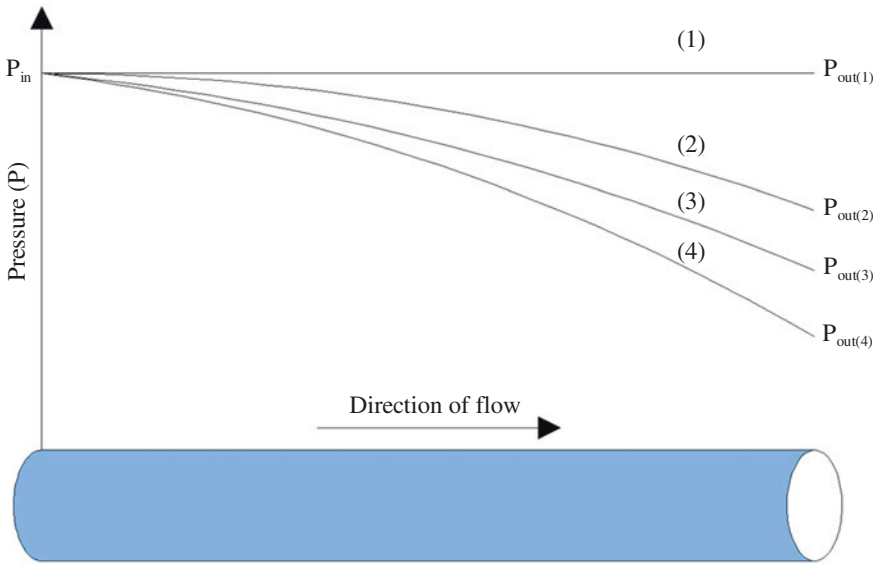


Fig. 4.1 Schematic diagram of pressure drop along a pipeline

$$PV = ZmRT \tag{4.1}$$

where P is the gas gauge pressure (N/m^2), V is the volume of gas (m^3), Z is the compressibility factor of gas, m is mass of the gas (kg), R is the universal gas constant (J/m^2 K) and T is the gas temperature (K).

4.1.3 Properties of Flowing Fluid

4.1.3.1 Viscosity: Dynamic and Kinematic

All fluids will resist any force tending to cause one layer to move relative to another layer. However, this resistance is offered only while the movement is taking place. As a result, as soon as the external force is removed, the flow subsides due to the resisting forces. There is no tendency for the fluid particles to revert to their original positions when the fluid motion ceases. This resistance to movement of one fluid layer adjacent to another is said to be due to the viscosity of the fluid. The forces resisting the relative movement of fluid particles are called *viscous forces* and are dependent not only on external forces, but on the nature of the fluid itself.

"Thick" fluids such as treacle or tar are said to be very viscous; they do not flow very easily because the forces resisting relative movement between the fluid particles are strong. "Thin" fluids such as water flow more readily and gases even more so. If the *viscous forces* resisting the overall motion of a fluid exist between layers of individual fluid particles, then these layers must be moving at different velocities. The force opposing the motion of the fluid is the result of a stress being set up within the fluid between each layer, where the stress (T) is equal to the force (F) divided by the area over which the force acts (A). For the straight and parallel motion of a given fluid, the tangential stress between two adjacent layers is proportional to the velocity gradient ($\delta u/\delta y$) in a direction perpendicular to the layers. Thus:

$$\tau = \frac{F}{A} \propto \frac{\delta u}{\delta y} \tag{4.2}$$

Assuming that the velocity varies only perpendicularly to the flow, then in the limit as δy tends to zero then:

$$F = \mu A \frac{\mathrm{d}u}{\mathrm{d}y} \tag{4.3}$$

The constant of proportionality μ is called the coefficient of dynamic (absolute) viscosity of the fluid.

The value of μ is a function of the temperature of the fluid. In the case of a gas, μ is a function of pressure, although only at extremely high or low pressures (both of which are outside the normal pipeline operating pressures). The unit of dynamic viscosity is kg/ms or Pa s. There is no name given to this unit, which for practical purposes is too large. More commonly used is the *centipoise* (cP), which is 1,000 times smaller than the Pa s. Table 4.1 provides the value of μ at metric standard condition (MSC).

Table 4.1 Typical value of dynamic viscosity for different gases

Gas	Dynamic viscosity (Pa.s)
Hydrogen	8.4×10^{-6}
Methane	10.7×10^{-6}
Air	17.0×10^{-6}
Water	$1{,}150.0 \times 10^{-6}$

Table 4.2 Typical value of Kinematic viscosity for different gases

Gas	Kinematic viscosity (mm^2/s)
Hydrogen	105.0
Methane	15.4
Air	14.0
Water	12

As we have seen, the *viscous forces* are proportional to the viscosity (μ), and since the forces of inertia are proportional to the density (ρ), the ratio μ/ρ, known as the *kinematic viscosity* (v), is a useful concept. The SI unit of kinematic viscosity (m^2/s) is too large for most practical purposes. A more commonly used unit is the centistoke (mm^2/s). Table 4.2 shows typical values of kinematic viscosity at MSC.

The magnitudes of the numerical values are reversed. For example, the absolute viscosity of air is about 1/70 that of water. The kinematic viscosity is about 12 times that of water. This is because the density of air is much lower than the density of water.

Because the viscosities of gases are generally of the same order of magnitude, it is possible to use correlations to calculate the viscosity of a gas mixture from the known viscosities of the individual gases that make up the mixture. Such a correlation is:

$$\mu_m = \frac{\sum \mu_i x_i \sqrt{\mathrm{MW}_i}}{\sum x_i \sqrt{\mathrm{MW}_i}} \tag{4.4}$$

where Mm is the absolute viscosity of the mixture, μ_i is the absolute viscosity of gas i, x_i is the mole fraction of gas i in the mixture and MW_i is the molecular weight of component i.

4.1.4 Pressure and Altitude

4.1.4.1 Pressure Head

The pressure at any point in a liquid is determined only by the vertical distance of the point of measurement below the free surface of the liquid. Pascal's law tells us that this pressure will be the same in any horizontal direction and is unaffected by

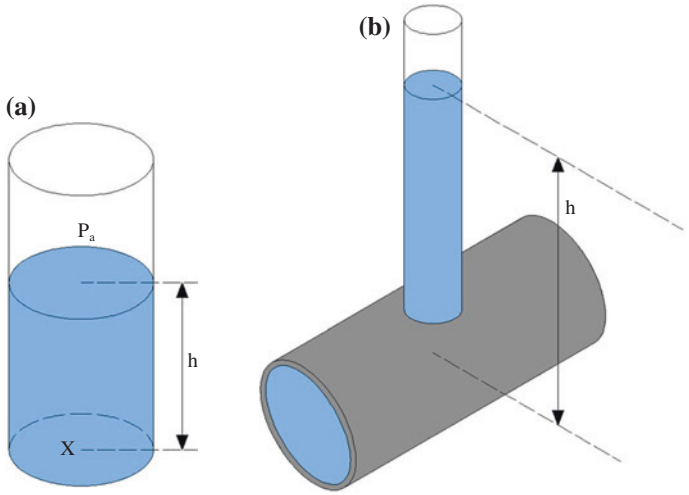


Fig. 4.2 Pressure at a point *X* in a fluid column (**a**) and the pressure head concept (**b**)

the shape or size of the vessel containing the liquid. In Fig. 4.2a, the absolute pressure of the liquid at point *X* is equal to the atmospheric pressure (P_a) plus the additional pressure resulting from the height of the liquid (h). This concept can also be applied to physically enclosed liquids where, for example, the pressure in a pipe containing water could be represented by a column of water extending upwards to a height h above the pipe as shown in Fig. 4.2b.

In each case, the pressure is given by:

$$P = \rho g h \tag{4.5}$$

where h is the pressure head in metres of water, p is the density, and g is the acceleration due to gravity (9.81 m/s^2). Clearly, trying to measure the pressure of gas in metres of gas is impractical, but we can measure it as an equivalent height of another fluid, for example water, by using a water-filled *manometer.*

4.1.4.2 Effect of Altitude

Equation 4.3 is only valid because the density of a liquid is assumed to remain constant over the height h. Gases are compressible, thus to determine the pressure at one point in a static column of gas relative to another point at a different elevation: first consider that the gas at any point to be in equilibrium; this must be so since the gas is not moving. Then, we consider a thin element (δh) over which the density can be considered sensibly constant, as shown in Fig. 4.3.

For the element to be in equilibrium, the forces on δh must balance (force *on bottom surface = force on top surface + weight of gas in element*). Thus, the pressure between the bottom and top can be expressed as:

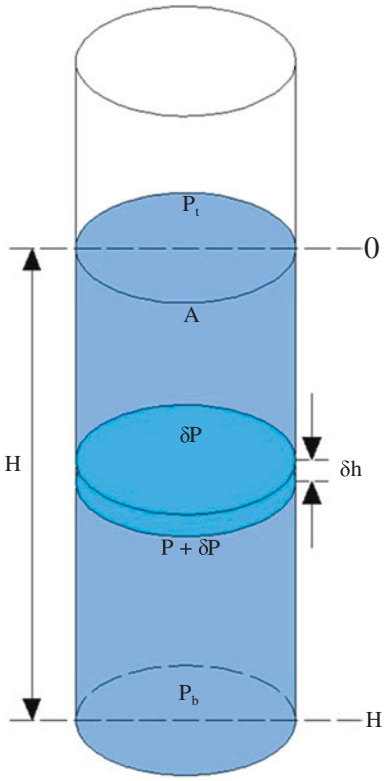


Fig. 4.3 The element δh in equilibrium

$$P_b = \exp\left(\frac{gH}{ZRT}\right)P_t \tag{4.6}$$

Or by substituting for $R = 287/S$, Eq. 4.6 becomes:

$$P_b = \exp\left(0.03417\frac{SH}{ZT}\right)P_t \tag{4.7}$$

where S is the relative density (air/gas), P_b is the bottom pressure (N/m^2) and P_t is the top pressure (N/m^2).

It should be clear from the results above that large elevation changes are required to produce even moderate corrections to the absolute pressure. As a rule of thumb, a static column of natural gas produces a pressure "head" equivalent to 1 % of the gas pressure for every 150 m of elevation. This may not appear to be very significant, but accuracy of pressure data is important for the effective utilisation of line pack and for modelling the operation of the transmission system on a daily basis. The equilibrium of the gas assumes that the pressure below the thin element ($p + \delta p$) is greater than the pressure above it (p). The pressure for

natural gas with no buoyancy effect is when the atmospheric conditions are the same as MSC, the density of the gas would need to be the same, assuming the density of air is 1.225 kg/m^3, with relative density of natural gas being typically 0.5 at a range of pressure. Therefore, using ideal gas law:

$$P = \rho_g R_g T_g = 1.225 \times \frac{287.06}{S} \times 288.15 \tag{4.8}$$

And for natural gas of relative density 0.59, this gives p 171,742 Pa or approximately 700 mbar (gauge). So, above 700 mbar, there is no buoyancy effect and Eq. 4.7 can be used, although as we have seen the effect is negligible unless large elevations are involved. The elevation differences within medium or intermediate pressure distribution systems are not great enough for Eq. 4.7 to have any effect so its use is restricted to transmission systems that traverse mountainous terrain. Below 700-mbar (gauge) natural gas is buoyant, relative to atmospheric air, so we must approach the problem from a different direction. Consider the vertical pipe, shown in Fig. 4.3, which is holding a static column of natural gas. There are effectively two columns of gas in equilibrium, one of natural gas inside the pipe, and one of atmospheric air outside. Because we are considering very low pressures, the density of each can be considered constant over the height *h.*

As can be seen from Fig. 4.4, if the gas gauge pressure was measured at the base of the column, it would be:

$$P_g + (\rho_a - \rho_g) gh \tag{4.9}$$

There is therefore an increase in the gauge pressure reading due to the difference in the densities of gas and air. This "increase" in gauge pressure would be negative if the gas density was greater than that of air (as with propane or butane, for example). This effect is only apparent because the density of low-pressure natural gas is of the same order of magnitude as atmospheric air. It is not apparent when measuring the pressure of a liquid, because the density difference is too great for variations in the density of air to be noticed. Of course, there is no change in the absolute pressure of the gas, so before carrying out any flow calculations, we must always correct any gauge pressure readings to a common datum. The gauge pressure correction factor, $(pa - Pg)gh$, can be simplified for general use by introducing the relative density and converting the pressure units to mbar, which leads to:

$$\text{Low pressure gauge correction} = 0.12(1-S)h \tag{4.10}$$

This simplified version of the gauge pressure correction factor is restricted to low-pressure use only—up to 75 mbar (gauge). The restriction applies because the use of the relative density requires that the gas is at or near MSC. As the pressure increases, the density increases, and as we have seen, natural gas has approximately the same density as air at 700 mbar (gauge). If accurate corrections are required to gauge readings above 75 mbar, they should be made using $(\rho a - \rho g)\, gh$.

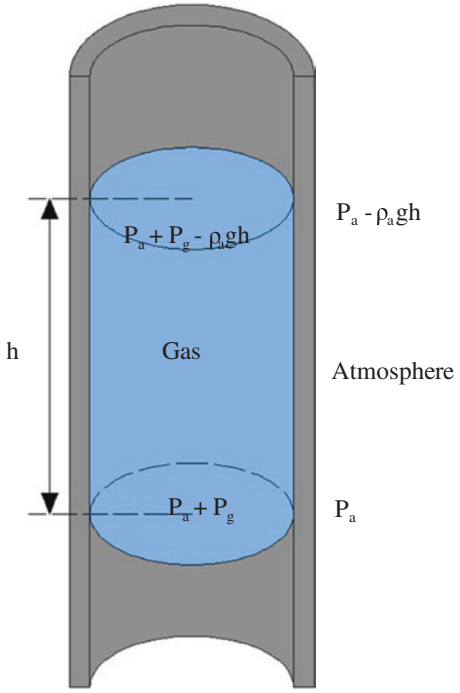


Fig. 4.4 Vertical pipe with static column of natural gas column of natural gas

4.1.5 Laminar and Turbulent Flow

4.1.5.1 Boundary Layer

Physical examination of a fluid flowing in a pipe shows that there is a thin layer of fluid adhering to the pipe wall and that the velocity of this layer relative to the pipe wall is zero. This zero-velocity layer affects the successive layers of flowing fluid. The normal distance from the pipe wall at which the flowing fluid ceases to be affected by the stationary layer is called the *boundary layer thickness*. The idea of a stationary layer of fluid, particularly gas, may seem surprising but it is true even for the smoothest of pipe materials. It is in the boundary layer that the fluid velocity changes with distance from the boundary, from zero at the pipe wall to the velocity of the main flow stream, approaching this velocity asymptotically as shown in Fig. 4.5. In the boundary layer, the velocity gradient normal to the pipe wall, *du*/*dy, is* large, so even a low value of viscosity is very significant and may not be disregarded even though its effect on the main flow may be negligible.

Beyond the *boundary layer thickness*, the free flow is affected much less by the fluid viscosity and in many practical applications may be treated independently as though the boundary layer did not exist

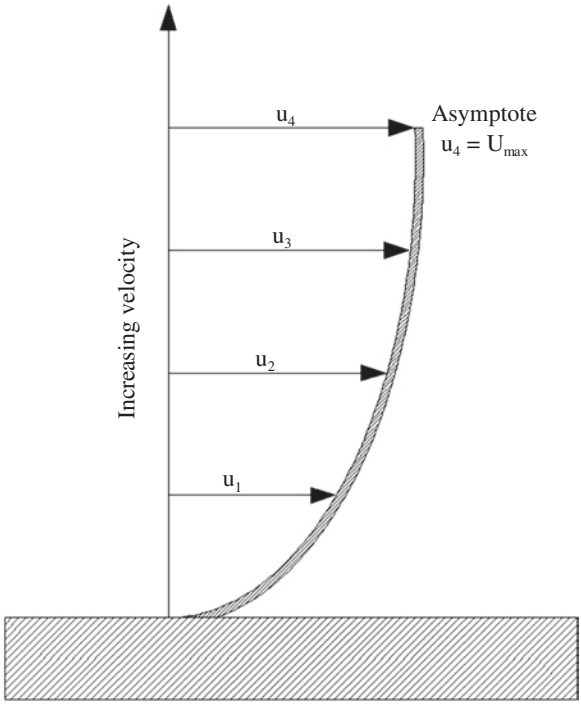


Fig. 4.5 Changes in fluid velocity with distance from the boundary surface

Fluid flow may be conveniently divided into two types of flow:

(i) *laminar flow* where the viscous forces tending to resist fluid movement predominates, creating a boundary layer that effectively extends to the centre of the pipe from the pipe wall.
(ii) turbulent flow where the *viscous forces* are restricted to a thin layer that extends only a short distance in from the pipe wall. The characteristics of turbulent flow will be described in Sect. 4.1.5.3.

4.1.5.2 Laminar Flow

In *laminar* flow (sometimes called *viscous flow*), the *viscous forces* predominate and the entire flow could be defined as a boundary layer. This usually occurs at low velocities. It is the viscosity of a fluid that determines the resistance to movement of fluid particles between parallel layers of the fluid. If the viscous forces are predominant, then the flow of that fluid will occur in layers. For a circular pipe, these layers must consist of concentric "tubes" of fluid. The resulting *velocity profile* is parabolic where the velocity of the fluid layers increases from zero at the pipe wall to a maximum value at the centre. In most pipe flow problems, it is not

the velocity of the gas that is of interest but the *quantity* being transported. Thus, the volume flow rate (Q) is given by the continuity equation as:

$$Q = U.A \tag{4.11}$$

where u is the velocity and A is the flow area, i.e. the cross-sectional area of the pipe. However, we cannot use the maximum velocity since this is not representative of the velocity profile, nor can we use the minimum, since this is zero. The actual volume flow rate is given by:

$$Q = U_{\mathrm{a}}.A \tag{4.12}$$

where U_{a} is the average velocity. In the case of the velocity profile for *laminar* flow, $U_{\mathrm{a}} = 0.5\ U\mathrm{max}$. The use of the average velocity (U_{a}) is valid provided the flow is steady and continuous. Laminar flow can be mathematically analysed without the use of any additional coefficients or experimental data. This analysis leads to Eq. 4.13, which is known as the Hagen-Poiseuille equation:

$$Q\frac{\pi d^4 \Delta P}{128\,\mu \mathrm{L}} \tag{4.13}$$

where Q is the volume flow rate, d is the diameter of the pipe, μ is fluid viscosity, ΔP is pressure loss over the length of the pipe and L is the length of the pipe. The two important features of *laminar* flow are: (i) the volume flow rate (Q) is inversely proportional to fluid viscosity (μ) and (ii) the pressure loss ΔP is directly proportional to Q.

4.1.5.3 Turbulent Flow

Turbulent flow can be described as flow where any of the fluid particles are not travelling in straight parallel lines; they begin to take on random and irregular flow patterns, although the principal motion of the fluid along a pipe is still maintained leading to fluid mixing. Figure 4.6 shows the transition of the flow from *laminar* to turbulent regime.

The *laminar* layer adjacent to the pipe wall is known as the *laminar sub-layer*, and the flow is described as being *partially turbulent.*

The boundary layer, remember, is the distance from the pipe wall at which the flowing fluid ceases to be affected by the stationery layer, so in Fig. 4.6c, it is defined by the distance at which the velocity profile no longer changes with distance from the pipe wall. So within the boundary layer, we have two types of flow occurring, the *laminar* sub-layer and an outer turbulent region. Further increases in flow rate (and hence velocity) cause the laminar sub-layer to become thinner. Theoretically, the layer would continue to decrease in width indefinitely as the fluid velocity increases. However, in practice, the laminar sub-layer becomes—so thin that it no longer covers the surface roughness of the pipe. In this situation, elements of the internal surface of the pipe are in contact with areas of turbulent flow. When this occurs, the fluid flow is described as *fully developed* turbulent flow.

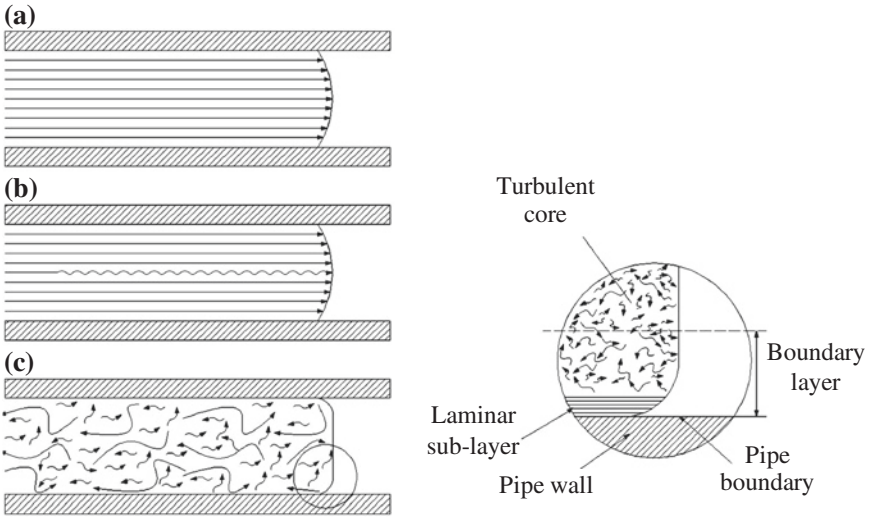


Fig. 4.6 The transition of flow from laminar to turbulent

4.1.6 Predicting Flow Types

Osborne Reynolds in the late nineteenth century argued that any general law governing the flow of fluids must be concerned with the criteria that make the behaviour of the fluid similar to the behaviour of that fluid, or any other fluid, when similar conditions are imposed on it. For fluid to be similar, the shape of the conduits in which they are flowing must be similar, and the forces causing the fluid to move must also be similar, that is when the forces acting on a particle in one flow are in the same ratio of magnitudes as the forces acting on a particle at the corresponding point in another flow. For natural gas flow, the first condition is usually satisfied since the gas is transported in circular pipes.

Reynolds reasoned that the ratio of the inertial force to the viscous force (see previous Sect. 4.1.3) would not only determine similarity of flow but would also indicate the *change* of flow regime between *laminar* and turbulent flow. This is providing of course that the shape of the conduit is similar in all cases. So, for a circular pipe of diameter *d,* Reynolds Number (*Re*) is given by:

$$Re = \frac{\rho \overline{U} \mathrm{d}}{\mu} \tag{4.14}$$

It represents the ratio of the forces of inertia to the *viscous forces* and hence is dimensionless. A high value of ρ, d or U^{-} or a small value of μ gives a high value of Re, and of course, the converse is also true. Natural gas has a low value of viscosity and a low value of density, but do you remember. The kinematic viscosity (see Sect. 4.1.3) for natural gas is large, compared with water for example, so this tends to lead to high Re values. When this is combined with a large diameter, as

with some transmission pipelines, the Reynolds number under these circumstances would be very high. Re cannot change if the pipe dimensions and gas viscosity remain constant because Reynolds number is an indication of fluid flow similarity. If the quantity of gas flowing in at one end is the same as the quantity leaving at the other and then the flow is not just similar, it is the same. For steady state flow, the mass flow rate remains constant along the length of the pipe, so as the pressure (and density) decreases the volume flow rate (Q) must increase and since $U^{-} = Q/A$, the velocity also increases. It is this increase in velocity that compensates for the fall in density and maintains a constant value of *Re*.

For natural gas, since both density and velocity change with falling pressure, we usually calculate *Re* as though all the flow occurred at MSC. Of course, the flow cannot actually occur at MSC, but this does not matter because the value of *Re* is determined by complementary values of ρ and U^{-}, so MSC are just as valid as any other set of conditions.

For steady state gas flow, the mass flow rate of gas entering a pipe is equal to the mass flow of gas leaving the pipe, so the *continuity equation is:*

$$\rho_1 A_1 \bar{U}_1 = \rho_2 A_2 \bar{U}_2 = \rho_s A_s \bar{U}_s \tag{4.15}$$

This is valid not only for the continuity of mass flow in the pipe but also for expressing the mass flow rate of the same gas at MSC. Thus, by using Eq. (4.14) and substituting Eq. (4.15), therefore, the *Re for natural gas* can be expressed as:

$$Re = 25,043 \frac{Q}{d} \tag{4.16}$$

Assuming Q_s to be in m^3 (st)/h and d to be in mm and $\rho_a = 1.225$ kg/m^3, $\mu = 10.38 \times 10^6$ kg/m s and relative density $S = 0.6$.

A high value of *Re* indicates that the forces of inertia dominate the flow, so it is likely to be turbulent. When *Re* is small the *viscous forces* are the most significant so the flow will probably be *laminar*. As a general rule, the following types of flow can be expected for the given ranges of *Re*. These are: $Re < 2{,}000$ laminar, *Re* 4,000 > partially turbulent and $Re > 10^7$ fully turbulent.

The region between $Re = 2{,}000$ and $Re = 4{,}000$ is known as the *critical zone.* It is so called because the flow cannot easily be defined as either *laminar* or turbulent; it is the point at which the *inertia forces* are approximately equal to the *viscous forces*.

4.1.7 The Effects of Friction on Flow

Friction in fluid flow, as with any other process, results in a loss of energy, but in order to be able to lose *energy*, the fluid must first *possess* it! The idea that fluids possess energy was first postulated by Euler in the eighteenth century. Euler describes the *total energy* possessed by a moving fluid as follows:

$$\mathrm{d}x + \frac{\mathrm{d}p}{\rho.g} + \frac{u.\mathrm{d}u}{g} = 0 \tag{4.17}$$

where x is the elevation above some arbitrary datum, P is absolute pressure, u is fluid velocity, ρ is the fluid density and g is gravitational acceleration.

The density of natural gas ρ varies with pressure. So, it would be necessary to relate ρ and p using the equation of state. For now, let us consider an incompressible fluid, but we will return to the question of a compressible fluid in a later section.

$$x + \frac{P}{\rho.g} + \frac{u^2}{2g} = \text{Constant} \tag{4.18}$$

This is generally known as Bernoulli's equation.

Each of the terms in Eq. 4.18 possesses the unit of length, and so the equation describes the *total energy head* of a given fluid. The quantity x is the elevation above some arbitrarily chosen horizontal datum and represents potential energy. The quantity $u^2/2\ g$ is known as the velocity head and represents *kinetic energy.* The quantity $p/\rho g$ is known as the pressure *head* and represents the energy available from pressure forces. This energy is not stored, in the sense of a compressed fluid. It is only available if there is a pressure difference in the fluid system that will allow the transmission of this pressure energy. In fact, none of the terms in Bernoulli's equation represent stored energy, but together they represent the total mechanical energy to be *transmitted* by the fluid.

Let us return to fluids in an ideal fluid system in which there are no transmission losses, so the total energy head will remain constant between any two points in the system. However, including the transmission losses, which is caused mainly by friction that can be denoted by h_f, Eq. (4.18) at any two points in the fluid can be expressed as:

$$x_1 + \frac{P_1}{\rho_1.g} + \frac{u_1^2}{2g} = x_2 + \frac{P_2}{\rho_2.g} + \frac{u_2^2}{2g} + h_f \tag{4.19}$$

In *laminar* flow, the fluid moves in layers with a stationary layer of fluid at the pipe wall, so all the pressure loss is due to the properties of the fluid [refer to Eq. (4.13)]. There is no contact between the pipe wall and *the flowing fluid.*

In fully turbulent flow, the surface roughness of the pipe wall is in contact with the flowing fluid, all traces of laminar flow have disappeared and the pressure losses are due entirely to the friction caused by the fluid flowing over the pipe wall surface.

For partial turbulence, where there is a *laminar* sub-layer adjacent to the pipe wall, the losses are due to a combination of pipe wall properties and fluid properties. Let us now consider frictional head loss in laminar and turbulent flow regime.

4.1.8 Frictional Head Loss in Laminar and Turbulent Flow

4.1.8.1 Frictional Head Loss in Laminar Flow

By using the Hagen-Poiseuille equation for *laminar* flow (4.13), the continuity Eqs. (4.12) and (4.19), the frictional head loss in laminar flow (due to friction between layers of fluid) is seen to be inversely proportional to *Re* and the frictional effects can be represented directly by *16*/*Re,* which can be expressed as:

$$h_{\mathrm{f}} = \frac{4L\overline{U}^2}{d.2g} \times \frac{16}{Re} \tag{4.20}$$

4.1.8.2 Frictional Head Loss in Turbulent Flow

Turbulent flow is highly complex due to the random movements of fluid particles superimposed on the main flow, since these random movements are unpredictable, there is, unlike *laminar* flow, no complete analytical solution for turbulent flow. However, if we look at the result of the above activity, we see that of the parameters involved, only *16*/*Re contains* the value of the fluid viscosity, which is the root cause of the head loss in laminar flow. In a general equation for head loss, including turbulent flow, we might expect an expression of the form:

$$h_{\mathrm{f}} = \frac{4fL\overline{U}^2}{\mathrm{d}.2g} \tag{4.21}$$

where *f* is a general friction factor.

As early as 1841, Weisbach had developed a resistance equation using a dimensionless coefficient to describe the energy lost by a fluid in motion. At around the same time, the French engineer Darcy was working on similar ideas, as expressed in Eq. (4.21).

This is commonly called the Darcy–Weisbach equation, but you may see it referred to simply as the Darcy equation. It is identical to that suggested by the extension of the *laminar* flow equation where *16*/*Re* is replaced by an empirical friction coefficient (*f*). Because *f* is an empirical coefficient, the method of evaluating it is not universal and various methods have been developed for this purpose. In particular, you should be aware that equations presented in other texts might use a value of *f* that is a multiple of the Darcy friction factor.

In North America, it is common to use a friction factor that is four times that defined by Eq. 4.20; this does not matter as long as the numerical coefficient in the equation is adjusted accordingly. In Britain, Transco uses both $u^{-}/2\ g$ and *L*/*d* as performance criteria in the specifications for the supply of fittings to be used in their gas distribution system.

4.1.9 Friction in Turbulent Flow

The parameters that affect the friction losses are represented in Fig. 4.7. Figure 4.7a represents the theoretical concept of a pipe with a perfectly smooth internal surface. If such a pipe existed, the *laminar* sub-layer would always completely cover the pipe wall even though it would become very thin at high velocities. In such circumstances, the main body of turbulent flow would never come into contact with the pipe wall so it is reasonable to assume that the wall plays no part in the resistance to flow. In fact, as with laminar flow, only the value of *Re* would be important.

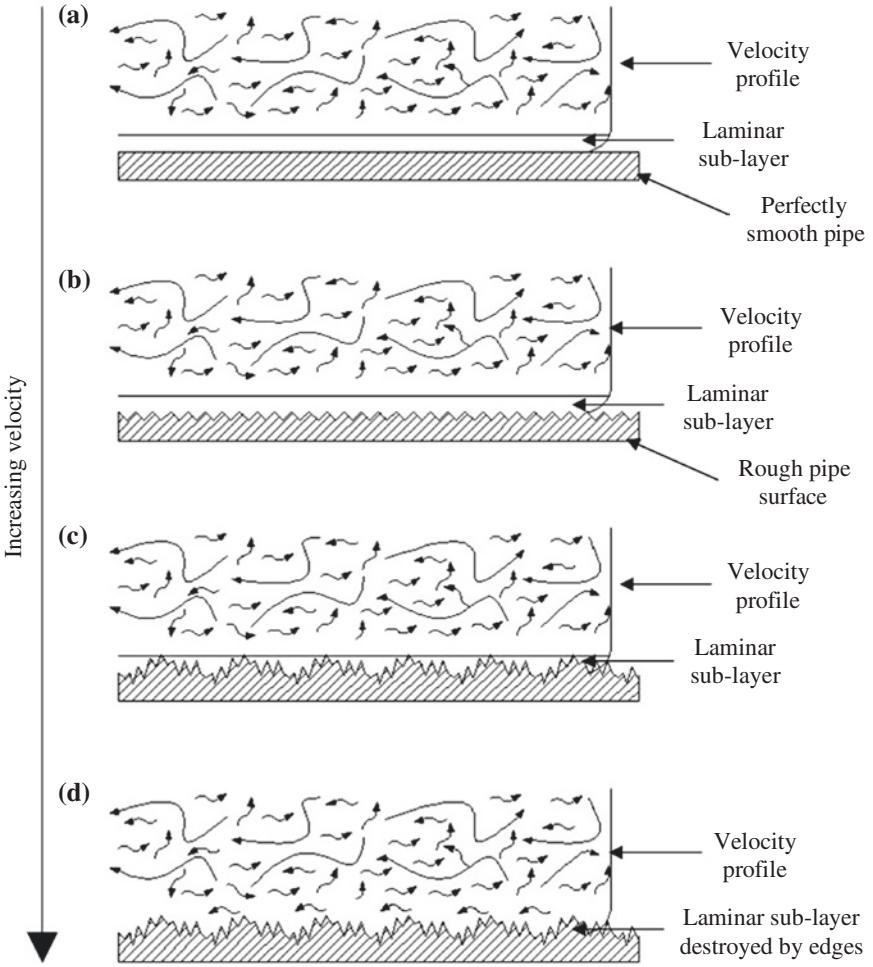


Fig. 4.7 Effect of friction losses

Figure 4.7b represents a real pipe with an internal surface, which consists of small particles, creating a certain amount of roughness. In this case the *laminar* sub-layer is thick enough to completely cover the surface roughness, so, as with a theoretically smooth pipe, the turbulent fluid flow does not come into contact with the internal pipe surface. It behaves as though it were a smooth pipe and so only the value of *Re* is significant.

In Fig. 4.7c, the *laminar* sub-layer has become thinner due to increasing velocity and now some of the roughness peaks are just protruding through the sub-layer and into the turbulent flow. The flow is no longer independent of the internal pipe surface, but it is still partly dependent on *Re* since there is still a laminar sub-layer.

Finally, in Fig. 4.7d, increased velocity has caused the *laminar* sub-layer to shrink even more, allowing the pipe roughness to protrude further into the turbulent flow stream causing eddies to form around each particle that destroys the remains of the sub-layer. Since there is no laminar sub-layer, the influence of fluid viscosity disappears and the flow is no longer dependent on *Re,* only the surface roughness of the pipe.

4.1.9.1 Relative Roughness

The progression from smooth pipe flow to rough pipe flow is obviously dependent on the surface roughness. However, the description of fully developed turbulent flow as rough pipe flow does not indicate that the pipe is actually very rough, only that the roughness is totally responsible for the friction losses. The degree of pipe roughness is described by its *relative roughness*.

It is not easy to express the internal roughness of pipes in a quantitative manner, because we all have a different idea of what "rough" is. All surfaces, no matter how well polished, are to some extent rough, in so much as there are minute particles making up the surface of the material. The irregularities created by these minute particles vary greatly in shape, size and spacing, making an absolute measure of roughness very difficult. If we could make measurements of the height of these surface particles, we could arrive at a value for the mean height, which over a considerable area, could be expected to influence the frictional resistance to flow. Extending Reynolds' reasoning on the similarity of flows would indicate that a relative roughness rather than an absolute roughness would be significant in determining the characteristics of the flow. In the case of circular pipe, this is:

Relative roughness = Mean height of surface particles (Σ)/Internal pipe diameter (d)

This is also shown diagrammatically in Fig. 4.8.

4.1.9.2 Smooth and Rough Pipe Laws

Figure 4.7 represents the perfectly *smooth pipe* shown in Fig. 4.7a, but it also represents the situation shown in Fig. 4.7b where the *laminar* sub-layer completely covers the surface roughness. You should note that the smooth pipe friction factor (*f*sp) is dependent only on *Re.*

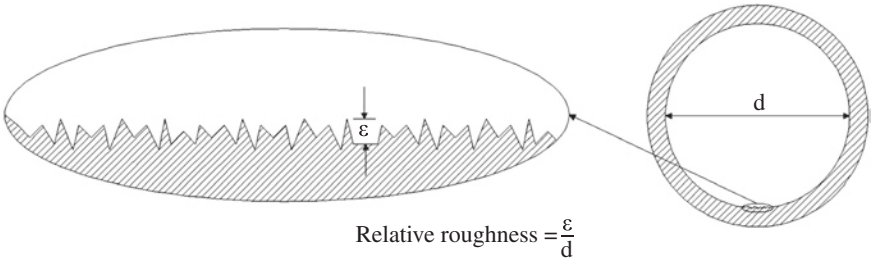


Fig. 4.8 Relative roughness of the internal surface of a pipe

The *rough pipe friction factor* (f_{rf}) is dependent only on the values of d and Σ (the relative roughness), and since these are constants for a given pipe, the friction factor is also constant.

The inverse of the relative roughness appears in the rough pipe law. In some textbooks, particularly, those published in the USA, the inverse is used instead of the relative roughness. The ratio d/M is known as the roughness factor.

4.1.9.3 The Moody Diagram

The American engineer Lewis F. Moody, in a synthesis of earlier work, presented his concept of fluid flow behaviour in the form of a diagram, which now bears his name—the Moody diagram. Figure 4.9 is a Moody diagram based on the Darcy friction factor (for *laminar* flow $f = 16/Re$). Remember, you may see other diagrams based on multiples of this, for example $f = 64/Re$, for laminar flow. Friction factor should always be checked using the diagram to identify the following:

i. the *laminar flow* region
ii. the *smooth pipe*curve
iii. a series of lines representing values of *relative roughness*
iv. the *critical zone* between *laminar* and turbulent flow (*B*)
v. the region of *complete turbulence* where the rough pipe law applies (*D*)
vi. the *transition zone* between the smooth pipe and rough pipe regions (*C*).

In Moody's diagram (Fig. 4.9), the curves representing the transition from smooth pipe to rough pipe are based on Colebrook's equation and his experimental data using commercial grade pipes. These curves are used widely for predicting friction factor values in general pipe work systems

4.1.10 General Flow Equation

In order to develop the general flow equation, it is necessary to make some assumptions about the pipe system and the flow through it. As we will see, the assumptions do not place any excessive restrictions on the use of the equation. The

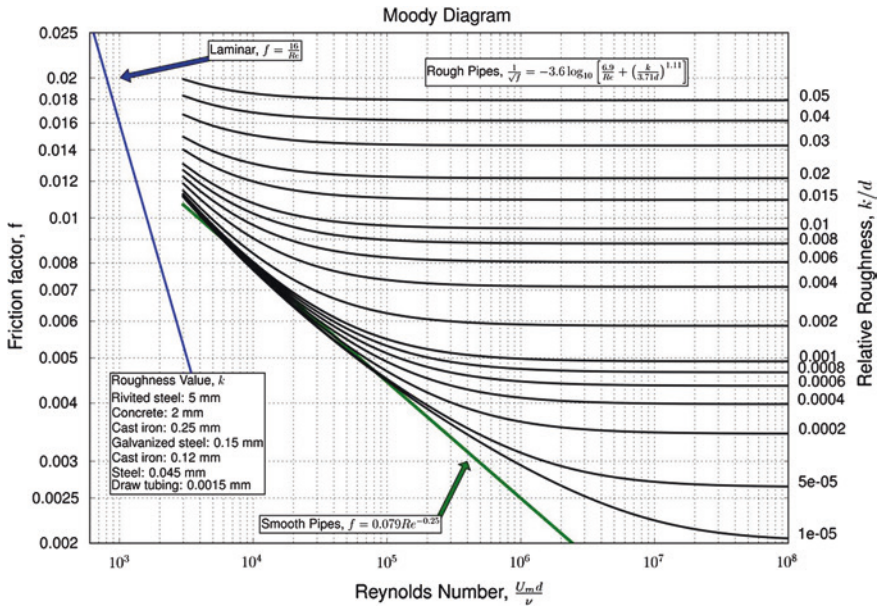


Fig. 4.9 Moody diagram

flow process is isothermal, changes in the kinetic energy of the gas are negligible, the pipeline is horizontal, no mechanical work is done by or on the gas, the energy loss due to friction is given by the Darcy–Weisbach Eq. (4.21) and the flow is steady state. Using Eqs. 4.17–4.21, the gas velocity at standard conditions for a given pressure drop can be expressed as:

$$\bar{\mathrm{U}}_s = \frac{T_\mathrm{s}}{P_\mathrm{s}} \sqrt{\frac{\left(P_1^2 - P_2^2\right)\mathrm{d}.R_\mathrm{g}}{2f.L.Z.T}} \tag{4.22}$$

The actual gas velocity varies along the pipe. It does not occur at standard conditions, but this is a convenient way of expressing it because it allows us to solve Euler's equation for gas flow and can be converted to a volume flow rate at standard conditions. Thus, the *general flow equation* can be written as:

$$Q = \frac{7.574 \times 10^{-4}}{\sqrt{f}} \frac{T_\mathrm{s}}{P_\mathrm{s}} \sqrt{\frac{\left(P_1^2 - P_2^2\right)\mathrm{d}^5}{S.L.Z.T}} \tag{4.23}$$

In this equation, all the units are self-consistent, but for general use, it is convenient to express it in terms of the units that are normally used:

• Pipe length (L) m	Gas pressure (p) bar
• Pipe diameter (d) mm	Gas flow rate (Q) m^3 (st)/h
• Gas temperature (*T*) *K*	

This equation applies over all pressure ranges and is the basis for many of the flow equations used in the analysis of transmission and distribution networks.

In summary, the general flow equation, using gas industry units, is given by:

$$Q = \frac{7.574 \times 10^{-4}}{\sqrt{f}} \frac{T_s}{P_s} \sqrt{\frac{\left(P_1^2 - P_2^2\right) d^5}{S.L.Z.T}} \tag{4.24}$$

The general flow equation can be simplified, for use in MP systems, to the following:

$$Q = \frac{1.269 \times 0^{-2}}{\sqrt{f}} \sqrt{\frac{\left(P_1^2 - P_2^2\right) d^5}{S.L}} \tag{4.25}$$

For use in LP systems, the general flow equation can be simplified further to

$$Q = \frac{5.712 \times 10^{-4}}{\sqrt{f}} \sqrt{\frac{(P_1 - P_2) d^5}{S.L}} \tag{4.26}$$

4.1.11 Friction and Smooth Pipe Law

As we are now familiar with the flow equations used for gas transmission and distribution systems, let us consider the practical implications of using the smooth pipe law. In particular, let us examine the use of an efficiency factor for modelling the energy loss more accurately.

The transition from smooth pipe flow to fully turbulent (rough pipe) flow is shown as a series of curves in the Moody diagram. The curves are asymptotic to the smooth pipe and rough pipe curves and represent Colebrook's data. So far, gas flow has been described as following the smooth pipe law, but no real system can follow it exactly because real systems are not perfectly smooth. Therefore, some significant factors such as actual pipe roughness, pipe joints, welds, bends, tees, fittings, valves, dust and debris.

In the case of gas transmission pipelines, these factors are uniformly distributed throughout the system and form a very small proportion of the overall pipeline length. The US Bureau of Mines' tests indicate that the deviation from the smooth pipe law could be represented by a fixed percentage of the smooth pipe flow. This leads us to the idea of an efficiency factor that could correct the smooth pipe law to actual conditions which can be expressed as:

$$E = \frac{Q_{\text{actual}}}{Q_{\text{smooth pipe}}} \tag{4.27}$$

where E is an efficiency factor that converts the smooth pipe transmission factor into an actual transmission factor.

Taking the general flow equation as an example, this could be written as

$$f_{\text{actual}} = \frac{f_{\text{sp}}}{E^2} \tag{4.28}$$

A typical value of E is 0.95 for a new pipeline and is constant for a wide range of Re. However, for distribution systems, this must be treated somewhat differently because of the higher incidence of drag-inducing elements. These elements induce energy losses, which can be expressed as a number of velocity heads ($u^2/2$ g).

The total energy loss is obtained by adding together the velocity head losses of all the elements. Since there are a significant number of elements, the overall energy loss is sufficient to create a deviation from the smooth pipe law, which is dependent on the gas velocity. This deviation from smooth pipe conditions is not great enough to be considered as a Colebrook-type transition (for new pipes at least), but is sufficient to cause a deviation that increases with velocity. So for low-pressure distribution systems, we have:

$$f_{\text{actual}} = \frac{f_{\text{sp}}}{e^2} \tag{4.29}$$

Values of e at two velocities are given in the Table 4.3 for polyethylene pipe. Values of e selected from the Table 4.3 with the nearest velocity to that required are usually adequate for general calculations.

4.1.12 Other Flow Equations

The general flow equation developed in Sect. 4.1.10 is the basis for all gas flow equations and is valid for all pressure and flow conditions found in practice. One could come across other flow equations with other names, but these are all variations of the general flow equation. Usually, it is the method chosen for the determination of the transmission factor that gives its name to the flow equation. In this section, we will look briefly at some of these and comment on their usefulness.

Before the introduction of computers, the need to solve the general flow equation by hand led to the development of some empirical but explicit friction factor relationships. Here are some of them

Blasius	$\frac{1}{\sqrt{\text{f}}} = 3.558\, Re^{0.125}$
Meuller	$\frac{1}{\sqrt{\text{f}}} = 3.350\, Re^{0.130}$
Fritsche	$\frac{1}{\sqrt{\text{f}}} = 5.145(Re.\text{d})^{0.125}$
Polyflo	$\frac{1}{\sqrt{\text{f}}} = 5.350\, Re^{0.076}$
Panhandle "A"	$\frac{1}{\sqrt{\text{f}}} = 6.87\, Re^{0.07305}$

Blasius was the first to suggest a relationship of this kind, so equations of this kind are known as "Biasius-type" equations. Polyflo and Panhandle "A" are perhaps the best known, but they are all approximations to the smooth pipe law. Other friction

Table 4.3 Typical pipe-sizing criteria

Nominal pipe size (mm)	SDR	Maximum internal bead height (mm)	Velocity = 1.5 m/s efficiency factor		Velocity = 4.7 m/s efficiency factor	
			e_6	e_{12}	e_6	e_{12}
90	11	5.0	0.91	0.95	0.87	0.93
125	11	5.0	0.92	0.96	0.89	0.94
180	17	5.0	0.94	0.97	0.92	0.96
250	17	6.3	0.92	0.96	0.89	0.94
315	17	7.9	0.90	0.95	0.87	0.93
355	17	8.9	0.89	0.94	0.86	0.92
400	17	10.0	0.87	0.93	0.83	0.90
500	17	12.5	0.84	0.91	0.91	0.88

e_6 represents 6 m between butt welds and
e_{12} represents 12 m between butt welds

factor relationships that depend only on the pipe diameter, such as Unwin, Pole, Weymouth and Spitzglass, are crude approximations and *should not* be used.

The Panhandle equation is typical of gas flow equations that use a transmission factor relationship of the Blasius type. Since it represents gas flow reasonably accurately, albeit within restricted ranges of *Re,* we can adopt the general form as a basis for describing gas flow in pipe networks.

Take another look at the Panhandle "A" flow in Sect. 4.1.12. All equations of this type may be represented by:

$$\Delta P^2 = K.Q^n \tag{4.30}$$

where n is the flow exponent and K is the resistance factor and are chosen according to the type of fluid and pipe system

The LP gas flow equation has the general form

$$\Delta P = K.Q^n \tag{4.31}$$

At 75 mbar, the absolute pressure is 1.08825 bar, so the maximum pressure loss that can be accommodated using the LP equation is 2 % of this, i.e. 22 mbar.

4.1.13 Gas Velocity

We mentioned earlier about the effect of velocity on Reynolds Number and the compensating effect of density, which results in a constant value of *Re* along the length of a pipeline for a fixed mass flow of gas. We also extended this concept using the continuity Eq. (4.14) to allow us to use the volume flow of gas expressed at MSC in the general flow equation.

This is a very convenient way of expressing the flow; however, the actual conditions in the pipe are affected by the change in velocity (u) and density (ρ), for example dust migration is caused by high gas velocity [2].

The highest velocity occurs at the end of a pipeline where the pressure (and hence the density) of the gas is lowest. In order to maintain the same mass flow as the density falls, the velocity must increase. The Institution of Gas Engineers codes of practice recommends that the maximum velocity in a pipeline should not exceed 20 m/s. This is a practical constraint based on the damage that could be inflicted by gasborne dust, not a theoretical velocity limit. Because of the importance of gas velocity in the management of dust, it is often necessary to know what the gas velocity is likely to be at a particular point in the system. The average velocity is given by $u = Q/A$, but we cannot use the flow rate calculated from the flow equations directly to obtain the velocity. The flow equations give the gas flow rate in m^3(st)/h, so the velocity calculated from $u = Q/A$, would not be the actual gas velocity. The actual gas velocity is lower because the gas is under pressure. We must correct the flow rate to actual m^3/h before calculating the velocity.

From the ideal gas law, we get:

$$\overline{U} = \frac{Q.P_s}{A.P} \tag{4.32}$$

where P_s, is the pressure at MSC and p is the actual gas pressure at the point of interest.

Remember, the gas velocity increases in the direction of flow as the pressure decreases, so at any particular point in the system where we need to know the velocity, we may not know the actual pressure. Usually, it is only convenient to measure the gas pressure at locations where the system is above ground, but the point at which we need to know the velocity may be some distance away. Take a look at Fig. 4.10. This shows a pipe where the value of velocity is required a distance L from the point at which the gas pressure is measured. Using the following equation enables us to calculate gas velocity in any distribution main operating at pressures up to 7 bar (gauge):

$$u_2 = 353.68Q.P_s\sqrt{\frac{\mathrm{d}}{\mathrm{d}^5.P_1^2 - 6211.2Q^2f.S.L}} \tag{4.33}$$

The speed at which any gas can travel is limited to the speed at which sound can travel in the same gas. The speed of sound in Methane gas is 396 m/s. In gas pipelines, this very high velocity is not a practical constraint because the gas normally f1ows at less than 20 m/s. However, when gas escapes in an uncontrolled manner from a hole or pipe break, it will escape at the speed of sound provided that the pressure ratio of the internal pressure to the atmospheric pressure is more than the *Critical Pressure Ratio.* This ratio varies according to the gas composition but for a typical natural gas, it is approximately 2:1. For the critical pressure ratio to be 2:1, then the absolute pressure of the gas in the pipe must be twice the absolute atmospheric pressure. Taking the atmospheric pressure to be approximately 1 bar abs, then the internal pipe pressure would need to be 2 bar abs, or approximately l bar(gauge). Therefore, in all gas supply systems, except low-pressure distribution systems, escaping gas will travel at the speed of sound. The speed of sound

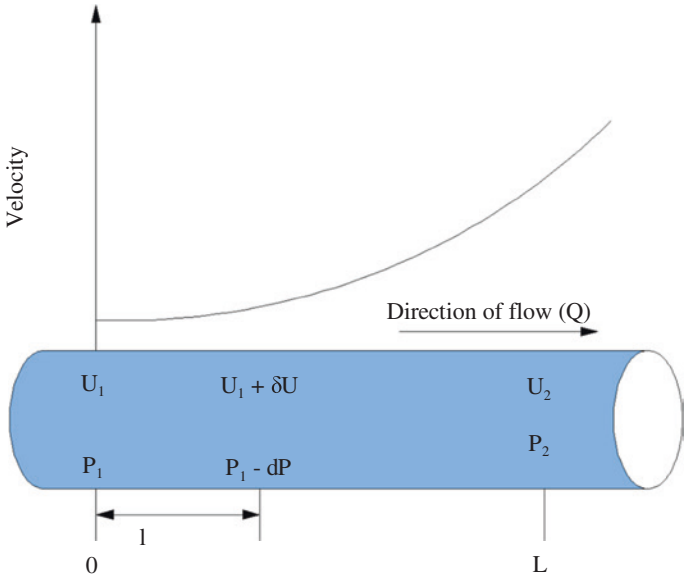


Fig. 4.10 Change of velocity with pressure and distance

is a physical property of a gas, and it has important implications for the design and operation of pipeline systems operating under *transient flow* conditions as discussed in the next section.

4.2 Network Analysis

4.2.1 Introduction

Natural gas is one of the world's largest sources of primary energy supply. For many years, this type of gas was flared in the production of crude oil, but gradually it began to replace town gas, mainly from coal. Over the years, gas reserves have grown at a substantial rate and natural gas has developed into a necessity across the world.

For the efficient, safe and reliable exploitation of natural gas resources, transportation of natural gas from various locations to consumers is of prime importance. The gas industry is a very capital intensive industry. It can be suggested that the economics of its transportation may well be as important as the safety aspects. Therefore, design engineering and economic judgment are very important in the viability of any gas project.

In gas transmission systems, the gas flows are constant, and most operate 24 h a day, 7 days a week, 365 days a year. However, particularly in long-distance pipelines, time-varying processes do exist in the systems due to the starting

and stopping of compressors or the sudden increase and decrease in consumer's demands. Therefore, because of these variations, unsteady state and transient conditions have to be considered in the design of gas transportation systems. These systems are known as "networks". In simple terms, it is the simulation of flow and pressure in pipes and the process of ensuring that a gas network can meet the specific operational duties required of it and/or the design criteria under consideration.

Gas flow in pipes is a complex process and the parameters associated, such as velocity, pressure and density that aim to illustrate the behaviour and state of a fluid are not always constant. They vary from one point in the pipe to another, and thus, it can be suggested that problems involving the flow of gas can occur.

It is apparent that modelling a network is a complex task. In this report, the important concepts of unsteady and transient analysis will be examined along with highlighting the electronic calculating equipment and equations that are used in network analysis systems. The advantages and disadvantages of using such systems in the gas industry today will also be elaborated on and a typical industry calculation will be provided to present how the maximum and minimum operating pressures of a network can be calculated. Every effort has been made to provide a comprehensive examination of network analysis and the reasons why it is such a popular choice in the gas industry. In order to support the descriptive investigation of network analysis, several workplace examples have been used throughout this report. These examples have been taken from scotia gas networks (SGN), one of the major gas distribution and transmission networks currently operating in the United Kingdom (UK), predominately in Scotland and the southern regions of England.

Before the introduction of network analysis techniques, the only solution in solving gas transmission and distribution networks was by trial and error. Estimating the flow of natural gas through each individual pipe and then making subjective judgments until a balance was achieved and a solution obtained. Obviously, this was a long and difficult process to undertake and the margin of error was high.

In 1936, a method known as the "*Hardy Cross*" was introduced into the industry. This method attracted a lot of attention across the world; however, it still required a large number of calculations to be made and thus was only practical when addressing small networks. It was not until 1965 that the first pipeline network analysis solution was solved on a computer and ever since; this method has taken off and has been widely adopted across the world. There are now several methods of network analysis systems introduced in the gas industry and most of them adopt the Hardy Cross solution in order to sufficiently manage the network and control the parameters associated with transient flow.

Transient flow in natural gas systems is flow where the flow velocity and pressure are changing with time. When changes occur to a fluid systems such as the starting or stopping of a compressor, closing or opening a valve, a pipeline failure or large demand surge then transient flow conditions arise; otherwise, the system is steady state. Often, transient flow conditions persist as oscillating pressure and velocity waves for some time after the initial event that caused it.

Transient flow can result in significant transient pressures that may exceed the design limit of the pipes and fittings. Therefore, a detailed understanding of transient flow analysis is very important. Transient analysis deals with the dynamic nature of pressure and flow in gas transmission systems. The behaviour of the gas flow in the pipeline is derived from the mass balance, momentum balance and the energy balance equations. These problems are commonly solved using a finite difference method known as the method of characteristics. The equations generated by this method can only readily be solved with a computer, which relates pressure, flow and their derivatives. Hence, network analysis exists.

4.2.2 General and Industrial Applications

4.2.2.1 General Application

The general application of network analysis is described as an influential system that is used across many industries. These industries include electricity, petrochemical, water and gas. Network analysis is also frequently used in the process pipework design and power station cooling water systems. Inevitably, it is apparent that network analysis is a well-used system in a whole variety of industries across the world.

Gas distribution companies model their low-pressure networks in the same way that water companies make use of network analysis. Each of them adopts the basic steady state equations shown in Sect. 4.1.12. Furthermore, electricity companies across the world also use the same techniques for modelling their urban supply networks. In recent years, the derivation from methods used to analyse large electrical networks has been introduced and used in the gas, petrochemical and water networks. These basic electrical circuits can be analysed using Ohm's law: $V = \mathrm{IR}$, where, $V =$ voltage drop (volts), $I =$ current (amps) and $R =$ electrical resistance (Ohm). The terms in a basic steady state Eq. 3.30 ($\Delta p = KQ^n$) are analogous to each of the terms in Ohm's law. Voltage drop (V) is analogous to the fluid pressure loss (Δp); electrical resistance (R) is analogous to the resistance factor in fluid systems (K) and current (I) is analogous to the flow rate (Q). From analysing this theory, it is apparent that fluid networks are analogous to electrical networks, with the main difference between the equations being that the fluid flow equation is nonlinear and this affects the way in which we approach the solution of gas networks.

4.2.2.2 Application to Gas Systems

As recalled by Babonneau et al. [1], "problems dealing with the design and the operations of gas transmission networks are challenging". However, the introduction of network analysis has had a significant role to play in easing the difficulties and problems that can arise and has provided gas networks with more success when designing

and operating either simple single delivery lines or large complex networks. In the UK gas industry, network analysis is carried out in accordance with the guidelines set out in industry standard documents; IGE/GL/1: Planning of Gas Distribution Systems Operating at Pressures Not Exceeding 7 bar and IGE/GL/2: Planning of Transmission and Storage Systems Operating at Pressures Exceeding 7 bar.

Network planning and operational analysis are supported by the creation of network models using network analysis software systems, to simulate network performance under various criteria. Network analysis simulates the performance of gas networks to enable the effects of physical changes to be anticipated and understood. It provides a mean of determining the dynamic interface requirements of gas infrastructure systems, and it delivers essential information to support business cases for the operation of and investment in safe, economic and efficient networks, pursuant to the obligations under the Gas Act 1986 applicable to gas transporters in the United Kingdom.

Once the basic network analysis model has been established, it can then be adopted over a large array of application uses. Network analysis is used for a number of purposes including pipeline design; hydraulic analysis; the development of operation plans; and analysis of alternative scenarios. During the design stage, it can:

- Calculate the maximum/minimum throughput of gas through any particular pipeline system.
- Optimise pipe dimensions and pressures in the network.
- Calculate compressor requirements.
- Determine the optimal location of various equipments as shown in Fig. 4.11 contained within the network, such as compressors, valves, regulators and the operating envelopes and discharge temperatures they can withstand.

For existing pipelines, network analysis can be introduced to:

- Assess the ability of the existing network to meet specific demand conditions as may be specified by the appropriate Gas Transporters Licence obligations.
- Calculate the optimum flow distribution/transmission requirements.
- Track the composition of gases (e.g. calorific value).
- Identify and remove bottlenecks.
- Provide a receipt of delivery volumes and support the provision of quotations for new loads.
- Provide the analytical capability to help understand and plan for the complex dynamics between pipeline operation and pipeline economics.

Additional applications include the analysis of “what if” scenarios, commonly referred to in the industry as “scenario planning”. This investigates the implications of loss of input supply or damage to equipment, such as loss of compressor power, i.e. emergency procedures. It also allows the feasibility of new contract requirements to be investigated, for example planning network reinforcement or replacement of a particular area; if reinforcement or mains replacement was required, it would be necessary to satisfy that load data and load growth rates are accurate and that there are no interruptible loads on the system. Various solutions for reinforcement or mains replacement would be considered and the least cost

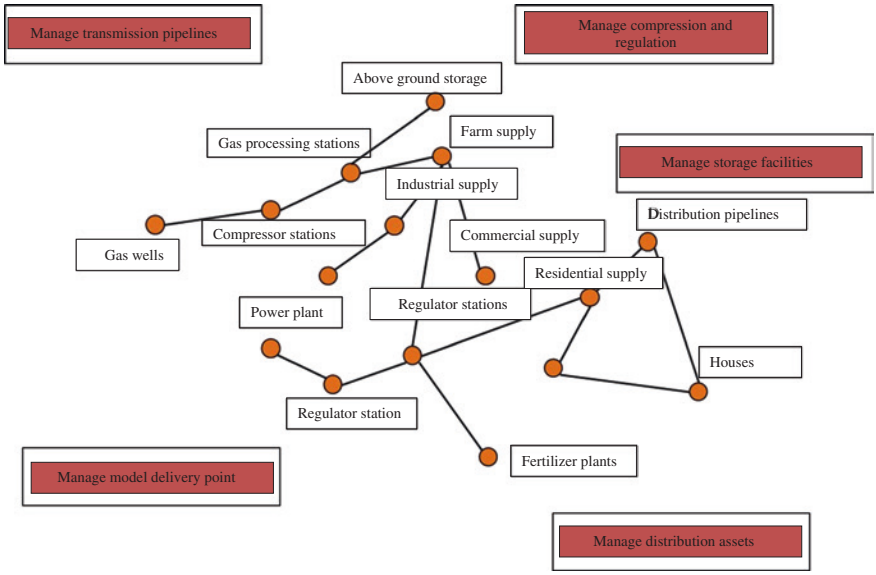


Fig. 4.11 Determination of optimal location of equipment [11]

solution likely selected, whether it be through increasing the size of a pipeline on the network or some other factors.

This process was carried out with the click of two buttons, which illustrates why network analysis is such a popular choice across various industries. Further advantages will be highlighted later in this section.

4.2.3 Objectives and Input and Output Requirements of Network Analysis

4.2.3.1 Objectives of Network Analysis

When looking to apply network analysis systems to the gas industry, there are several objectives that aim to be met. These can be summarised as:

- Determine current and future network capacity requirements, i.e. flow in all pipes.
- Determine current and future physical network capacity limitations, i.e. pressures at all pipe junctions.
- Ensure that the networks are capable of meeting maximum demand that will occur, on average, in not more than 1 winter out of 20 years.
- Determine and assess network enhancement options.
- Record the processes.

In most gas distribution companies, there is a network analysis suite of documents that highlights the policies and procedures that must be followed, such that these activities can be carried out in a defensible and consistent manner throughout the company and also the industry.

However, it is not applicable to only state that these are the main objectives of network analysis; it is also essential to fully appreciate that in order to satisfy these objectives gas distribution networks are required to know some necessary information:

- The fixed pressures at the supply points.
- The loads or off-takes from the system.
- Maximum/minimum pressures at various points in the network.

4.2.3.2 Input Requirements

Network analysis models must be developed, which represent the physical parameters of the network. These physical parameters are referred to as the input requirements of the system and comprise of pipe length, pipe diameter/condition, source and pressure regulation equipment operating characteristics and load data. Expanding on each of these parameters individually, it will allow a comprehensive representation of the input requirements of the network to be drawn. However, firstly, it is essential to touch upon the flow equation. Although this is not strictly input data, as it is already built into the network, it is still essential to comply with the network conditions. It is recognised that the form of the equation used varies depending on the preferred method of calculating the friction factor. The method chosen will undoubtedly affect the values of the constant (K) and exponent (n).

Pipe Length

From the low-pressure flow equation shown earlier, it is apparent that the pipe length (L) is directly proportional to the pressure loss, and therefore, any error will be directly reflected upon the value of K.

Since the input data are taken from the information recorded at the time of construction, for example off a transmission pipeline greater than 7 bar, then it is vital that this information is extremely accurate, as this reflects upon the information being applied to the network system.

Pipe Diameter

Standard theory concurs that the internal diameter of the constant K is proportional to $1/d^5$. Hence, this implies that if the input data for pipe diameter is inaccurate by 1 %, this will result in a 5 % error in the value of K and a 2.5 % error in the

flow calculation determined by the network analysis system. For this fundamental reason, it is vitally important that the information input is accurate. However, this is not easy and there are several difficulties that can arise that result in not so accurate information being applied.

Nominal diameters of pipes are not a good indication of the true internal diameter, significantly due to the manufacturing process. In some cases, pipeline diameters are measured to their external size and not internal, thus generating abnormalities. Therefore, in order to achieve the most accurate results, the wall thickness value must be used. Not only is the pipe diameter determined by its size but also the quality of the pipe; especially, internally is important. The internal layers of the pipe will play an imperative role in the flow of gas through a pipeline.

Source Pressure

The term "source pressure" is defined as any point where gas enters the network as shown typically in. This is generally associated with pressure reduction stations (PRS). At these points, accurate readings can be determined and related to the network analysis generated. In transmission networks, the source pressure is associated with supply terminals, compressors or regulator installations for steady state analysis. However, for unsteady and transient flow, it cannot be assumed that parameters will stay the same.

Load Data

This type of input data can only be assumed, due to the difficulty of establishing what the expected demands on the systems under peak conditions are. Gas networks look to collate metre readings on a quarterly basis from all their customers supplied from the network. This then allows them to analyse the peak flows and determine the times of year when loads are high, as previously shown in this chapter in relation to annual load factor and seasonality load demand. Moreover, to illustrate the load here Fig. 4.12 is the typical daily load factor. For example, in the winter of 2010 in the United Kingdom, especially in Scotland, a vast amount of snow has fallen and temperatures had dropped below what was anticipated, this has resulted in the gas distribution network in the United Kingdom operating at a sufficiently high load demand. This is essentially due to more and more people having to work from home, due to difficult travelling conditions and thus, heating, cooking, etc., is used more frequently throughout the day.

It is difficult for the gas distribution networks to anticipate these increases in demand; therefore, they aim to design their networks to meet the 1:20 peak 6-min demand expressed as an hourly rate. This means that the system will be designed to cope with a maximum demand that could occur, on average, in not more than 1 winter out of 20 years.

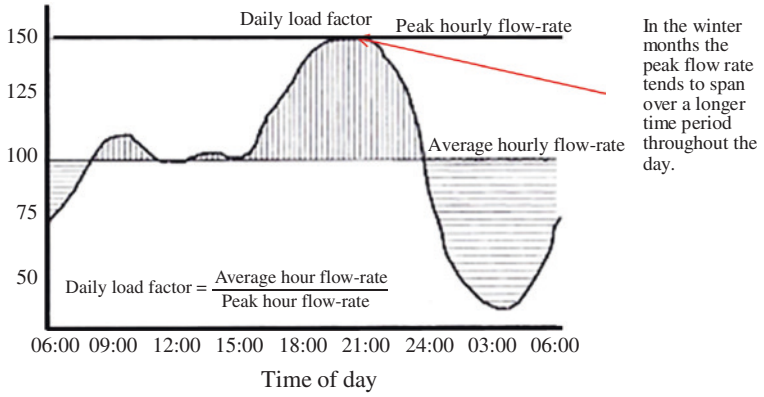


Fig. 4.12 Typical daily load factor [4, 5]

Other load data information that the gas transporters will need to assess is by the loads taken from large industrial customers. These loads tend to be consistent throughout the year and are therefore required to be accurately controlled to ensure that sufficient supply has been established to provide a constant flow under the peak demand forecasts. However, care has to be taken when allocating such loads to these customers to ensure that these high loads do not have a disrupting effect on various other aspects of the network.

Additionally, another problem with assessing the peak demand is the rate of growth and the positioning of different rates of growth accurately within the network. Some areas are declining due to depopulation; some are saturated and can be considered negligible, whereas major cities are currently undergoing rapid development and considerable rates of growth. In most cases, the gas networks make estimates and closely analyse trends in the load growth; however, as the density of these cities becomes more and more of a significant concern, this scenario will soon become unachievable and in the years to come other adequate solutions will need to be developed.

4.2.3.3 Output Requirements

The primary requirements of the output from any network analysis programme are that it should precisely signify the actual network. In order to achieve this, it is necessary to ensure that four main aspects are effectively carried out.

- Ensuring input accuracy.
- Load monitoring.
- Pressure surveys.
- Use data from pressure and flow surveys.

Expanding on each of these aspects briefly; starting with ensuring that the input is as accurate as possible. It is essential that the input information is accurately applied, in

order for the output to be regarded as a definite factor. In most cases, a database of information about each of the pipes in the network is used as the input programme. These databases will include information on new pipes that have been recently laid, plus additional information in regard to pipe diameters, material condition, excavation history, etc. All this information can be retrieved from the database.

Additionally, occasional load-monitoring exercises are undertaken to assess the accuracy of assumptions made in determining the load data. This can be achieved by measuring the flow in the pipes and evaluating the number of customers being supplied. However, this method is expensive to undertake and is not carried out on a regular basis.

Pressure surveys are another option. These can be carried out over a well-defined area, where total loads can be accurately assessed from metre readings taken in the area.

Lastly, using data from pressure and flow surveys to apply accurate efficiency factors to the pipes within the survey areas is another way in which the actual network can be represented.

4.2.4 Rules that Underpin All Network Analysis Methods

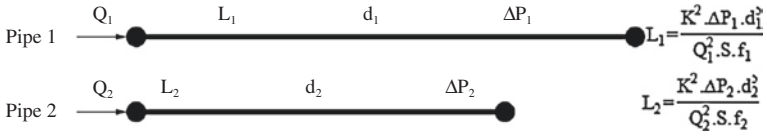
4.2.4.1 Simple Network

In this section, the underlying principles of network analysis for pipe systems starting with pipes in series and parallel and then moving onto simple loops are discussed. Specifically, the following will briefly will be described:

(i) *Equivalent length*
(ii) *Pipes in series*
(iii) *Pipes in parallel*

(i) *Equivalent Length*

Sometimes, when doing manual calculations for example, it is convenient to reduce the complexity of a problem by reducing a system of pipes in series to an equivalent length of a single pipe of uniform diameter. To illustrate the principle using the low-pressure gas flow Eq. (4.26) and applying it to two individual pipes as follows:



By assuming the flow rate in each pipe is the same ($Q_1 = Q_2$), the pressure loss down each pipe must also be equal ($\Delta P_1 = \Delta P_2$) and $f_1 = f_2$, thus pipe-2 can be expressed as an equivalent pipe-1:

$$L_2 = \left(\frac{d_2}{d_1}\right)^5 \times L_1 \tag{4.34}$$

(ii) *Pipe in Series*

Apply the principle of equivalent length to pipes in series, that is, end to end this is the simplest type of network. Pipes in series can be expressed in terms of one pipe of uniform diameter by using equivalent lengths. Take the following system:

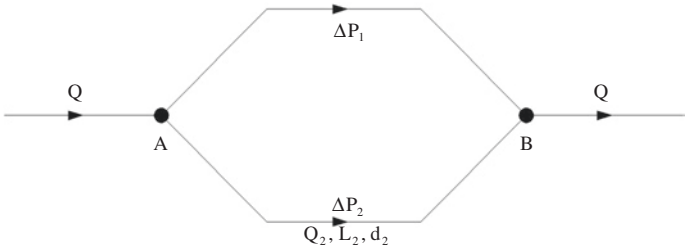


to express this system as one pipe of diameter d_{AB}, convert section BC into an equivalent length of diameter d_{AB}. For this to be valid, the flow rate in section BC must be the same as the flow rate in section AB, that is no off-take at *B*, and the overall pressure loss must remain unchanged. Thus, the total length of the system expressed in terms of d_{AB} is:

$$\text{Equivalent Length} = L_{AB} + \left(\frac{d_{AB}}{d_{BC}}\right)^5 \times L_{BC} \tag{4.35}$$

(iii) *Pipes in Parallel*

Although still simple in concept, parallel pipes are more difficult to analyse than pipes in series. We often need to know the ratio of flows in each pipe, and these may have different lengths or diameters. But first, let us consider simplifying the network to a single pipe of an equivalent diameter. This must be able to pass the same quantity of gas as the two parallel pipes with the same overall pressure loss. Consider the following example:



Applying again the LP Eq. (4.26) to the network and the two parallel pipes are joined at A and B, the pressures at these two points are common to both pipes which means that $\Delta P_1 = \Delta P_2$. Similarly, $L1 = L2 = L$ and $f1 = f2 = f$. Furthermore, the flow into the network Q must be equal to the sum of the flows in each of the parallel pipes, thus if we replace the two parallel pipes with an equivalent single pipe, the low-pressure equation for this pipe is:

$$Q = K\sqrt{\frac{\Delta p \cdot d_e^5}{f \cdot S \cdot L}} \tag{4.36}$$

where d_e is the equivalent diameter that allows the total flow Q to flow from A to B with the same overall pressure loss as the original system. Since this new pipe is equivalent to the original parallel pipe system then:

$$Q = K\sqrt{\frac{\Delta p \cdot d_e^5}{f \cdot S \cdot L}} = K\sqrt{\frac{\Delta p}{f \cdot S \cdot L}}\left(\sqrt{d_1^5}\right) + \left(\sqrt{d_2^5}\right) \tag{4.37}$$

We can extend Eq. 4.38 to account for any number of parallel pipes as follows:

$$d_e = (d_1^{2.5} + d_2^{2.5} + d_3^{2.5} + \cdots d_n^{2.5})^{0.4} \tag{4.38}$$

where

d_e equivalent diameter of a single pipe
d_1 diameter of the first parallel pipe
d_2 diameter of the second parallel pipe
d_n diameter of the nth parallel pipe

4.2.4.2 Network Structures

Before the underpinning rules of network analysis can be discussed, it is necessary to generate a level of understanding into the structure of a gas network. Typically, industry networks consist of numerous combinations of trees and loops:

The term "tree" can be defined as a series of pipes in which each pipe is connected to a source by only one route. All the pipes can be solved analytically as a series of individual pipes. Figure 4.13 illustrates a typical tree network that would supply each pipe or node in the network.

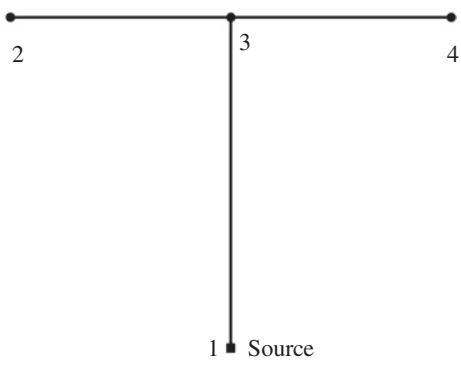


Fig. 4.13 Simple tree network

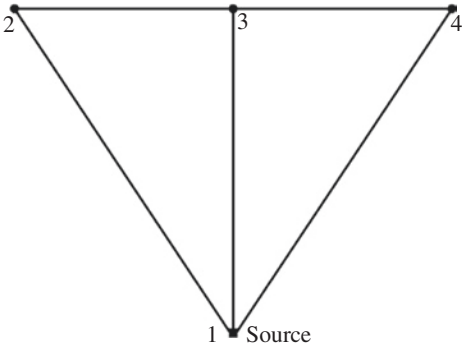


Fig. 4.14 Simple network consisting entirely of loops

A "loop" can be defined as a closed path that begins and ends at the same point and therefore has a pressure drop of zero. Figure 4.14 presents a basic loop network structure.

Some network analysis techniques require the network loops to be defined before the analysis can proceed. Hence, it is important to know where the loops are in the network structure. Figure 4.14 shows three loops; 1–2–3–1, 1–3–4–1 and 1–2–3–4–1. Any two of these loops would identify all the pipes in the network, such a set of loops is known as an independent set.

Essentially, networks that have no loops can be considered tree networks and are relatively easy to deal with. However, in most cases, loops are present and the network solution can be very complicated. Some network analysis techniques, such as the Hardy Cross method, requires the network loops to be determined prior to conducting the analysis. Therefore, it is important to know where the loops are in the network structure.

Whether it is a tree or loop network, they both have an essential role to play in the success of network analysis.

4.2.4.3 Kirchhoff's Laws

As previously mentioned, loops and nodes have an important role to play in network analysis. The flow of gas through nodes and loops is governed by a set of rules known as Kirchhoff's Laws. Gustav Robert Kirchhoff, a Prussian physicist, first stated the analogous rules for electricity networks.

Kirchhoff's First Law

The first law states that:

- The volume flow entering a given node is equal to volume of flow leaving that node in a specified period of time.

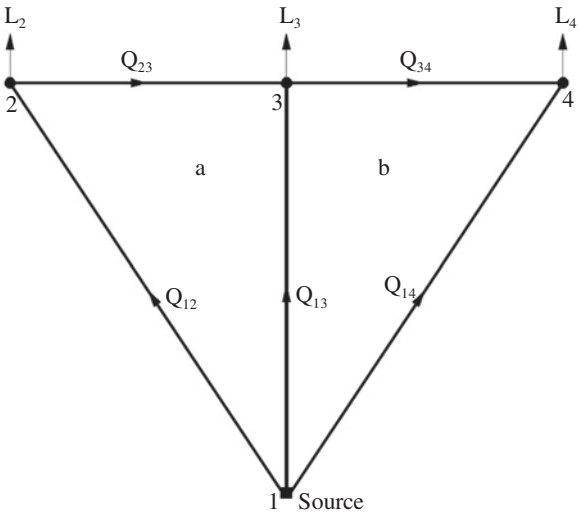


Fig. 4.15 Simple network with load nodes and pipe flows

From this, we can state that the algebraic sum of all flows entering and leaving a given node is equal to zero as shown:

$$\sum Q = 0 \tag{4.39}$$

This law is the basis for all nodal equations. From Fig. 4.14, it is evident that a network consists of nodes and flows. Using Fig. 4.15 below, it is possible to produce a set of equations that conveys Kirchhoff's first law which states the algebraic sum of flows entering and leaving a given node is equal to zero.

From the simple network above, the following equations can be derived for each node, as shown in Table 4.4.

There is one other additional equation based on the first law. The rule that the sum of the inflows must equal the sum of the outflows also applies to the complete network as well as individual nodes. Therefore,

$$\text{Source Flow} = L_2 + L_3 + L_4 \tag{4.40}$$

In summary of the first law, it has been established that five equations exist and along with these, five unknown pipe flows arise. Unfortunately, these five equations are not independent, which means it is not possible to solve the entire network using Kirchhoff's first law; therefore, it is necessary that the second law is also used.

Table 4.4 Nodal equations using simple network

Node 1	Source flow	$=Q_{12} + Q_{13} + Q_{14}$
Node 2	Q_{12}	$=L_2 + Q_{23}$
Node 3	$Q_{23} + Q_{13}$	$=L_3 + Q_{34}$
Node 4	$Q_{34} + Q_{14}$	$=L_4$

Kirchhoff's Second Law

The basis of the loop equation is Kirchhoff's second law. This law states that:

- At any given instance in time, the pressure difference between any two nodes in a network is fixed and is the same for every flow path between those two nodes.

From this equation, we can state that the algebraic sum of the pressure drops around any given loop is equal to zero:

$$\sum \Delta P = 0 \tag{4.41}$$

Therefore, because only one pressure value can exist at a given node in a network, this law also applies to the difference in pressures squared, as shown:

$$\sum \Delta P^2 = 0. \tag{4.42}$$

Unless both laws are satisfied, a network will not be balanced.

Underpinning Rules

In review of the foregoing sections, it is evident that two simple rules underpin all network analysis systems. These are: (i) what goes in must come out and (ii) the pressure loss between any two nodes provides the same value whatever way the flow travels.

4.2.4.4 Comparison of Methods

Hardy Cross (Mesh) Method

Network analysis based on the Hardy Cross Method uses Kirchhoff's first and second law (Eqs. 4.41 and 4.42, respectively) for obtaining the value of flow rate and pressure in each segment. If the equations have not satisfied Kirchhoff's law, a series of iterations will be done using. Use Newton's method (for non-linear equations) until the values for pipe flows within the required tolerance are reached. Computer solutions using the Hardy Cross method were developed originally because of the simplicity and convenience of the method, particularly when dealing with large steady state networks such as low-pressure distribution systems. Because each loop is considered individually, only the information relevant to one particular loop need be resident in the computer's memory at any time. For early computer systems, this was a considerable advantage compared to the substantial memory requirements for simultaneous solution techniques. The Hardy Cross method cannot be used for transient analysis because Kirchhoff's laws, on

which Hardy Cross is based, are not applicable. Transient problems are solved by a direct solution of the simultaneous equations arising from the dynamic models for pipelines, compressors and other network components. Transmission systems, although physically larger than distribution systems, do not contain very many pipes or loops; therefore, the memory requirements for the network are not as great, but the computational requirements are more complex.

The advantages of the Hardy Cross method are:

- Convergence is achieved rapidly because of the quadratic equation
- The convergence is insensitive to the starting value
- The disadvantages of the method, however, are:
- A set of loops is required to be defined
- Time consuming because each loop must be considered individually and sequentially

The Nodal Method

Generally, the nodal method is similar to the Hardy Cross method except this method does not require the network loops to be defined, a process which can take up a significant proportion of the overall solution time of a computer. However, the disadvantage of this method is the convergence heavily depends on the initial estimate for the nodal pressures.

The advantages of the method are:

- The method does not rely on interconnected loops within the network and can be applied to any network where the requirements for fixed pressure nodes and load nodes are met.

Moreover, the disadvantages are:

- Square root function requires a large number of iterations to reach a practical solution. In addition, the divergence and oscillation are the risk of using this method.
- Time consuming because each node is considered individually and sequentially throughout the network.

Newton–Raphson Mesh Method

Unlike one Hardy Cross method in which each iteration on a loop takes no account of the effect of that loop on the other loops in the network and consequently contains flow correction for loops, Newton–Raphson Mesh method has a simultaneous solution that interacts one loop with another to be modelled. The advantages of using this method are the reduction of iterations required to solve the network.

Newton–Raphson Nodal Method

The Newton–Raphson method enables the simultaneous solution of all the equations derived from the application of Kirchhoff's first law at each load node (pressure defined nodes, such as sources, are not included). The matrix equation is constructed from an equation where n is the number of load nodes in the network. As a result, the matrix contains many more zero elements, allowing a greater degree of compaction. The advantages, as with the Newton–Raphson mesh method, are the improvement of solution speed by reducing the number of iterations required for convergence of the network.

Hybrid Method

This is a combination of Newton–Raphson mesh and nodal methods to get the advantages of the mesh and nodal methods for achieving a solution. Unlike Hardy Cross method, this method can only be achieved by using matrix methods.

In fact, the hybrid method is a variant of the Newton Nodal method. In the original Newton Nodal method, flow variables were eliminated before the equations were linearised; while in the hybrid method, the equations are linearised before the flow variables are eliminated and the problems are avoided.

4.2.4.5 Advantages and Disadvantages of Using Simulation Modelling

This section of the report examines the various simulation modelling software systems that are commonly used throughout the gas industry to analyse a network. There are numerous advantages of using network analysis to simulate gas flow and pressure within a natural gas network. However, with these advantages fall several disadvantages. In this particular section, a flavour of the advantages and disadvantages that are associated with the various network analysis systems shall be determined. This will allow a comprehensive understanding into why network analysis is so popular not only in the gas industry but also in many other industries across the world.

In the United Kingdom, SGN adopts the SynerGee system for network pressures below 7 bar and uses Graphical Falcon for networks operating above 7 bar (Fig. 4.16).

Advantages of Simulation Modelling

There are many reasons why, industries, especially gas are attracted to simulation modelling on their networks. The biggest of these advantages is likely cost. Designing, building, testing, redesigning, rebuilding, retesting of a gas pipeline network is an unrealistic option and would also be extremely expensive. Simulation modelling allows the building and rebuilding phases to be removed from the loop, therefore generating faster solutions.

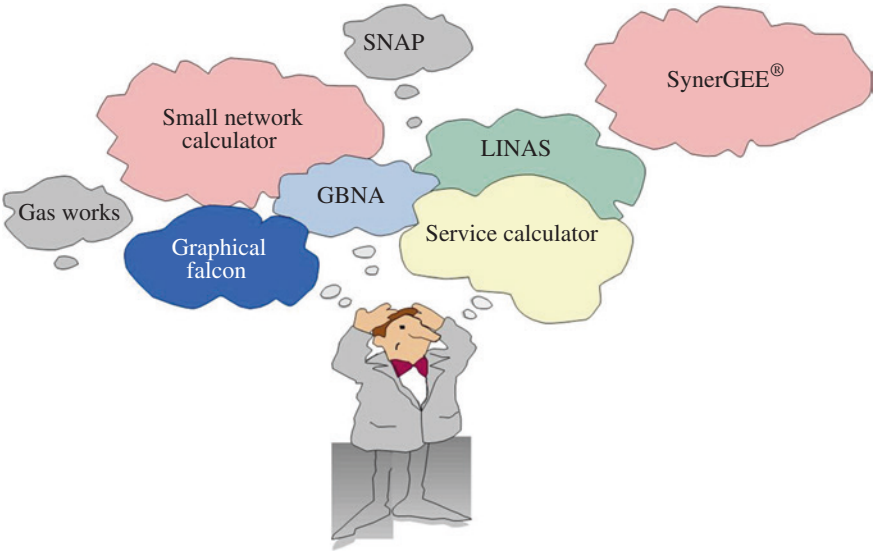


Fig. 4.16 Common network analysis systems [3]

These solutions will also significantly benefit the capital costs associated with operating the network. The modelling will allow an accurate assessment of the pipe diameter requirements necessary to cope with the pressure criteria of the pipe to be easily established. This is a huge advantage and will reduce material costs when constructing the necessary pipeline.

Another evident benefit is the reduction in time that the engineer will spend designing the network [6, 7]. This is significantly due to the vast array of detail that a network analysis system can provide the engineer with. For example, if the equations and calculations mentioned previously were to be determined manually, this would take days, however with simulation modelling an engineer can produce a whole network in a matter of hours because the system can generate all the necessary results at the click of a button.

"What If" scenarios as mentioned in Sect. 4.2 are also easily resolved using simulation modelling, which is another massive advantage for a gas distribution network. This allows them to assess the effectiveness of the network in the event of an occurrence happening. The example presented provides a brief insight into what happens to the existing network when a pipe is removed. From this example, it allows the advantages of simulation modelling to be demonstrated further.

Disadvantages of Simulation Modelling

Due to the nature of some of the underlying concepts of pipeline modelling, network analysis can never be absolutely accurate. The relationship between flow and

- Altitude of the pipeline system.
- Thermal properties of the burial material.
- Burial depth of the pipeline, as shown in Figure 4.17

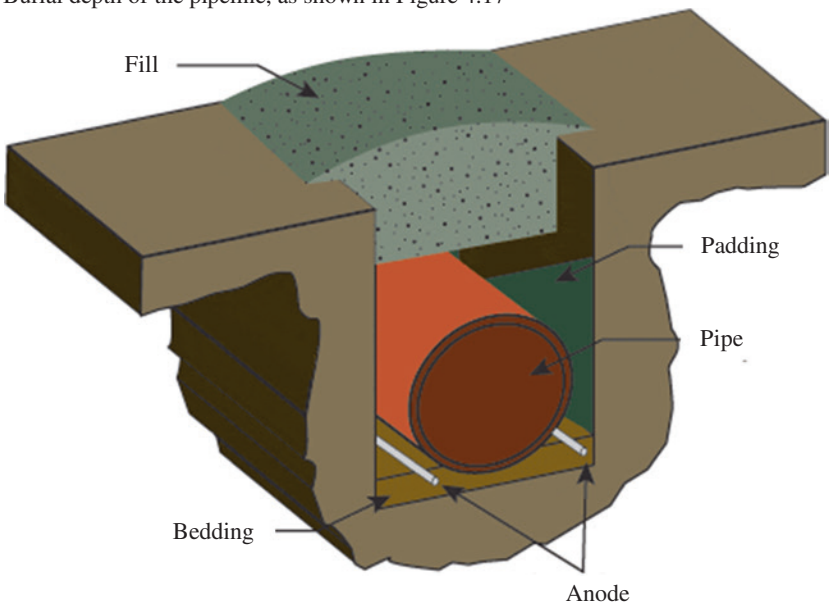


Fig. 4.17 Burial depth and materials used [10]

pressure can be encapsulated in an equation. However, various other parameters that are less obvious tend to be forgotten about, such as:

- Internal roughness and/or efficiency of the pipeline.
- Temperature of the gas.
- Compressibility of the gas.
- Altitude of the pipeline system.
- Thermal properties of the burial material.
- Burial depth of the pipeline, as shown in Fig. 4.17.

Most of these elements cannot be reasonably or accurately known. For example, the thermal properties of the burial material might rely on knowing the sand content and water content of the soil: the former of these can vary greatly over a short distance, and the latter may need correction for seasonal variation [8, 9].

Additionally, there is no guarantee that all the data and input received are correct, for example the pipe may not be PE 80; in fact, it may be PE 100 as shown typically in Fig. 4.18. Also the quality of workmanship may not be sufficient or the positioning of the pipeline may be incorrect; however, when using network analysis, it implies that the network is to the highest standard and runs a model regardless of pipe conditions, extra bends, etc.

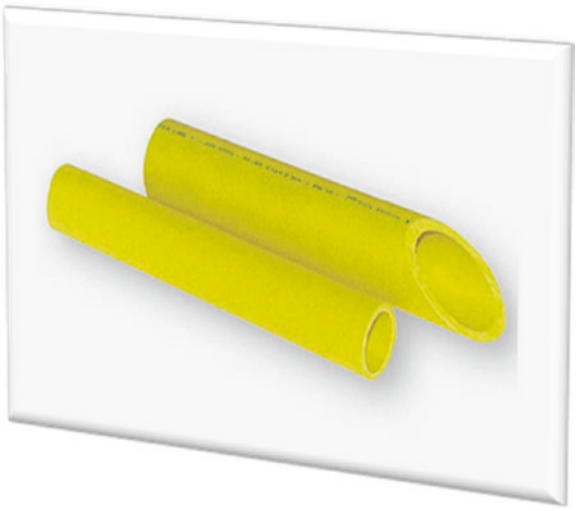


Fig. 4.18 Differences between PE pipes [12]

Hence, this means that all flow equations must be derived empirically, which conveys that any particular flow equation will be accurate only within a special context (e.g. a range of pressures). Thus, stating that network analysis is not a 100 % accurate method, although many gas networks make use of the information in this way.

4.3 Principles of Transient Flow

Previous sections were concerned with steady state flow. However, because gas is compressible, to model accurately the flows that take place in gas pipelines or during a pipe break, the concept of unsteady state or transient flow should be applied. The difference between steady state and unsteady state flow is where the flow parameters, such as pressure and mass flow rate, vary with time; so transient analysis is the analysis of the time-dependent variations in these parameters (the transients). For example, in a pipe break, the pressurised gas is released into the atmosphere in an uncontrolled manner and the pipeline will not empty rapidly because of the much larger quantity of compressed gas in the pipe. Clearly, steady state equations are inappropriate for the analysis of pipe breaks and they should not be used to determine the rate of gas leakage. The quantity of gas escaping depends mainly on the absolute pressure of the gas in the pipe and the resistance to flow created by the shape and size of the fracture area. If the gas supply into the pipe is maintained, the gas will escape at a rate determined by the gas pressure. If it is possible to isolate the fractured section, the pressure will decline exponentially with the consequent reduction in the rate of gas escaping.

Unlike a water pipeline for example, where a fracture results in an immediate loss of pressure, a gas pipeline contains compressed gas, so the dynamics of the decompression are much slower; this fact has important consequences for pipeline design, which are embodied in IGE/TD/1, the Institution of Gas Engineers recommendations for gas pipelines. The main difference between unsteady state flow and steady state flow is the time period between input and extraction of fluid from a system. The time difference between the input and extraction of fluid introduces the concept of storage, and in gas pipeline systems, this storage is used to smooth out variations in the flow at the demand end of the pipeline so that the inlet flow does not change significantly. The time difference allows gas to be extracted from the pipeline by reducing the pressure (rather than introducing more gas at the input) in the same way that water was drawn off from a barrel. This feature allows pipeline operators to manage the daily demand variations without the need to alter the rate at which they buy gas from the offshore producers. Let us have a look at this feature in more detail.

Figure 4.19 shows two steady state pressure profiles for a pipeline. Profile (1) represents the conditions for a steady state flow rate Q_1 and profile (2) represents the conditions for a new steady flow state flow rate Q_2.

The area under curve (2) is less than the area under curve (1) and since these areas are representative of the volume of gas in the pipeline, the line-pack volume, it follows that a volume of gas, represented by the hatched area, has been

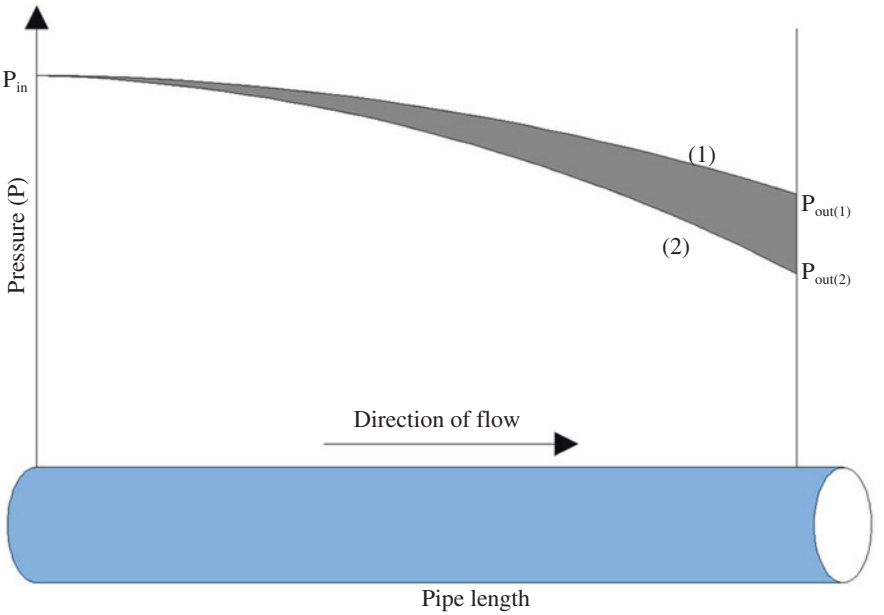


Fig. 4.19 Line-pack storage resulting from two steady state pressure profiles in a pipeline designed for gas transmission only

withdrawn from the pipeline, the line-pack storage volume. This gas has been used by the customers at the end of the pipeline causing the outlet pressure to fall from Pout(l) to Pout(2), this means that a finite period of time has elapsed between profile (1) and profile (2). The customers have effectively been using stored gas between time (1) and time (2). In fact, between these times, the rate of demand could have been higher than Q_2 for a limited period because the higher demand does not need to be transmitted along the pipeline, it is only drawn off at the downstream end.

The line-pack storage volume, represented by the hatched area, is much larger than that shown in Fig. 4.19. Taking the line-pack storage a stage further is by designing the pipeline specifically for this purpose. To do this, the pipeline should be designed and constructed to withstand a higher pressure than that required purely for gas transmission.

Profiles (1) and (2) in Fig. 4.19 are exactly the same as those in Fig. 4.18. But what is the main difference between the two diagrams?

The difference is that P_{in}(l) is higher than the original Pin creating more storage between the two profiles. To release this storage at the required time, the inlet flow rate can be restricted to force the inlet pressure lower, but in practice, it is more effective to set a constant flow rate at the inlet (the daily average) and allow the variations in demand at the outlet to absorb the stored gas. The principle of line-pack storage has been introduced, but how is the storage volume calculated? This is described in the next section.

4.3.1 Calculation of Line-pack Storage

To calculate the volume of usable line-pack storage, the gas volume should be found which is represented by the areas under each of the profiles shown in Fig. 4.20 and subtract one from the other. Figure 4.21 shows a pipeline operating under steady state conditions. By considering an elemental volume of gas, we can integrate the pressure profile along the length of the pipeline to find the total volume of gas under the profile.

The volume of the incremental section is $A.\mathrm{d}l$ so, from the real gas equation of state, the volume of gas in the section at standard conditions ($\mathrm{d}V_s$) is given by:

$$\mathrm{d}V_s = \frac{A \cdot \mathrm{d}l \cdot p \cdot T_s}{Z \cdot p_s \cdot T} \tag{4.43}$$

The pressure gradient in the pipe is Ap^2/L, so the pressure (p) at array point distance 1 from the start of the pipeline is given by:

$$p = \left(p_1^2 - \frac{1}{L}\Delta p^2\right)^{0.5} \tag{4.44}$$

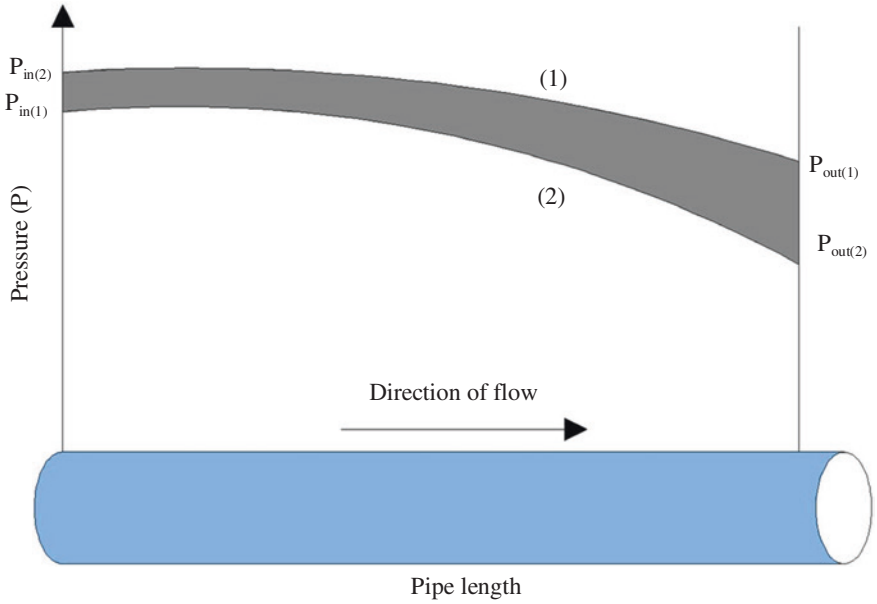


Fig. 4.20 Line-pack storage resulting from two steady state pressure profiles in a pipeline designed for gas transmission and storage

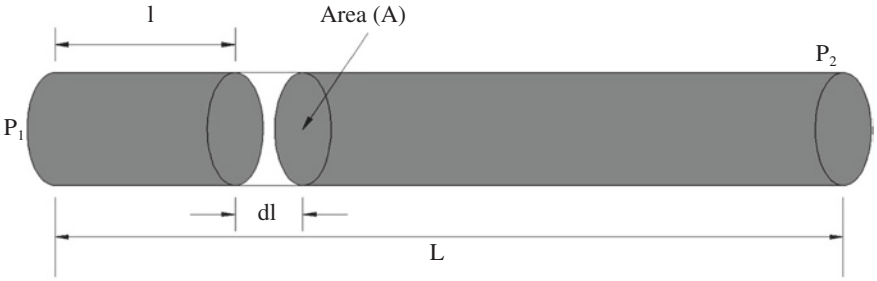


Fig. 4.21 Pipeline operating under steady state condition

Substituting for p in Eq. 4.47 which, when integrated and rearranged, gives

$$V_s = \frac{A \cdot L \cdot T_s}{Z \cdot T \cdot p_s} \frac{2}{3} \left(\frac{p_1^3 - p_2^3}{p_1^2 - p_2^2} \right) \tag{4.45}$$

The volume of gas in the pipeline (V_s) is the physical volume of the pipeline ($A.L$) corrected to standard conditions using the average pressure (P_m). This is the same value of P_m for dealing with the compressibility factor. Thus:

$$\frac{A \cdot L \cdot T_s \cdot p_m}{T \cdot p_s \cdot Z_m} \tag{4.46}$$

Or in terms of diameter, with d expressed in mm, the difference between the volumes at the maximum and minimum conditions can be expressed for profiles (1) and (2) as:

$$\Delta V_s = \frac{0.7854}{10^6} \frac{T_s}{p_s} \frac{L \cdot d^2}{T} \left[\left(\frac{p_m}{Z_m} \right)_1 - \left(\frac{p_m}{Z_m} \right)_2 \right] \tag{4.47}$$

where AV_s is the volume of line-pack storage expressed at the standard condition T_s and P_s and Z_m (1 and 2) are the maximum and minimum flow rates.

4.3.2 Dynamics of Pipelines

In a water pipeline, the dynamics are very fast because the fluid is incompressible and so changes in flowrate result in very rapid changes throughout the system, often resulting in "water hammer". The equivalent of water hammer is not noticeable in HP gas pipelines because the gas, being compressible, absorbs the changes in p over a period of time. However, it can be seen in some LP situations because the gas behaves more like an incompressible fluid, e.g. where large load changes occur in a small volume of pipe.

Now look again at Fig. 4.18. This represents a pipeline where the inlet pressure is controlled at a constant value (say, 70 bar abs), so changes in pressure at the outlet, caused by changes in the flow rate, do not affect the value of the pressure at the inlet. All the pressure variations have been absorbed by the gas over the pipeline length. However, a change in the demand flow rate from one steady state condition (Q_{mm}) to another steady state condition (Q_{max}) would lead to a change in the inlet flow rate from Q_{mm} to Q_{max}, the time taken for this change to be apparent at the inlet is determined by the rate at which the change can be propagated through the gas from the outlet to the inlet. This, of course, is the dependent on the volume of line pack that must be used up. The length of time that it takes for flow rate change to propagate through the pipeline can be estimated by the difference in the flow rates using the following equation:

$$\text{Approximate transient time} = \frac{\text{Volume of linepack storage } (\Delta V_s)}{\text{Change in flowrate } (\Delta Q_s)} \tag{4.48}$$

Figure 4.20 represents a pipeline designed for line-pack storage as well as transmission where the inlet flow rate is held at a constant value while the downstream outflow varies according to the demand. In this case, we are interested in the pressure transients, that is the time taken for a change in pressure at one end to be observed at the other end and as with flow rate changes, we are dealing with quite long transient times.

Since both pressure and flow transients take a finite length of time to propagate along the pipeline they must travel with a certain velocity, so we can think in terms of a "wave" travelling through the gas.

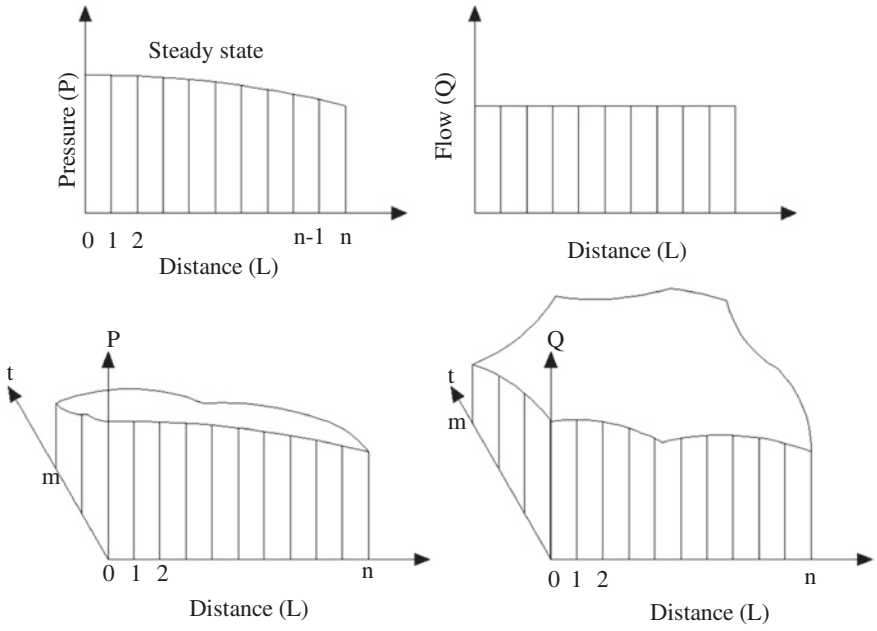


Fig. 4.22 Comparisons of steady state and unsteady state flow

Figure 4.22 illustrates the difference in concept between steady and unsteady state flow resulting from the inclusion of the time dimension. The steady state case is two dimensional in pressure and distance or flow and distance. The curve representing the pressure loss along the pipeline length can be described by n values and the flow, being constant, is described by one value along the whole length.

In the unsteady flow situation, Fig. 4.22, each of the n pressure values requires a further m values to describe the change in pressure with time; therefore, a total of $m \times n$ pressure values is required and since the flow also varies with time, $m \times n$ flow values are also required. This represents a great increase in complexity and so careful consideration must be given to the problem to be solved.

The disturbances introduced into the gas at one end of a pipeline will take a finite length of time to reach the other end; the length of time depends on the velocity at which the "wave" travels, which in turn depends on the type of disturbance, the gas pressure and the characteristics of the pipeline t_1.

Here is an example of how a pressure transient might travel along a pipeline. The response of a pipeline *to* a step change in pressure at its inlet is shown in Fig. 4.23. At the inlet conditions ($t = 0$), the flow is steady state and is represented by the pressure profile from P (0, 0) to P (0, L) at the outlet. At time t_1, the inlet pressure rises from P (0, 0) to P (1, 0) but the pressure at the outlet P (1, L) is still at the original outlet pressure. At each increase in time, a higher pressure is seen at increasing distances along the pipe until at time t_{10} the pressure transient has reached the outlet of the pipeline, after which time the flow settles back to steady

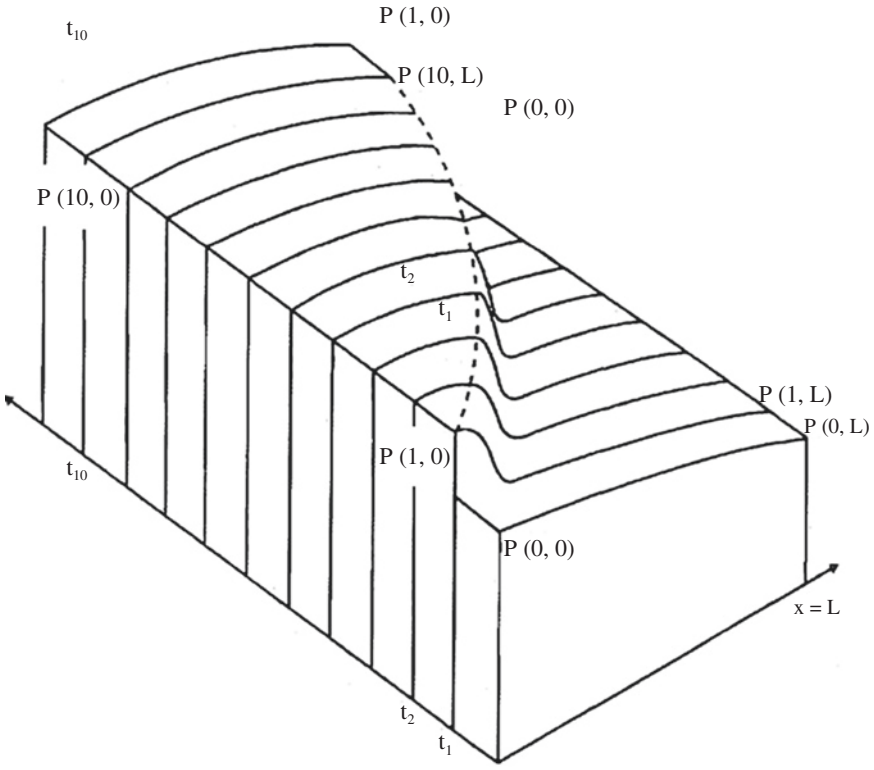


Fig. 4.23 The response of a pipeline to a step change in the inlet pressure

state. The average velocity of the pressure transient is given by the length of the pipeline (L) divided by the travel time, which in this case is $(t_{10} - t_1)$. The velocity $L/t_{10} - t_1$) is represented by the dotted line from P (1, 0) to P (10, L).

Figure 4.23 from above shows that the velocity of the transient in two dimensions; time and distance. Figure 4.24 shows the transient progressing from t_1 to t_{1o} over the length of the pipeline. Notice that the path is curved, which indicates that the velocity is not constant; however, for the purposes of discussion here, it is assumed constant.

In practice of course, there is no single transient as we have described but many different transients superimposed on each other. These are caused by the continual changes in demand and operating conditions which occur throughout each day and so transients are being generated all the time and at differing rates. When a change in flowrate or pressure is imposed on a gas pipeline, the disturbance propagates through the gas at the local velocity of sound; this is the maximum velocity at which transients can travel. However, due to pipe wall friction, the disturbances are smeared out over a large space and time interval as they travel along the pipe. For a large disturbance, the main effect at a remote point is seen long after the

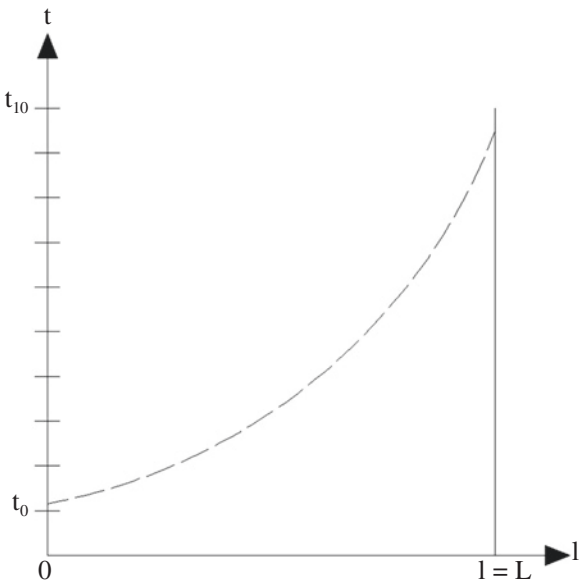


Fig. 4.24 Progress of transient between t_1 and t_{10}

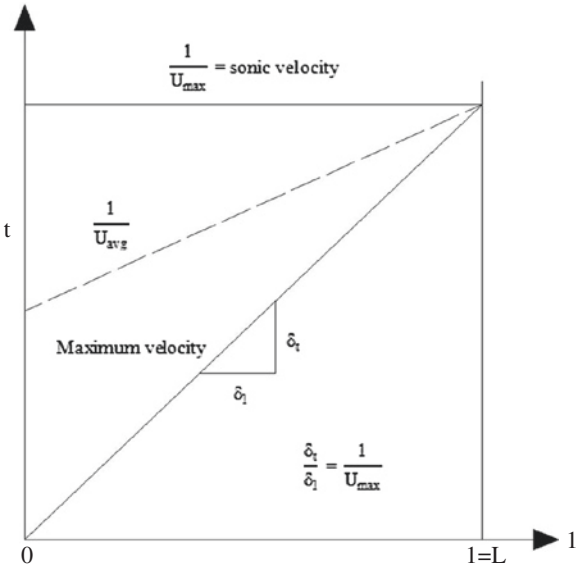


Fig. 4.25 Velocities associated with transient

original wave front has passed. This delayed effect is referred to as a "slow transient" and is the main interest for modelling pipelines. Figure 4.25 shows the range of velocities to be considered.

Because unsteady state flow problems are so complex, we need to optimise the method of solution; this means selecting a solution procedure which only deals

with the transients of interest. So transient analysis programmes are designed to model the pressure and flow transients that result from the action of friction rather than the disturbances moving at sonic velocity.

Here are some examples that are not suitable for analysis using a pipeline model.

- The velocity of gas escaping from a pipe break is equal to the sonic velocity, which is outside the range of transients covered by a pipeline model. Pipe breaks must therefore be modelled using a programme specifically designed for this purpose.
- The dynamics of control valves are determined by the valve characteristics, the method of valve control and the pressure of the gas. In order to achieve accurate control of the gas pressure or flow rate the valves are designed to have a fast response time, this allows the valve to move quickly to correct any variation in the measured pressure or flow rate. The dynamics of control valves are therefore much faster than pipeline dynamics and outside the range of pipeline models. Special modules using different solution techniques must be built into a transient analysis programme if required to simulate the interaction of valves and pipelines.
- As with control valves the dynamics of compressor stations are much faster than pipelines, they are also very complex, so special solution techniques must also be used to model the performance of compressors and their interaction with pipelines.

Because the dynamics of control valves and compressors are so fast compared to those of pipelines, a reasonable approximation can be made by assuming that the output from these devices is calculated on steady state basis. This simplifies the modelling considerably, but you should be aware of the limitations of such an approach.

4.4 Concluding Remarks

In the preceding chapter, a thorough evaluation of the principles of steady and unsteady state gas flow and network analysis were provided. Understanding gas flow and network analysis that are the most comprehensive system available today provides various industries such as the gas industry with a suitable simulation model of their ever evolving and continuously functional distribution and transmission network. There are a significant number of application uses for network analysis along with a whole list of advantages as to why the gas industry adopt such a system, one of the main ones being to determine safe operational flow and pressure through their pipes.

The mathematics behind all systems appears to be well specified and documented and allows various sets of calculations to be established with several clicks of a mouse. However, it is worth mentioning that network analysis cannot

be deemed to be 100 % accurate; however, it is the best solution available and provides numerous valuable pieces of information. To conclude, network analysis is a fundamental aspect in the success of transporting gas to millions of customers worldwide. Network analysis systems will continually be improved and this will only assist the gas industry in providing a consistent supply of natural gas throughout the United Kingdom and across the world for many years to come.

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Chapter 5
Natural Gas Measurement Systems

5.1 Temperature and Heat Flux

5.1.1 Introduction

Throughout the years, the gas industry has witnessed numerous different measurement techniques and experimental investigations relating to natural gas engineering, consequently making it extremely difficult to keep up to speed with the evolving technology that is being implemented across the world.

In the gas industry, the measurements of temperature, pressure, gas flow and gas quality are some of the key parameters that are considered. However, due to the copious amount of measurement techniques that are available, it is impractical to detail each of these in this chapter. For this reason, the rationale is to capture the key points from the vast array of measurement techniques being used across the gas industry and present them in a clear and concise manner. In this chapter, attempt has been made to provide the knowledge necessary to understand the operational principles of instruments that are available in the gas industry to monitor parameters such as temperature, pressure, gas flow and gas quality. Contained within each of these sections will be a brief overview of the parameter being measured, the various types of instruments that are available to use for each parameter, along with the specific options which exist. At the end of each related section, all the instruments that have been highlighted are then shown in tabulated format which includes a brief operations of feature of the instrument, the range, the accuracy, advantages and disadvantages and application of each of the devices. This will provide the reader with sufficient data on the various instruments that can be used to measure the parameter being discussed. The conclusions drawn from each of the sections will then be presented at the end of the chapter. References are also provided with regard to design features of each instrument for further consultation.

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Starting with temperature and heat flux in this section, one can see that, in general, it is not easy to define temperature in a practical way; therefore, it tends to be expressed using words, such as hot and cold. When any such object has a low temperature, it is referred to as cold and when a particular temperature is higher, it is referred to as hot. However, it is not practical to use sensations and notions as a means of measuring temperature, this is not scientifically correct. Fortunately, several properties of materials act in a consistent manner with temperature, hence forming the basis for accurate measurement of temperature.

When a hot body is brought into contact with a cold body, heat is transferred from the hot body to the cold body until both these bodies have equal temperature. At this point, it is referred to as thermal equilibrium. The zeroth law of thermodynamics states that if two bodies are in thermal equilibrium with a third body, they are in thermal equilibrium with each other. This zeroth law serves as a basis for the validity of temperature measurement.

Temperature scales aim at using a common basis for temperature measurement. Throughout the world, there are many scales that can be used; the most common is known as Celsius. However, the Kelvin scale is the temperature standard most commonly used in engineering principles and has the same incremental scaling as the Celsius scale, but fixes its origin at absolute zero (0 K = −273.15 °C). All temperature scales are based on some easily reproducible states, such as the freezing and boiling points of water. The Celsius scale adopts 0 °C as the freezing point of water and 100 °C as boiling point, whereas the Fahrenheit scale makes use of a historical scale at which water freezes at 32 °F and boils at 212 °F. It is also necessary to understand the relationship between temperature and absolute temperature, shown below. In this case, there are no negative temperatures, and the lowest attainable temperature is absolute zero. The absolute temperature scale in the SI system is the Kelvin scale as shown in Fig. 5.1 and can be expressed as:

$$\text{Temperature (Kelvin)} = \text{Temperature (Celsius)} + 273.15$$

When zero is read on a temperature scale, it does not mean the point at which no heat is contained in the body being measured for Fahrenheit and Celsius scales. The presence of heat in any mass causes the molecules to vibrate. The more heat there is, the greater the oscillation becomes. When all molecular motion ceases in a body, this is the point at which there is no heat remaining in that body. This occurs at −273.4° on the Celsius scale and −460.16° on the Fahrenheit scale.

5.1.2 Thermometry Types

5.1.2.1 Expansion Thermometer

In most forms of temperature measurement, the variation of a substance with temperature tends to be the method commonly used. Expansion thermometers are divided into four expansion classes: solids, liquids, gases and change of state. In each of

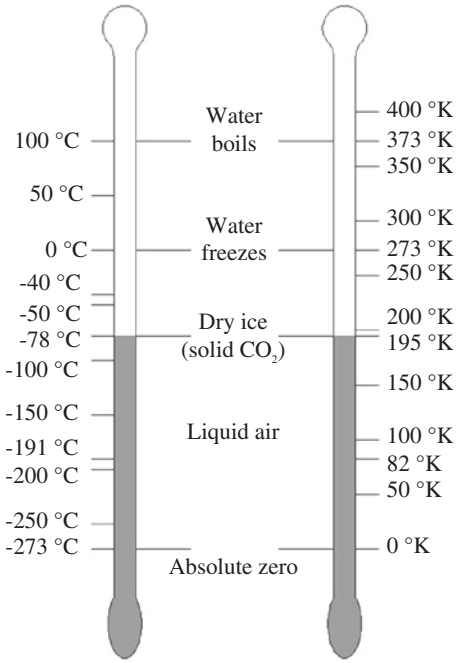


Fig. 5.1 The temperature scale in SI unit

these cases, the substances have a constant recognisable change when temperature is applied to them. In this particular section, each of the classes being discussed either increase or decrease in size, this is known as expansion or contraction. Therefore, taking this into consideration, it is important to gain a thorough understanding of each of the expansion thermometer instruments that can be used under each particular class.

Solid Expansion

When two metallic strips with different expansion coefficients are applied under the same temperature, a differential expansion will occur on the strips. The increase in length of any side of a solid will depend on its original length, the rise in temperature and the coefficient of linear expansion.

The most practical instrument that is used for this particular class of expansion thermometer is known as bimetal strip. Figure 5.2 shows the process of two metals bonded together with different coefficients of expansion.

Liquid Expansion

In this class of expansion thermometers, an instrument referred to as liquid in metal thermometers tends to be used; however, liquid in glass thermometers is also extremely popular. See previous Table 5.1 for further information on this

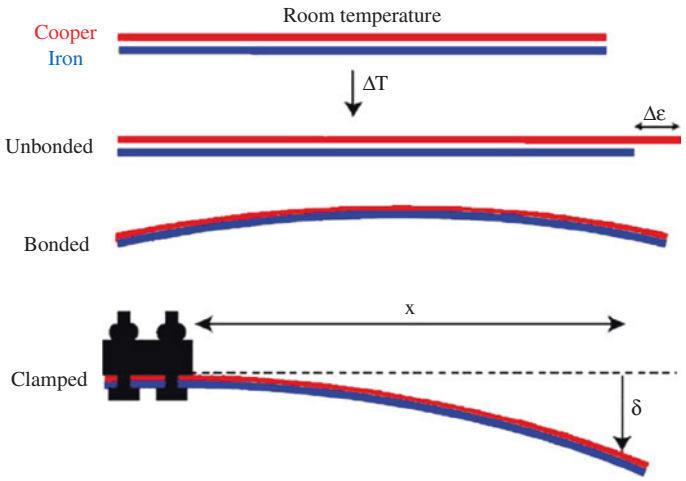


Fig. 5.2 Coefficients of expansion of bimetal strips

Table 5.1 Temperature ranges of thermometers

Liquid	Temperature range
Mercury	−38 to +600
Helium	−55 to +600
Alcohol	−80 to +50
Toluene	−90 to +30
Pentane	−200 to +30

thermometer and other instruments related to expansion of liquids. Table 5.2 also indicates the related temperature ranges of thermometers (Fig. 5.3).

Change of State

There are two different instruments that can be used for measuring the change of state class. These instruments are known as vapour pressure thermometers and Pyrometric Cones, Bars and Rings and are described in more detail in Table 5.3.

5.1.2.2 Electrical Methods

Temperature measurement based on the changes of electrical resistance of a metal with temperature has become popular because they are very convenient techniques and they furnish a signal that is easily detected. The measurement can be amplified and or used for control purposes. Furthermore, they are generally quite

Table 5.2 Expansion thermometer

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Bimetal strip (solid)	Two strips of metals are bonded together. When heated, one side will expand more than the other, and the result in bending is translated into a temperature reading by mechanical linkage to a pointer. Range: −35 to 550 °C, Accuarcy: ±0.5 to 1 % Narrow operating temperature range. Cannot be remotely indicated. Operation is dependent on external factors	A: Not expensive. Provide direct reading. Portable and do not require outside power D: Not as accurate as thermocouples and RTDs	Used to operate thermostats and steam traps. As a switch in a heater circuit. Used to drive a pointer on an instrument It provides compensation for changes in atmospheric temperature
Liquid in glass thermometer (liquid)	This is the most common type of thermometer. It consists of a glass tube known as the capillary tube that has a bulb at one end, a capillary hole down the axis connected to a reservoir in the bulb filled with a liquid. Usually filled with mercury and when heat is applied, the liquid level will rise and, respectively, if cooled will fall. Range: −80 to 450 °C, Accuracy: ±0.5 %	A: Has good expansion D: The glass can melt if exposed to high temperatures	Airplanes to measure the atmospheric icing conditions Measure temperatures of swimming pools and saunas. Used in medicinal examination
Liquid in metal thermometer (Liquid)	The bulb and capillary are completely filled with liquid. The change in volume on thermal expansion is indicated by a Bourbon tube. Range: −50 to 60 °C, Accuracy: ±1 %	A: Very accurate readings. Durable long stem are practical in industrial use D: The liquid in the thermometer may need to be shaken down prior to use	Air-conditioning units, burners, boilers, cookers, heat exchangers, water works tanks
Gas thermometer (Gas)	The effect of temperature on gas changes the pressure that is indicated by the movement of the Bourdon tube. If the change in volume of the tube as it curls can be neglected a fairly linear scale is produced. Range: −100 to 800 °C	A: Response is more rapid than liquid types. Can measure a wide range of temperatures D: Has to be calibrated before being used. Handling this thermometer is comparatively difficult	Process systems

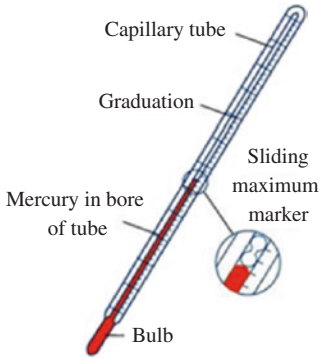


Fig. 5.3 Liquid in glass thermometer

Table 5.3 Change of state thermometers

Vapour pressure thermometer (Change of State)	The vapour pressure of a volatile liquid is directly proportional to its temperature. By using an expansion thermometer system to enclose an unstable liquid and a space into which it can vapourise a thermometer which uses this effect is produced. The vapour pressure depends only on the temperature of the interface which is normally within the bulb. It is therefore independent of the volume of the bulb, the capillary and the ambient conditions in which the capillary and Bourdon tube are situated. Range: 0–500 °C Accuracy: ±1.5 %	A: Is cheaper than mercury in steel. Does not suffer from ambient temperature effects. Can be made with smaller bulb. A large number of liquids can be used	Various processing systems
		D: Slow response rate	
Pyrometric cones, bars and rings (change of state)	The melting point of specific mixtures of certain minerals can be used to determine temperature. The melting point of particular minerals such as silicate and calcium carbonates is placed in a kiln in the form of cones, bars and rings and their reaction in the kiln determines the temperature. Range: 600–2,000 °C, Accuracy: ±10 %	A: Are useful if clarification is required that the temperature of a piece of equipment or material has not exceeded a certain level	Used in the ceramic industry to determine kiln temperature. Used with steam traps, when a trap exceeds a certain temperature, a white dot on the sensor label attached to the trap will turn black. Used also in the welding process
		D: Long response times. Accuracy could sometimes be high as results are based on interpretations	

accurate when calibrated properly. Table 5.4 provides details of a thermoelectric thermometer, also known as a thermocouple including other electrical methods for measuring temperatures. However, a number of factors govern the type of thermoelectric thermometer that can be used. In this section, these are briefly described.

Thermoelectric Thermometers (Thermocouples)

One of the most common industrial thermometers used today are the thermoelectric thermometers or thermocouples as they are generally referred too. The choice of

Table 5.4 Electrical methods of measuring temperature

Instrument	Description, range, accuracy	Advantages (A) and disadvantages (D)	Application
Resistance thermometer	Resistance thermometers also called resistance temperature detectors (RTD's) are temperature sensors that capitalise on the fact that the electrical resistance of a material changes as the temperature changes. They consist of spirals of fine wire wound on insulating formers. The variation of electrical resistance of the metal when heat is applied provides an accurate temperature measurement. Range: up to 1,000 °C, Accuracy: ±0.05 %	A: High Accuracy, can read up to 1,000 °C and above. Require no cold-junction measurement. Suitable for precision applications	Various process industries
		D: Rarely used above 650 °C, as it becomes difficult to prevent contamination by impurities of the metal sheath. Response speeds tend to be slower due to bulb size	
Thermistors	These are based on resistance change in a ceramic or polymer. This generates a higher precision with a limited temperature range. Can be classified into two types: positive temperature coefficient (PTC) and negative temperature coefficient (NTC). The resistance drops nonlinearly with temperature rise. Range: −100 to 300 °C , Accuracy: ±0.001 %	A: Have a high resistivity. Can respond quickly to temperature change	PTC thermistors are used as current-limiting devices for circuit protection
		D: May de-calibration at high temperatures. Use is limited to a few hundred degrees. Fragile and care must be taken	NTC thermistors can be used to provide high initial resistance to power supply circuits

(continued)

Table 5.4 (continued)

Instrument	Description, range, accuracy	Advantages (A) and disadvantages (D)	Application
Thermoelectric thermometers (thermocouple)	Consists of two dissimilar metals, joined together at one end. Changes in the temperature at the junction induce a change in electromagnetic force (emf) between the other ends. As temperature goes up, the output emf of the thermocouple rises, though not necessarily linearly. The current flowing in the loop can be detected by a galvanometer, which can also be used to convert heat into electric power. Range: 200–2,600 °C , Accuracy: ±0.1 %	A: Provides stable response. Capable of measuring a wide range of temperatures, up to 2,600 °C. The thermocouple junction may be grounded and brought into direct contact with the material being measured. Less expensive than other temperature sensors and can be easily installed	Thermocouples are widely used in science and the industry; applications include temperature measurement for kilns, gas turbine exhaust and diesel engines
		D: Two temperatures need to be measured. Are usually complex and generates room for error	
Measurement of electrical variable (Wheatstone Bridge)	In resistance thermometers, Wheatstone Bridge circuits are most commonly used. The current flows through the resistance in the bridge and also through the thermometer bulb. As a result, this generates heat in the thermometer producing an indication of change. To achieve the most accurate readings, it is important to reduce the current through the thermometer to as low as possible	A: High accuracy and sensitivity. Can be used to measure numerous quantities, such as capacitance, impedance	Various
		D: Requires calibration with the leads attached	

materials for a thermocouple is governed by a number of factors. They must develop a relatively large emf that does not fluctuate from the calibrated value when in use. The material must also be able to withstand the maximum temperature of use. Usually, the thermocouple is rugged and easy to install, and if properly connected, it provides an accurate temperature of its junction. However, care must be taken to ensure that the correct compensating cable is used and the correction for the temperature of the cold junction is made as typified in Fig. 5.4. As a result, there are four types of thermocouples commonly used for industrial application, which are described here.

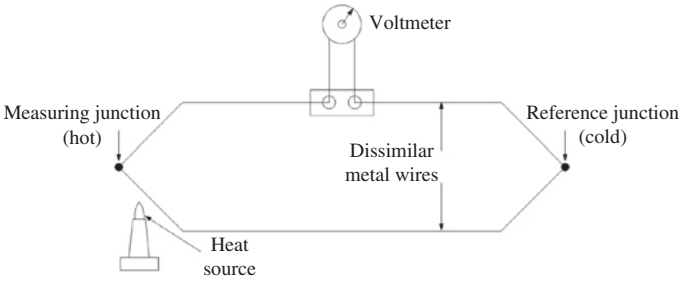


Fig. 5.4 The basic principle of a thermocouple

Base-metal Thermocouples

Base-metal thermocouples range from −200 to 1,200 °C. The different types that are available are listed below.
Note that

1. Copper/constantan elements are usually only used for laboratory measurement.
2. Iron/constantan elements are used up to 800 °C under reducing conditions. They deteriorate rapidly under oxidising conditions or in the presence of acids.
3. Chrome/aluminium alloys are suitable for up to 1,000–2,000 °C under oxidising conditions.

These thermocouples can be obtained in a mineral-packed type of construction which gives strong, small diameter-flexible thermocouples.

Rare-Metal Thermocouples

Rare-metal thermocouples are those based on platinum metals. They have relatively low thermoelectric power when the material is measured in response to a temperature difference across that material. This allows them to produce a good accurate measurement and is the preferred types for standard or reference thermocouples. It is advised that all rare-metal thermocouples should be used along with a good-quality sheath (protective cover) that is resistant to the temperature and conditions of use, as they are vulnerable to contamination by dirt and grease. The sheaths come in many different materials, such as mild steel, stainless steel, and chrome/iron and are all receptive to varying temperatures and conditions as well as oxidation and mechanical damage. The three types of thermocouples listed below are used in relation to rare-metal thermocouples and can be used in oxidising and inert atmospheres continuously at temperatures up to 1,600 °C.

	Type	Range (°C)	Accuracy (%)
1	S (Pt/Pt10 %Rh)	0–1,600	±1
2	R (Pt/Pt13 %Rh)	0–1,600	±1
3	B (Pt6 %Rh/Pt13 %Rh)	0–1,600	±1

Electrical Variable Measurement

In Table 5.4, the Wheatstone Bridge circuit is described. This method is one way in which to obtain a resistance and temperature from current flowing through a circuit. This method generates heat in the thermometer and causes its temperature indication to change.

5.1.3 Radiation and Optical Pyrometry

Radiation pyrmometry typically shown in Fig. 5.5 is the type used when the temperature is too high or at the point in which it is difficult to put radiant energy onto the junction of a thermocouple at which temperature is required.

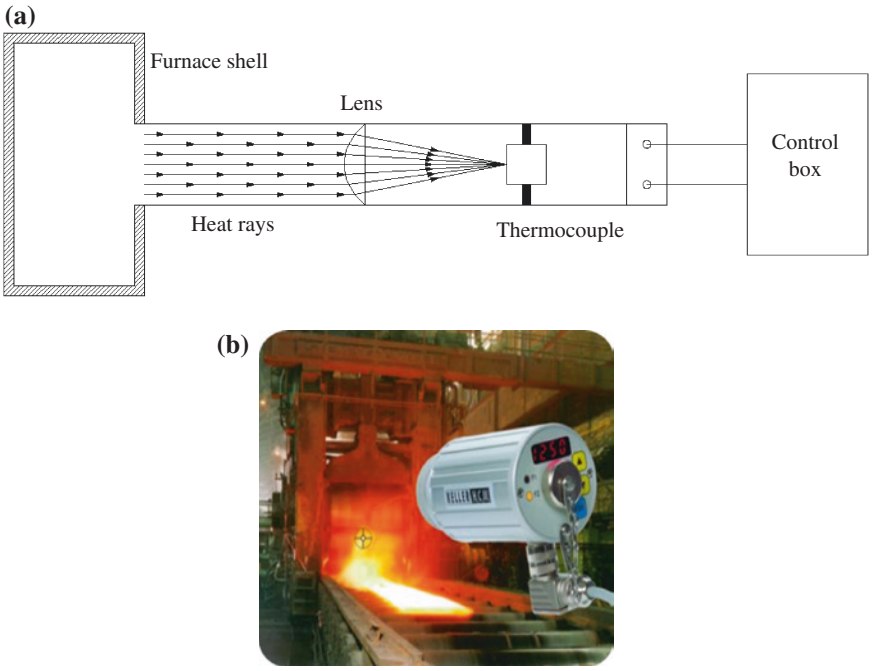


Fig. 5.5 Radiation pyrometer. **a** Design features. **b** Typical instrument in continuous catching operation [1]

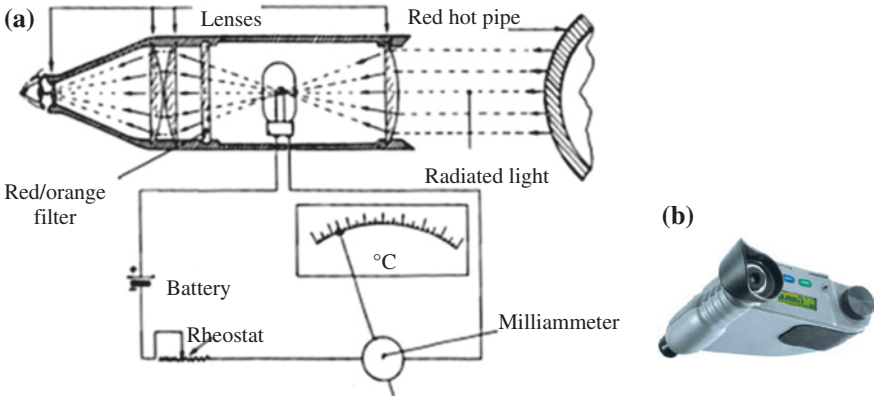


Fig. 5.6 Optical pyrometer. **a** Design features. **b** A typical illustration of the instrument [2]

Some solids, such as metals, when hot begin to emit light with a very dull red colour, and as the temperature increases, the colour becomes a brighter red, orange, yellow and so on. The colour emissions occur at a particular temperature and hence can be used as a means of measuring the temperature if suitable calibration is adopted. This process is known as *optical pyrometry* (see Fig. 5.6) and the instruments associated with this type of measurement technique are also highlighted in Tables 5.5 and 5.6.

The principle of operation of radiation and optical pyrometer is also shown in Figs. 5.5 and 5.6.

Table 5.5 Total, partial and chopped radiation pyrometers

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Total radiation pyrometers	This is a detecting element which converts the radiant energy to electrical energy. Non-contact temperature sensors use the concept of infrared radiant energy to measure the temperature of objects from a distance. After determining the wavelength of the energy being emitted by an object, where leads are taken from the sensor to the measuring instrument where the use of integrated equations takes into account the body's material and surface qualities to determine its temperature	A: Does not require direct contact with the object being measured. Can be used in circumstances where the objects are moving	Used in assembly lines
		D: Installation cost is high. Requires maintenance	Furnaces, process manufacturing industry

(continued)

Table 5.5 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Partial-radiation pyrometers	An optical system which ensures that radiation from the target either falls or is focused on the detecting element. It is most likely that the waveband is determined by the detector itself. Detectors in use are photo-electric, photoconductive and photoemissive types	A: There is the possibility to select a sensitive waveband to suit particular requirements	Food processing engineering
		D: Installation cost is high. Requires maintenance	
Chopped radiation pyrometers	An optical system which selects only radiation in the required waveband for detection. The radiation entering is periodically interrupted by a mechanical device, which is usually a rotating disc with alternate open and opaque segments. It is possible to use AC amplification and sensitive detectors. The waveband can then be chosen to suit particular applications. Accuracy: ±1 %	D: Installation cost is high. Requires maintenance	Various

Table 5.6 Ratio, reflecting and optical pyrometers

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Ratio (two colour) pyrometers	An amplifier and recorder or indicator which measures and displays the output of the detector. Basically two partial-radiation pyrometers built into one. The two halves are sensitive to different wavebands and the temperature is obtained from the ratio of the outputs. Range: 700–1,800 °C	A: Reading is independent of the target paragraph emissivity provided emissivity is the same in both wavebands. The reading is independent of target size, provided the target surroundings are at least 300 °C colder than the target. The reading is in depended of partial obscuration of the target and of atmospheric absorption	Used in cement kilns and for the measurement of the temperature of liquid metals
		D: More expensive than single-waveband partial-radiation pyrometers	

(continued)

Table 5.6 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Reflecting hemisphere pyrometers	The radiation detector is mounted at the top of the gold-plated reflecting cup, which is placed on the hot surface. The temperature is determined without knowing the emissivity of the surface, since it is measured essentially in black body conditions. A silicon cell would be the preferred detector option. Up to 1,300	Using a solar cell detector is faster and better able to withstand high temperatures It is less sensitive to emissivity errors and is more expensive	Molten metals, forging, ceramics, investment casting and calibration procedures
Optical pyrometers	This has some form of telescope, through which the high temperature object is observed. Usually, the telescope will have a light filter installed, to reduce damage to eyes. The telescope is focused on the object whose temperature is required. Colour filters can be introduced to the side of the object and the colour of the filter is matched with the colour of the object. There is a calibrated scale on the side of the object, which records the temperature. Range: Ranges: 700–1,400 °C, Accuracy: ±0.5	A: Accurate, robust D: Expertise	Various

5.1.4 Measurement of the Bulk Temperature of Solids and Liquids

This is described as one of the easier temperature-measuring problems encountered in industry. Provided that the sensor is completely surrounded by the medium and immersed to an adequate depth, no corrections are required for transient mass or heat transfer effects, conduction effects, radiation or velocity effects. The most common methods of measuring industrial bulk temperatures are by using thermocouples and resistance thermometers, which are referred to in Sect. 5.1.2.

5.1.5 Measurement of the Surface Temperature of Solids and Liquids

When heat transfer is of interest, it is essential to note that the surface is the only portion of the system that is accessible for measurement of temperature. In this case where the surface is accessible, measurement can be made with surface pyrometers such as the radiation and optical methods discussed in Sect. 5.1.3. Each of these instruments is easy to use and is suitable for discontinuous measurements.

Where the surface may not be accessible, for example in industrial environments, it is necessary to fix the detecting element in place prior to heating the surface (the elements used are thermocouples). This option is simple; however, precautions must be taken to avoid errors that can be introduced through changes in local transfer conditions, uncertainty about the positioning of the hot junction and heat loss by conduction along the thermocouples. In order to reduce this occurring, couples should have smaller diameters and be well insulated.

5.1.6 Gas Temperature Measurement

Measuring gas temperature is very difficult as a gas is rarely in thermal equilibrium with its surroundings and in furnaces, its spectral characteristics are unknown.

As a result of this, there are only three instruments, shown in Table 5.7 that are suitable for the measurement of gas temperature for industrial furnace applications:

1. The suction pyrometer, i.e. metal and water-cooled metal (See also Fig. 5.7).
2. The Venturi Pneumatic Pyrometer.
3. The Schmidt radiation method

5.1.7 Heat Flux Measurement

On some occasions, it is necessary to make a direct measurement of the heat flux incident upon a point in a furnace wall. The three instruments used to measure heat flux are: plug type, radial disc and ellipsoidal (see Figs. 5.8, 5.9 and 5.10, respectively). Each of these are available commercially, but are usually made specially to suit a particular application. See Table 5.8 for particulars on each of these instruments.

Table 5.7 Measurement of gas temperature

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
The metal suction pyrometer	This instrument measures the true temperature of a gas. The sensor used is a type S	A: Instrument can be widely used. Has a high response rate. Is portable and robust and only requires AC power supply. Protects the thermocouple from radiation errors by shielding	Manufacturing industry
	Thermocouple is shielded by 3 concentric ceramic tubes to minimise the radiative exchange from the surrounding surface or flames. The gas that is drawn is through the stainless steel tube and then passes out of the instrument through the T-piece to the suction line. The emf generated by the thermocouple is fed to a socket housed in the assembly at the rear of the T-piece. Compensating cable is then used to connect the pyrometer to the measuring instrument. Range: 0–1100 °C, Accuracy: ±2.5 %	D: Shield can be blocked by dirty gases, therefore causing deterioration of the thermocouple and a slow response speed in some cases	
Water-cooled metal suction pyrometer	It was developed to be used as a high-temperature accessory for use with portable suction pyrometer described above. Range: 0–1,600 °C, Accuracy: ±2.5 %	A: Can be extended to read up to 1,800 °C. Has a rapid speed of response	Heat exchanger
		D: Requires a sufficient flow of water to avoid instrument damage	
The Venturi pneumatic pyrometer	Measures both gas and flame temperatures. The gas temperature is measured by comparing the gas density at the unknown temperature with its density at a lower temperature. The density measurements are achieved by venturi restrictions, which state that if ΔP_h and ΔP_c are the pressure drops across the hot and cold venturies, respectively, and if gas temperatures at these points are T_h K and T_c K, then $T_h \alpha (\Delta Ph/\Delta Pc)\, T_c$. Range: 0–2,500 °C, Accuracy: ±2 %	A: Has a very rapid rate of response	Manufacturing industry
		D: They are big and bulky	
The Schmidt radiation pyrometer	This technique provides a measurement of the effective temperature and emissivity of the gas. Two readings are taken through the gas with a total radiation pyrometer, one with a cold background the other with a hot background at a known radiation temperature. Alternatively, a twin-beam pyrometer can be used, one beam sighted on a hole in the furnace and the other on a hot region of the lining which contains a thermocouple. The temperature obtained is a mean value along the optical path through the gas and this can be unacceptable if large temperature gradients are involved. The accuracy also depends on the emissivity of the gas	D: They give nonlinearly weighted averages of unknown quantities across the combustor	Combustion chamber measurements
		Taking gas samples is difficult without disturbing the combustion process	

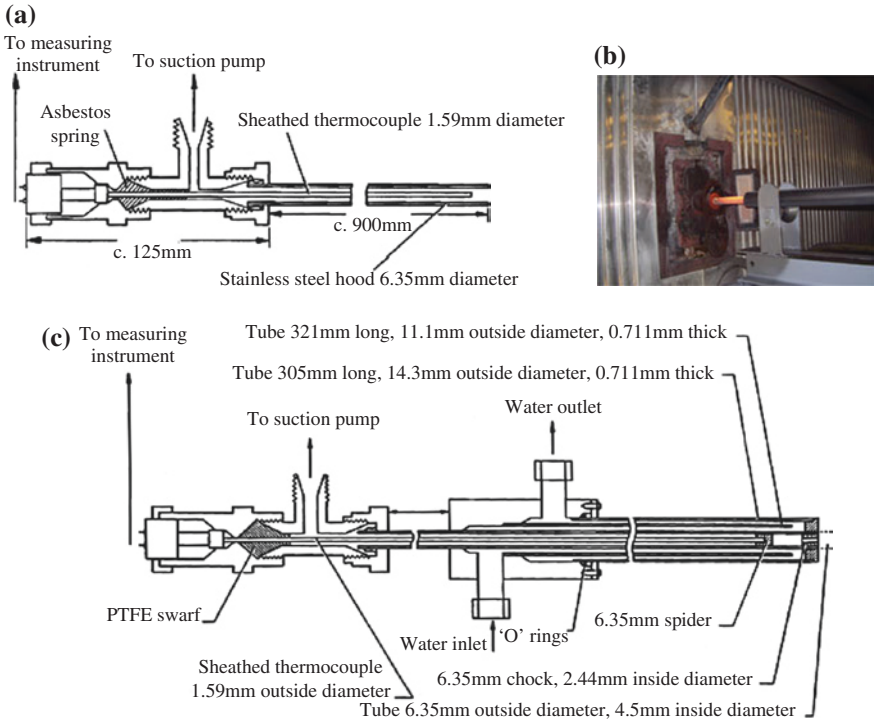


Fig. 5.7 Typical metal suction (**a**) [2, 3] with illustration (**b**) and water-cooled (**c**) pyrometers

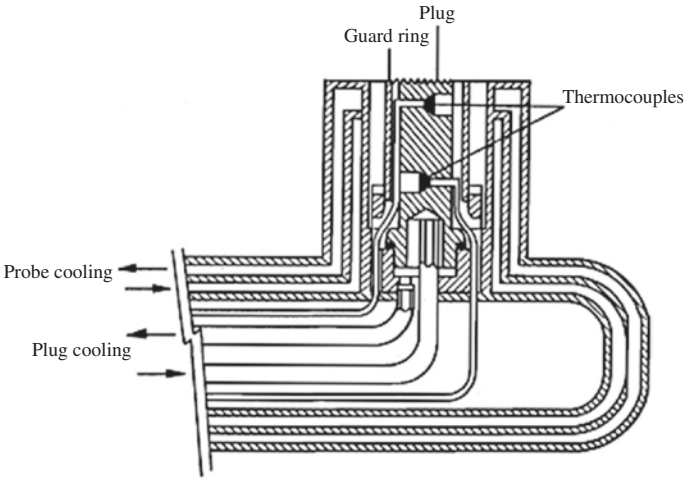


Fig. 5.8 Plug-type heat flux [4]

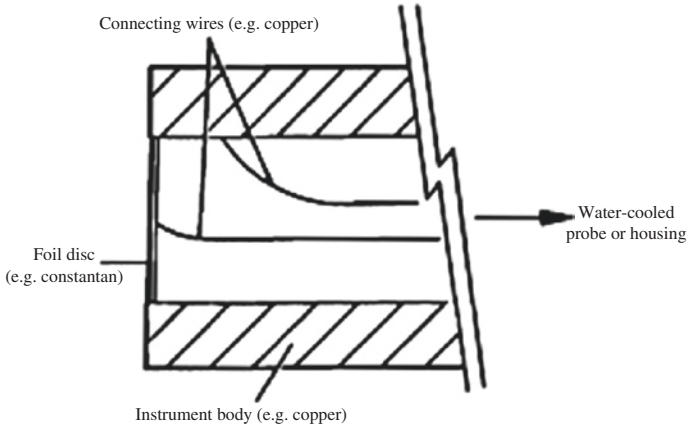


Fig. 5.9 Disc-type heat flux [4]

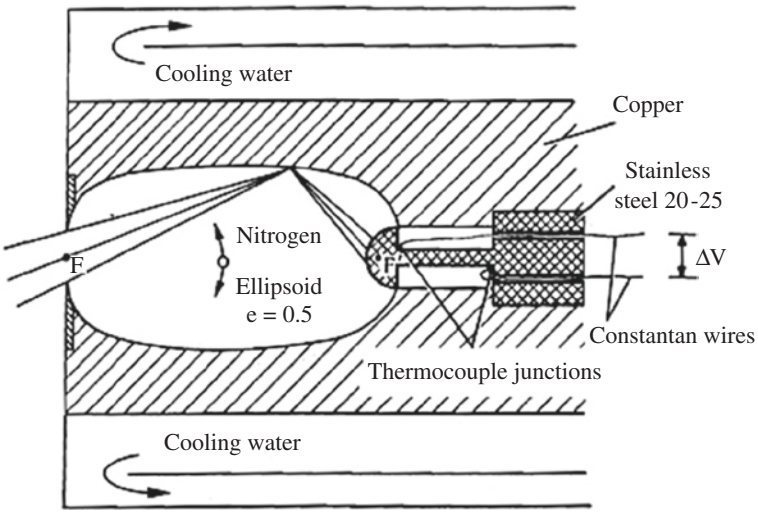


Fig. 5.10 Ellipsodial heat flux [5]

5.2 Pressure

5.2.1 Introduction

Pressure can simply be defined as a measure of force per unit area, and the units of pressure are measured in either Pascal (1 $N/m^{2)}$, bar (10^5 Pa) or pound per square inch, gauge/absolute and mmHg. This section will go on to discuss various pressure terms and measurements commonly associated with the gas industry.

Table 5.8 Plug, Radial disc and ellipsoidal heat flux measurement of heat flux

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Plug-type heat flux meter	The heat incident upon the plug face is conducted through the plug to the water cooling at the rear. A guard ring system ensures the laminar flow of heat through the plug and the whole assembly is mounted in a water-cooled probe. The measurement of the heat flux is made by two thermocouples mounted with their hot junctions a distance apart on the axis of the plug, the heat flow through the plug, and hence, the heat flux incident on the plug face can be determined using the formula: $Q = A\lambda_m (T_2 - T_1)/l$	A: Accurate D: Not simple in construction	Various
Radial disc heat flux meter	Also measures the total heat flux. Here, the heat flux incident on the disc is conducted away radially to the instrument body which is water-cooled. The output is obtained from the differential thermocouple formed by the junction of the disc circumference with the instrument body and the wire attached to the disc	A: Simple in construction and can be used in positions inaccessible to the plug type D: Less accurate and less robust. Each instrument requires individual calibration	Various
Ellipsoidal radiometer	Has been designed to measure the total radiation from a solid angle. Radiation that passes through the instrument apertures is focused by the ellipsoidal mirror onto the blackened hemispherical detector, which is effectively the front surface of a very simple heat flow plug. The output is obtained from thermocouple junctions on the plug axis. A nitrogen purge is fed to the ellipsoid through small holes around its mirror diameter to prevent entry of furnace gases and particles	D: Slow response time. Not sensitive to convectional heat transfer effects	Used in fire testing. Furnace and flame radiation measurement since the sensing element is not exposed to hot gases

Absolute pressure refers to the absolute value of the force per unit area exerted on the surface of the fluid. Absolute pressure is defined as the difference between pressure at given point in a fluid and absolute zero or perfect vacuum.

Gauge pressure is the pressure difference between the absolute pressure and the local atmospheric pressure. It can also be referred to as differential-pressure.

The local atmospheric pressure varies from one place to another depending on the temperature, altitude and local weather conditions.

Differential-pressure refers to the measurement of the difference between any two pressures. Pressure may be measured directly in two ways:

1. Pressure due to a fluid may be balanced by the pressure produced by a column of liquid of known density.
2. Pressure due to a fluid may be allowed to act over a known area, producing a force whose magnitude depends on the pressure.

5.2.2 Liquid-Column Pressure Gauges (Manometer)

There are various types of manometers used, which have different ranges and accuracies depending of the manometric fluid and configuration. The basic type is the U-tube manometer, which is a measuring device that measures pressure near atmospheric pressure. They are usually simple in construction containing monometric fluid, which is usually mercury due to its high density and low vapour pressure, as shown schematically in Fig. 5.11.

Liquid-column manometers have highly linear calibration with poor dynamic response. When measuring a vacuum pressure, the manometric fluid may evaporate and contaminates the vacuum if its vapour pressure is too high.

5.2.3 Force-Balanced Pressure Gauges

The force produced by a pressure can be measured by balancing it against a known weight or by the strain or deformation it produces in an elastic medium.

Measurement by balance against a known weight: In this kind of instrument, the force produced on a piston of known area is measured directly by the weight it will support. The pressure gauges operating on this principle are: (a) the piston type, (b) the ring-balance type and (c) the bell type as shown in Fig. 5.12. Table 5.9 (A) and (B) provides more details of these gauges.

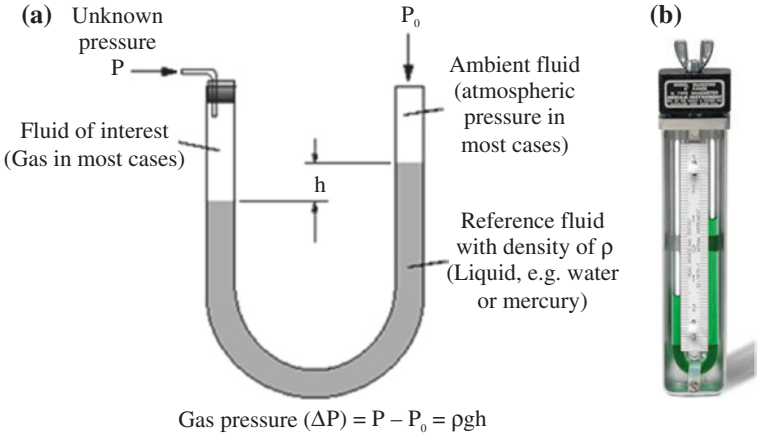


Fig. 5.11 Schematic diagram of U-tube manometer (**a**) and illustration of the device (**b**)

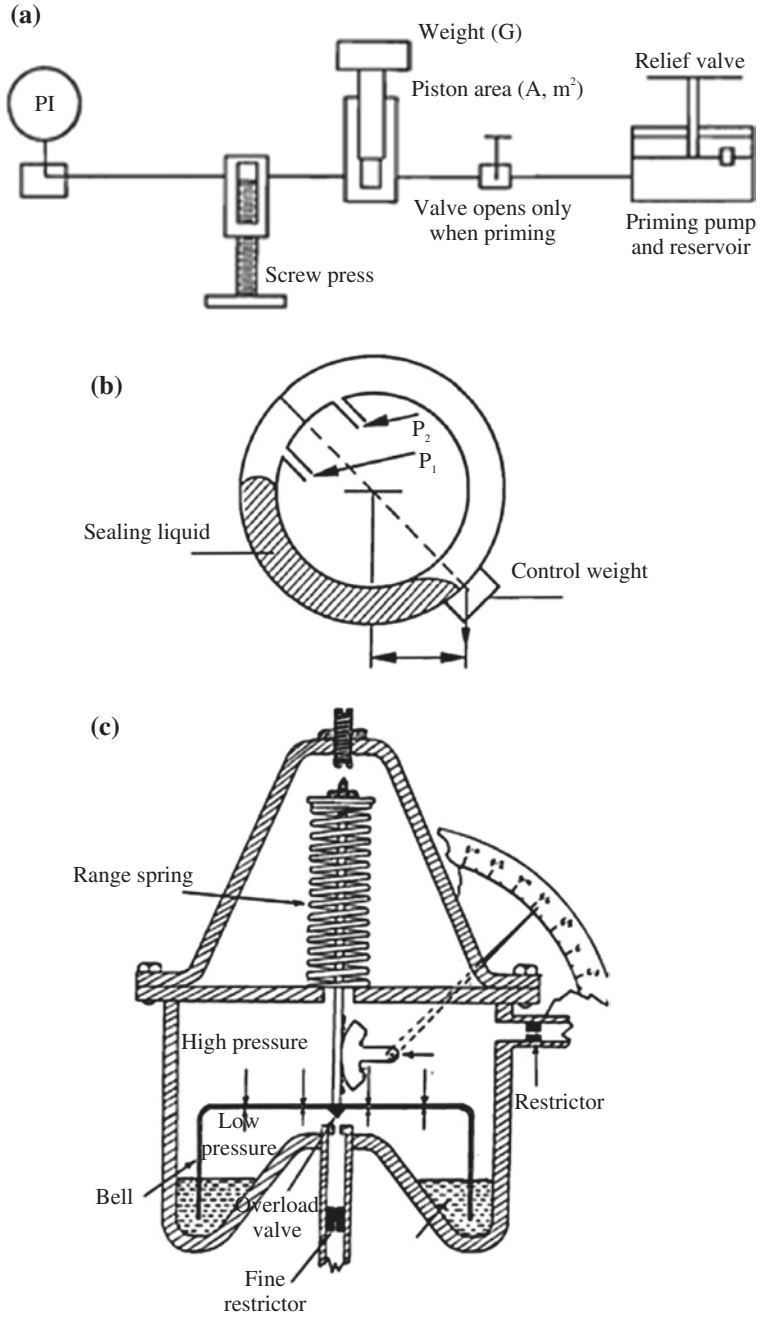


Fig. 5.12 Different types of pressure gauges. **a** The piston-type. **b** The ring-balanced. **c** Typical bell-type differential meter for a very low range (thin wall)

Table 5.9 Liquid-column pressure gauges

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
(A)			
U-tube manometer	U-tube manometer is a simple manometer which consists of a glass U-tube containing liquid. The difference in heights of the liquid in the two arms can be used as an indication of the pressure if the fluid is specified	A: It is used for measurement of differential-pressure	Simple in construction. Low cost hence easy to buy. Very accurate and sensitive. It can be used to measure other process variables
		D: Fragile in construction. Very sensitive to temperature changes. Error can happen while measuring the height	
Well-type manometer	The well has a cross section larger than that of the indicating tube rises appreciably. A direct reading of height can be taken from the scale which is designed to correct for the error introduced by the fall of level in the well. The nature of the fluid whose pressure is being measured and the pressure required would determine the kind of fluid used in the gauge	A: Used for standard calibration of other pressure-measuring elements. Used for more direct reading instrument than the U-tube manometer	They have no mechanically moving parts. They are not expensive. They are simple
	Up to 800 mmHg. In most cases, the uncertainty of a manometer reading is ± 0.5 of the smallest scale graduation	D: The higher pressure source being measured must always be connected to the well connection	
Inclined-tube manometer	The inclined tube may be straight or curved, in each case the gauge can be calibrated to give a flow measurement directly. Inclined-tube manometer is used to measure small differential-pressure. Up to 40 mm H_2O. Can read up to 2 mm. Accuracy: ± 3 %	Used to measure small differential-pressure	It is used for measuring pressure above or below atmospheric pressure. It is not expensive. It is simple and compact
(B)			
Differential liquid manometer	Differential liquid manometer utilises the movement of an interface between two liquids which have different relative densities (specific gravities), different colours and do not mix. The cross-sectional areas of the wells are usually ten times larger than that of the U-tube	Used to measure the flow dynamics of gas flow in pipe	Various

(continued)

Table 5.9 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Micro-manometer	Micro-manometer consists of a shallow U-tube system, the left-hand limb being in the form of an inclined sight tube. The right-hand limb consists of a reservoir which may be raised or lowered. The height of the reservoir in relation to the sight tube is adjusted by means of a vertical micrometre screw. The gauge is constructed so that the opposing effect of temperature change on the surface tension and density of the gauge liquid largely cancel out. 0–3 bar and 0–50 mbar	Can measure small differential-pressures accurately	Various

Measurement of strain or deformation: These pressure elements are mechanical devices that are deformed by the applied pressure. They possess elasticity, and when deformed, the stresses establish equilibrium with the applied pressure. The choice and design of the type of element used depends on the magnitude of the pressure to be measured. These are, namely (a) the bourdon-tube, the diaphragm type and (c) the bellows type. Figure 5.14 shows schematically typical force-balanced pressure gauges types. Table 5.10 (A) and (B) also provides further details of these pressure gauges (Fig. 5.13).

5.2.4 Pressure Transducers

Pressure Transducers are devices that use an electrical system in conjunction with mechanical gauges to convert the deflection, due to pressure changes into electrical signals. They are useful for measuring pressures continuously, such that the electrical signal supplied to a control system can be used to monitor the pressure variations. Figure 5.14 shows the schematic diagram of strain gauge pressure cell.

There are two major types of pressure transducers: mechanical and electrical as described briefly in the following with further details in Tables 5.10 and 5.11, respectively.

5.2.5 Mechanical Pressure Transducer

The pressure-sensing element in a mechanical transducer shown in Fig. 5.15 is composed of a metal diaphragm capsule which translates the applied pressure into mechanical displacement. The tiny movement is detected by a sensitive flapper

Table 5.10 Mechanical pressure transducers

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
(A)			
Piston pressure Gauge	The force produced on a piston of known area is measured directly by the weight it will support. If the pressure acting on the piston is p Pa and the area over which it acts is A m^2, the force produced will be PxA(N). If there is no friction then PA = F, the weight supported by the piston. The system shown is deadweight pressure tester which is used as a standard of pressure measurement for testing other types of instrument. Up to 55 MPa	They are widely used for maintaining calculable pressures in a range extending from about 3 kPa (gas media, absolute or gauge mode) to 1 GPa (hydraulic, gauge mode). This instrument can be used to calibrate virtually any pressure device	Easy to handle. It is usually compact and space-saving. Piston/cylinder assemblies are simple in configuration, and there are no mounting flanges or other geometric irregularities that can result in complicated stresses in the cylinder. It cannot be used to measure the pressure of a flowing fluid
Ring pressure gauge	The operating force depends on the difference between the pressures on the two sides of the partition. The control weight attached to the instrument is used for the calibration. The range is fixed by the nature and thickness of the material of construction and the pressure which the flexible connections will stand. The range of differential is fixed by the size of the size of the ring and the nature and quantity of sealing liquid	Used to measure the low pressure differential of the order of 100 Pa (1 mbar)	Gas industry and various

(continued)

Table 5.10 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Bell pressure Gauge	In the bell type of gauge, the force produced by the difference of pressure on the inside and outside of the bell is balanced against a weight or force produced by the compression of a spring. There are two types the thick-wall and thin-wall. In thick-wall instrument, the resultant force causes the bell to rise until equilibrium is reached. The thickness, density, cross-sectional area and the density of the sealing liquid are determined by range for which the instrument is to be used. In the thin-wall instrument, the controlling force is obtained by a spring. The travel of the bell is proportional to the differential-pressure to be measured. This type of instrument needs no overload device. The range will be determined by the modulus of elasticity of the spring and by the density of the sealing liquid. Range: 0–250 Pa and 0–3 kPa	Used to measure low static pressures	Gas industry and various
(B)			
Bourdon-tube Pressure Gauge	It consists of a narrow-bore tube of elliptical cross section sealed at one end. When the pressure to be measured is applied, the effect of the forces in tube is to straighten it so that the closed end is displaced. This displacement is magnified and indicated on the circular dial by means of mechanical linkages. The pressure changes must not produce forces which exceed the elastic limits of the tube so that it always returns to its original 'C' shape when the pressure is released. The tubes are made from a variety of materials in a variety of thickness. The material of construction depends on the nature of the fluid whose pressure is being measured and the thickness on the range of measurement required. Normal working pressure should not be greater than 60 % of the maximum pressure indicated. Range: 0.1–130 MPa Accuracy: ± 1 %	Industrial-type gauge design for fluid medium which does not clog connection or corrode 316 stainless steel type	They have vibration and shock resistant. It has resistant to corrosion because it made of stainless steel The filled fluid depends on the process and pressure medium to measure especially for food application

(continued)

Table 5.10 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Diaphragm Pressure Gauge	The movement of a diaphragm is a convenient way of sensing a pressure differential; this is used in diaphragm pressure gauges. The unknown pressure is applied to one side of the diaphragm whose edge is rigidly fixed and the displacement of the centre of the diaphragm is transmitted via a suitable joint and magnification linkage to the pointer of the instrument. In principle, the corrugated discs give a larger displacement and can be combined to form a capsule. The instruments are usually used above or below atmospheric pressure, and the material of construction is chosen to suit a particular application. Range: 2.5 mbar–0.4 MPa; Accuracy: 1–2.5 %	Diaphragm pressure gauges are used in the chemical industry, chemical fibre, synthetic fibre, oil, dyes, alkali and other industries. It is used when medium could be harmful to the pressure gauge	It measures fluctuating pressure better than bourdon gauge. Diaphragm seal prevents medium entering the gauge It cannot be used in a corrosive environment
Bellows Pressure Gauge	An alternative to the diaphragm stack is the bellows element which can be put into instruments to measure differential, absolute and gauge pressure. The material of construction includes beryllium, copper, bronze and stainless steel. The 'flexibility' of a bellows is proportional to the number of convolutions and inversely proportional to the wall thickness and modulus of elasticity of the bellows material. The 'compression modulus' of a bellows varies directly as the modulus of elasticity of the bellows material and the cube of the wall thickness. It is also inversely proportional to the number of convolutions and to the square of the outside diameter of the bellows. The spring rate can be increased by fitting an internal spring to oppose the force compressing the bellows. Measuring pressures up to 5 Mpa, though design for low pressure, Accuracy: ±1.00 %	Bellow elements are used in various applications where the pressure-sensitive device must be powerful enough to operate not only the indicating pointer but also some type of recording device	which converts the differential-pressure into a rotary movement by way of a torsion tube assembly and mechanical linkages

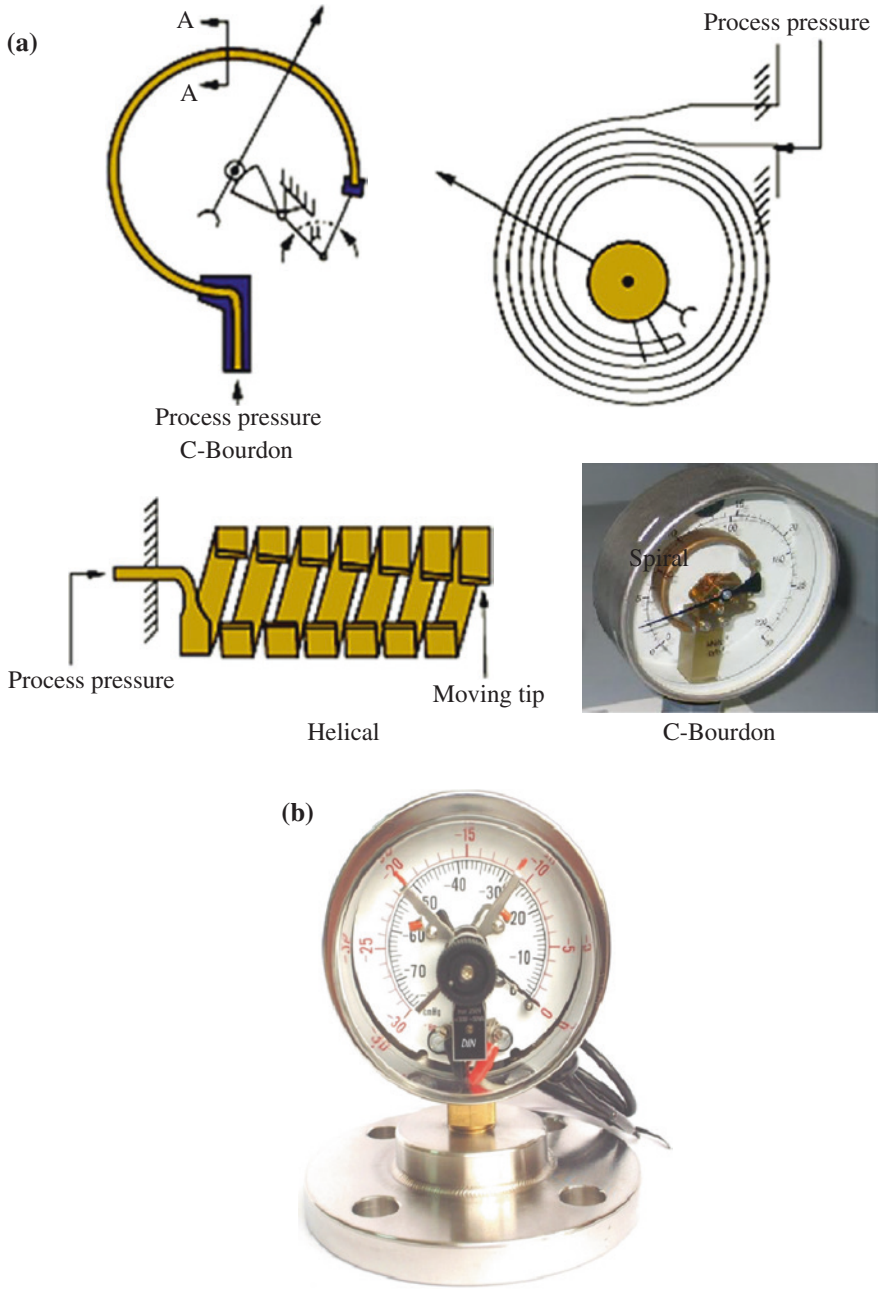


Fig. 5.13 Bourdon-tube (**a**) and diaphragm (**b**) strain type gauges

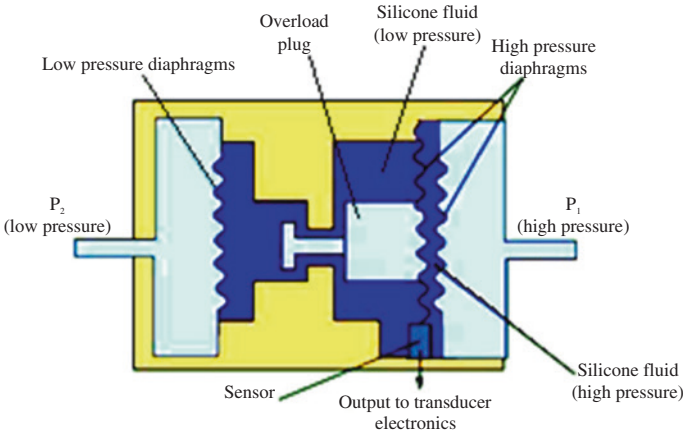


Fig. 5.14 Strain gauge pressure cell

Table 5.11 Electrical pressure transducers

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Strain gauge	In strain gauge pressure transducer, the round platform at the top could be a diaphragm which is pushed down by a rising pressure. In a weighing device, it would be part where the unknown weight is applied. When a downward force is applied to the platform (diaphragm), it pushes down on the cross-shaped spring. There are rods of insulating material extending above and below the cross and thin wire winding is arranged around these rods. When the centre of the cross is pushed down, all the rods will tilt inwards at the top and outwards at the bottom. Therefore, the resistance of the windings under the cross will increase, while the upper resistance will decrease due to the slight relaxation of tension in these wires. The Wheatstone Bridge uses full difference between these windings to give a good deflection of the meter for only small changes in force, 0.1– 0.25 % of full scale	They have fast speed of response Simple to maintain	Gas industry and various
		Small and easy to install	
		It can only measure narrow-span pressure	

(continued)

Table 5.11 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Load cell	Load cells operate on the same principle and are used mainly for weight measurement. This is basically a spring metal cylinder with a flanged base-plate. Four resistances are securely bonded on to the side of the cylinder. Two are arranged horizontally and vertically, they are constructed from long-track carbon-film resistors printed on flexible insulating sheets or they may be thin metal foil tracks. When a force—such as a weight—is applied to the top of the load cell, the walls of the cylinder becomes barrel-shaped under the load, only a very tiny movement occurs but it is sufficient to slightly stretch the horizontal resistors and slightly contract the vertical ones. Accuracy: 0.03–0.25 % full scale	A: The hydraulic type takes high impacts, insensitive to temperature. Pneumatic load cells intrinsically safe. Contains no fluids D: The hydraulic type is expensive and complex, while the pneumatic type has slow response and requires clean, dry air	Pneumatic load cells are sometimes used where intrinsic safety and hygiene are desired, and hydraulic load cells are considered in remote locations, as they do not require a power supply
Capacitor microphone	Capacitor microphones are suitable for measuring fluctuating pressures. Voltage is supplied from high-impedance source so that the charge on the capacitor remains constant .The fluctuating capacitance shows as a fluctuating voltage which can be amplified directly. The capacitor microphone is used due to the fact that small transducers essential to ensure adequate frequency response and also enable sitting close to the pressure-sensing element so avoiding long leads. Pressure transducers have to be calibrated since they are affected ambient temperature changes. Ranges: micron range to (70 MPa). Accuracy: within 0.1 % of reading or 0.01 % of full scale	A: Accuracy in reading D: Calibration required	Used to measure larger pressure differences than liquid manometers and other gauge types. In industry, it is used as a remote control transducer

and nozzle device that regulates the air output from the transmitting section of the cell. The output pressure operates a remote recorder (in a control room), and it provides a feedback force so that the output is directly proportional to the differential-pressure across the diaphragm.

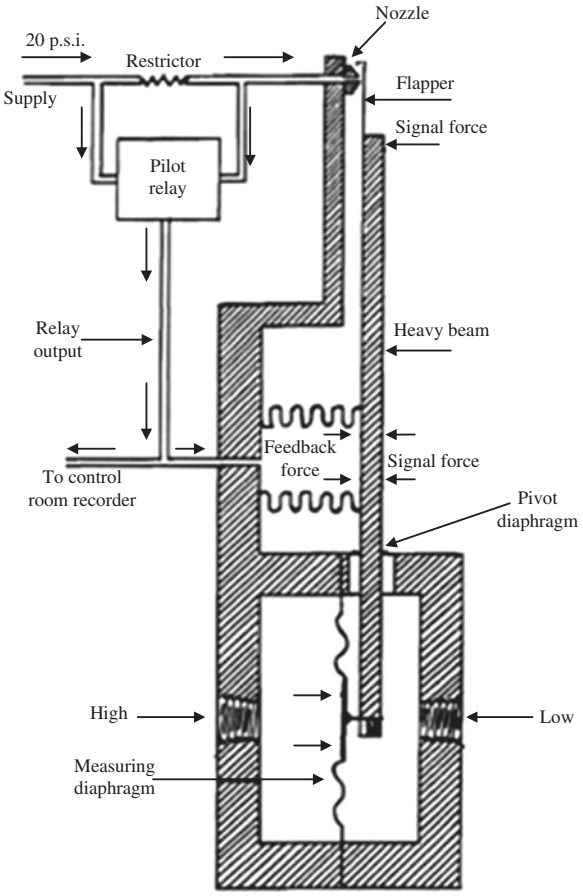


Fig. 5.15 Schematic diagram of mechanical transducer

They are available for temperatures as high as 300 °C without external cooling or up to 1,000 °C when fitted with special adaptors. The main advantages of these gauges in comparison with liquid manometers and other gauges are as follows:

1. Capable of transmitting results over a reasonable distance.
2. They can withstand high pressures and only react to the difference between the two applied pressures.
3. The air output can be fed directly into a control system, thereby automatically updating the controller.

See also Table 5.10 for further details.

5.2.6 *Electrical Pressure Transducer*

Electrical transducers convert responses from a pressure-sensing element into an electrical output and are much better than other gauges and mechanical transducers for remote reading, automatic recording or other signal processing requirements. The delays in response due to long pressure leads are eliminated. There are three types of electrical pressure transducers which are as follows: *strain gauge* put from El Kamki project*, load cell and capacitor microphone*, as shown schematically in Figs. 5.16, 5.17 and 5.18. Table 5.11 also provides for further details.

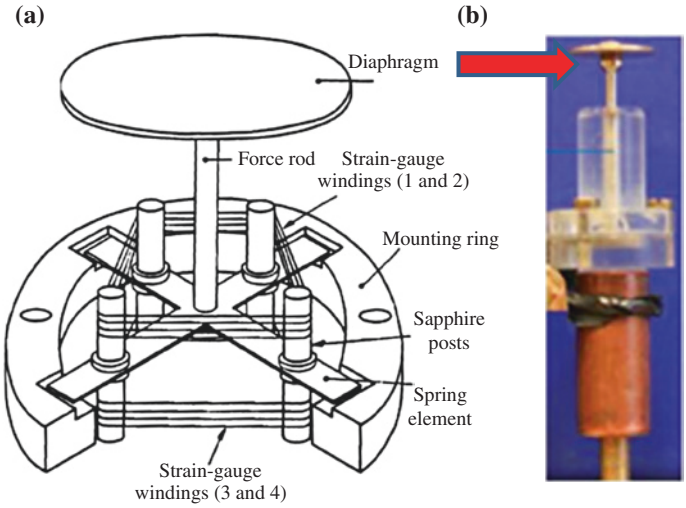


Fig. 5.16 Schematic strain gauge tranducer (**a**) and illustration (**b**)



Fig. 5.17 Typical load cell transducer

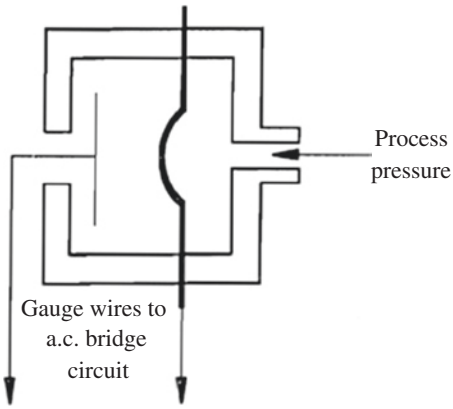


Fig. 5.18 Schematic diagram of capacitor microphone transducer

5.3 Gas Flow Measurement

5.3.1 Introduction

Measuring gas flow is a necessity. Gas suppliers need to know how much gas they are selling and thus how much they earn on the other hand, the gas consumer needs to know how much gas they consume and if the charges reflect that quantity prior to determining what meters to adopt. Part of meeting demand and supply gas measurement also answers other issues such as leakages. As the gas supply chain involves different stages such as transmission and distribution, the same is true with gas metering as domestic and non-domestic meters are required. The aim of this section is to present gas flow measurement devices from all stages; from laboratory instruments, domestic type of meters to those in high-pressure transmission lines used by producers and suppliers. At each stage, different meter specifications are required, and therefore, knowledge and experience are required in order to select the best meter suited for the task. The meters are presented in families according to their operating principle. They can all fall into two categories: those that extract energy from flow and those that add energy during measuring.

5.3.2 Pressure Differential Devices

When pressure differential devices are inserted into a pipe which carries a fluid, the device can cause a flow obstruction creating a pressure difference on either side of the device. Thus, the meters belonging to this family are known also as obstruction-type meters. These meters consist of a primary element that create a pressure change. The devices then sense the differences in pressure and transmit them by analogue or digital means in which the flow rate is estimated using Bernoulli's equation that is proportional to the square root of the pressure

difference across the obstruction. Obstruction of flow also means pressure loss, which is characterised by the discharge coefficient (C_d) a specification found in all differential-pressure meters. This family of meters is well documented in standards and codes of practice such as the BS 1042 [6].

5.3.2.1 Orifice Plates

The orifice plate shown schematically in Fig. 5.18, is a metal disc with a concentric hole that is placed into the pipe carrying the Flid being metered. They are simple and cheap and are available in a wide range of sizes. Thus, they are popular among gas industries worldwide since all natural gas that is bought and the largest sales are made using this type of meter. Table 5.12 provides further details with respect to this instrument (Fig. 5.19).

Figure 5.20 also shows the illustration of the orifice plate on a gas transmission line and the typical plate used in the laboratory.

5.3.3 Venturi Meter

The venturi meter is another obstruction device that is designed to minimise the pressure drop. It has a precision-engineered machining as seen from its shape as shown in Fig. 5.21 that accounts for the cost and accuracy. In addition, it has a smooth internal surface; thus, no sticking of particles from liquids or bubbles from gas could occur. The key point is the machining of the radius between the upstream cone and the throat where separation effects may be caused on the throat tapping being in the highest velocity region. A vital feature is the downstream diffuser which decelerates the flow and gives maximum pressure recovery. Thus, lower pressure loss but larger size. The size of the downstream diffuser can be reduced by up to 35 % without affecting the characteristics of the device, thus reducing costs. Figure 5.20 is a diagram that illustrates the principles of a venturi meter, and Fig. 5.22 is the examples of venturi at a construction site and one in a gas flow laboratory. Table 5.12 also provides more details for this flow measuring device.

5.3.4 Nozzle Meters

The nozzle meter is the result of the effort to reduce the size and cost of the classical venturi meter. Therefore, it is simple in construction but has higher pressure loss on flowing fluid. Figure 5.23 shows an schematic of a standard design of flow nozzle. Table 5.13 also gives more details specification of the instrument.

Table 5.12 Orifice plate and venturi meter

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Orifice plate	An orifice plate is a thin plate which has a hole in the middle with a bevelled downstream edge so that fluid or gas reaching it, is forced to converge as it goes through. Vena contracta is the point of maximum convergence and is located just downstream of the orifice. Thus, velocity and pressure change. After the vena contracta point, the flow profile retakes its original form. By measuring the difference in fluid pressure before the orifice plate and at the vena contracta (minimum pressure), the volumetric and mass flow rates can be calculated from Bernoulli's equation. The plate must withstand buckling; thus, it requires a sufficient thickness and is usually made of stainless steel. . For gas flow measurement, the orifice is placed concentrically within the pipeline. Key features:	A: Simple and robust. Well documented by standards [6] .Standard designs do not need calibration. Plate low purchase-priced measures liquids and gases	High-pressure gas metering on the large flows between suppliers and transmission companies. Fiscal metering
	Square upstream edge which sets the separation point thus contraction	D: Not available in many ranges. Performance changes with damaged plate mainly due to erosion which can produce changes in roughness and give an error up to 5 %. Vulnerable to poor flow profiles and swirl produced by governors, compressors, bends (up to 15 % error at 40^o angle). Not linear to flow rate output relation. Difficult to account accuracies which depends on complete installation and condition of flows. For high accuracy generates high secondary costs together with full installation. Collection of dust or debris before the orifice plate. A filter and flow straightener and regular plate's sharpness checking may reduce occurring problems	
	The conical entrance type has chamfered upstream edge and square downstream edge and is used on viscous fluids. It is designed to keep C_d constant to lower Reynolds numbers		
	Discharge coefficient $C_d = 0.6$ constant at numbers greater than 5×10^{-5} Re		
	Range: 3 to 4:1 difficult to account, depends on complete installation and condition of flows. Accuracy: 0.6 to ± 0.75 %		
Venturi meters	Venturi is a tubular set-up where fluid flows through a varying diameter pipe. A venturi tube typically has an entry cone of 30 degrees and an exit cone of 5 degrees. This geometry allows the venturi to maintain a practically constant coefficient of discharge when the Reynolds number (R_e) is greater that 3×10^5 at the throat of the tube. The geometry excludes curvature effects at the tappings having almost one-dimensional flow in the throat because of the acceleration. Discharge coefficient $C_d \approx 1$, constant at numbers greater than 3×10^{-5} Re. Measurement uncertainty of only ± 1 % 10–15 % of the pressure difference across	A: Compared to orifice: Higher efficiency and pressure recovery. Good performance at high β-ratios. More robust and less affected by internal erosion. Less sensitive to upstream disturbances	Used in water distribution system. Laboratory measurements. Not common for use in gas metering
		D: Occupies much longer length in line. More expensive to manufacture and install. Larger size difficult to handle. Not preferable in gas Industry	

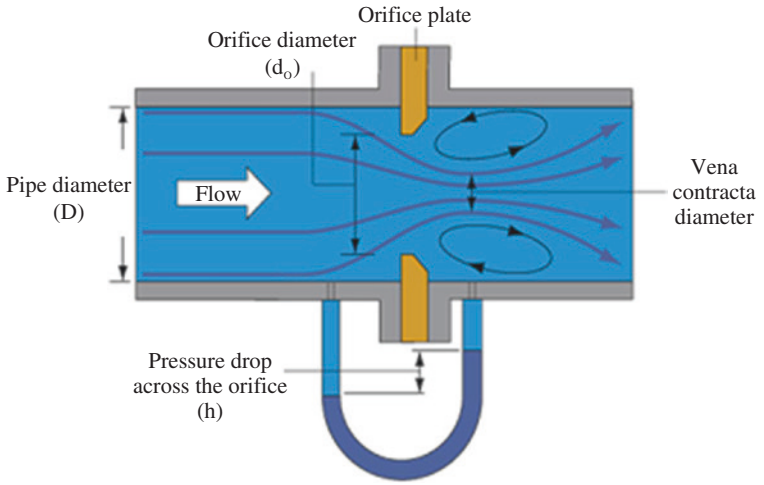


Fig. 5.19 Schematic diagram of orifice plate with D and ½ D tappings

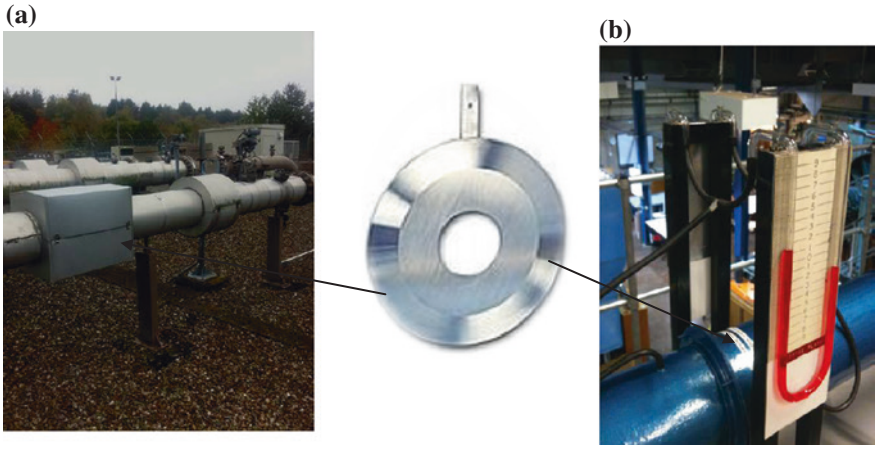


Fig. 5.20 Typical orifice plate installation on a gas transmission site (**a**) and orifice plate in a gas flow laboratory (**b**)

5.3.5 Pitot Tube

The Pitot tube is used for making temporary measurements of flow and has the ability to measure the local velocity of flow at a particular point instead of the average velocity measured by the other flow meters. This is very useful when examining a flow profile across a pipe cross section especially in the case of non-uniform flow, see also Table 5.13 and Fig. 5.24 for more details.

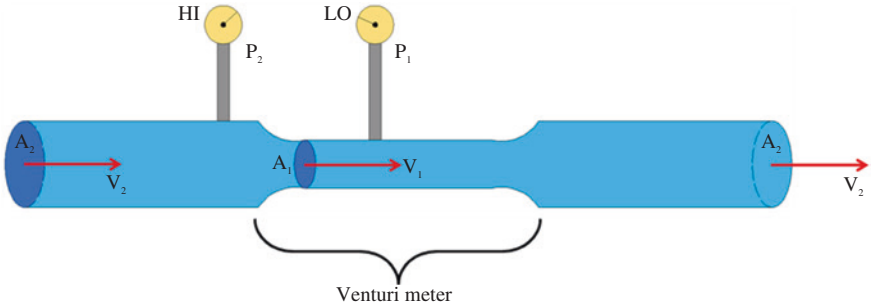


Fig. 5.21 Principles of a venturi meter



Fig. 5.22 Venturi meter on a construction site [7] (**a**) and in gas flow laboratory (**b**)

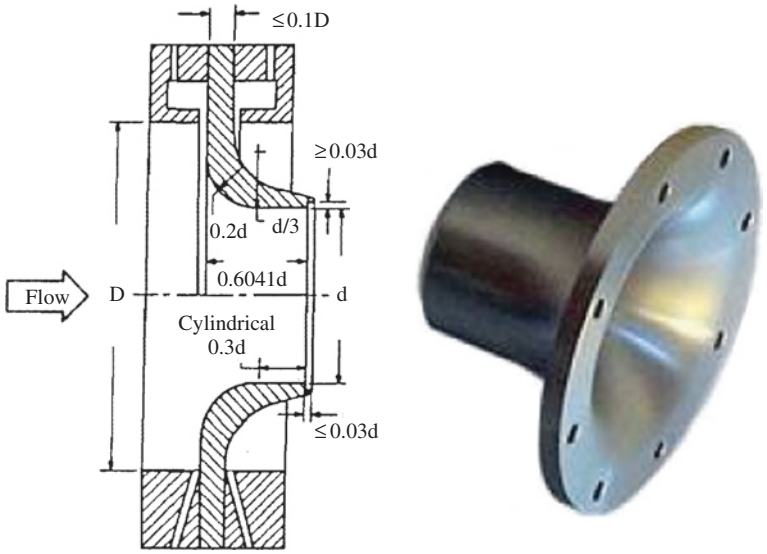


Fig. 5.23 Standard design of flow nozzle

Table 5.13 Nozzle meters, pitot tube, elbow and variable area meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Nozzle meters	The nozzle meter is a compromise between the orifice and the venturi meter. The nozzle resembles the compact orifice plate since the venturi's upstream cone is replaced by a nozzle and the diffuser is being discarded but yet keeps the favourable pressure loss characteristics of the venturi. The Dall tube is a nozzle variant and takes advantage of the streamline curvature at the throat tap producing higher differential-pressure for the same throat area. The Dall tube is used on water distribution systems. The critical (sonic) nozzles and sonic venturi nozzles have an accurately machined venturi tube so that at a certain flow rate the pressure at the throat is about half the upstream pressure. Pressure at the throat becomes sonic and a shock wave is formed as the flow expands after the throat, limiting the flow to a maximum value of Q_m (calculation formula to be inserted). Discharge coefficient $C_d = 0.9–0.99$. Comparable to venturies	A: Small and simple to use. Wear and erosion resistant. No sharp edges, higher C_d D: For very large flow rates, several devices must be used in parallel since the flow is limited by throat	Used as differential-pressure meters on gas flow systems. Critical (sonic) nozzles and sonic venturi nozzles in the calibration and testing of other types of gas meter. Can permanently check the accuracy of the orifice plate meter at site. Can prove turbine meters which are used in proving meters under test. In networks, critical nozzles can be used as flow limiters since once sonic flow is established only altering upstream pressure P_u can change the flow rate
Pitot tubes	The pitot tube employs the principle that if it is placed faced into a stream of fluid, then the impinging fluid will come to rest; thus, its kinetic energy will be converted into pressure. The pressure built up in the tube will be greater than in free stream by the 'impact pressure' due to the squared velocity in the stream. Thus, a difference is measured between pressure in the tube and static pressure of the stream. Static pressure is either measured by a tapping in the wall or by tappings in the Pitot-static tube. A pressure difference is obtained which is a measure of the 'impact pressure' and therefore of the velocity if the fluid at that point ±5 % [8]	A: Introduce negligible flow pressure loss. Cheap and easy to install D: Instrument for making temporary measurements of flow	Not permanently used in industry. Used for velocity and direction measurements in combustion studies and heat transfer work. The commercial 5-hole pitot tube is used greatly with furnace studies where recirculation and turbulent zones are present; thus, gas flow is unknown. With the 5-hole Pitot tube, both magnitude and direction of the gas velocity can be obtained

(continued)

Table 5.13 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Elbow meters	Due to the elbow bend and consequently to the centrifugal forces differential-pressure is generated between the inner and outer radii of the bend. In the case of gases because the differential-pressure is very low and a venturi can assist to increase it. Less accurate	A: Space saver. Pressure loss is significantly less than orifice plates	Compression station with space limitation
		D: Accuracy. Noise. Sensitive to upstream conditions	
Rotameters (variable area meters)	In its simple version, the device consists of a vertical tube with a bore tapered wider at the top containing a free piston (float). The flow in the instrument is vertically upwards and as the flow increases, the piston rises in the tube until a point of equilibrium is reached. In other words, the upward thrust of the piston generated by the differential-pressure of the metered fluid passing through the annular space balances the weight of the piston. Range: 10:1 sizes up to glass: 4 kg/s for liquids. Air flow: 0.05 m^3/s. Accuracy: up to ±0.5 %	A: High-accuracy metering-corrosive liquids and gases. Easily installed delivered calibrated from manufacturer for their intended use	Used in laboratory, development and industrial applications for monitoring flow rate of gas and air supply to burners
		D: If used for metering fluids with different densities, they must be corrected	

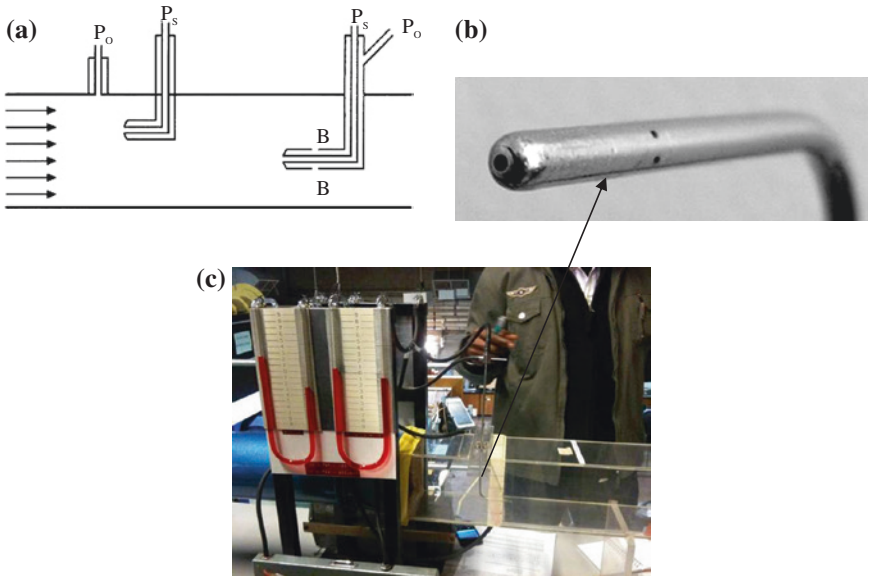


Fig. 5.24 Schematic of pitot tube (**a**) and illustration of pitot-static tube (**b**) and laboratory gas flow measurement (**c**)

5.3.6 Elbow Meter

Flow rate in this meter is measured by determining the differential-pressure developed between the inner and outer radii of the bend by means of two pressure taps located midway on the bend as shown schematically in Fig. 5.25. See also Table 5.13 for more details.

5.3.7 Variable Area Meters (Rotameters)

Employ the principle that the differential-pressure is maintained constant and the magnitude of the variation of the cross-sectional area of the flow is a measure of the flow rate as shown in Fig. 5.26 (see also Table 5.13).

5.3.8 Positive Displacement Meters

5.3.8.1 The Impeller (Lobe)-Type (Roots Meter) and the Rotary Vane-Type (Rotor-Seal Meter)

BS1179 defines a rotary displacement meter as 'a positive displacement meter in which the measuring compartment is formed between the walls of a stationary

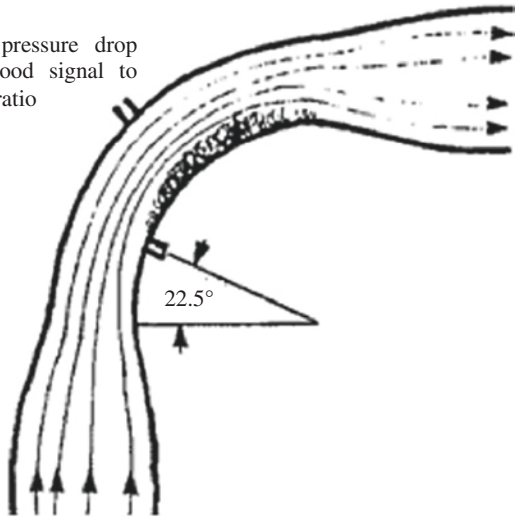


Fig. 5.25 Schematic diagram of typical elbow meter (venturi type)

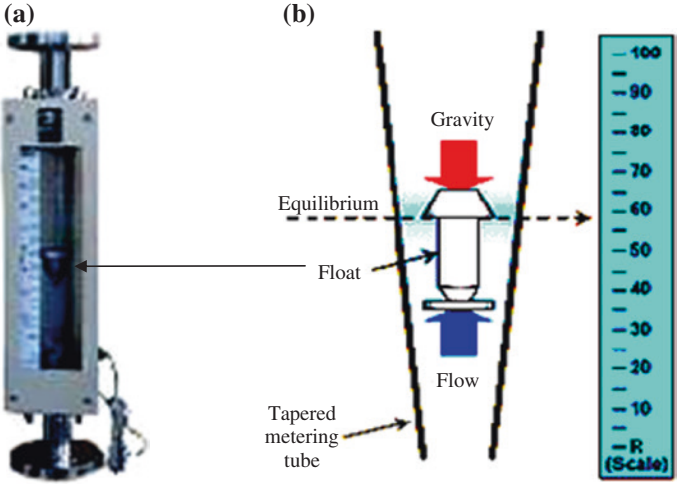


Fig. 5.26 Schematic digram of variable area meters (**a**) and a typical instrument (**b**) [9]

chamber and a rotating element or elements making substantially gas-tight contact with the walls'. They have been used in the industry for more than 70 years and account nearly 10 % [8] of the total number of flow meters in service. Positive displacement meters use mechanical parts to displace the fluid at measure in discrete known volumes. The impeller (lobe)-type (roots meter) and the rotary vane-type (roto-seal meter) meters are suitable for metering gas as shown schematically in Fig. 5.27.

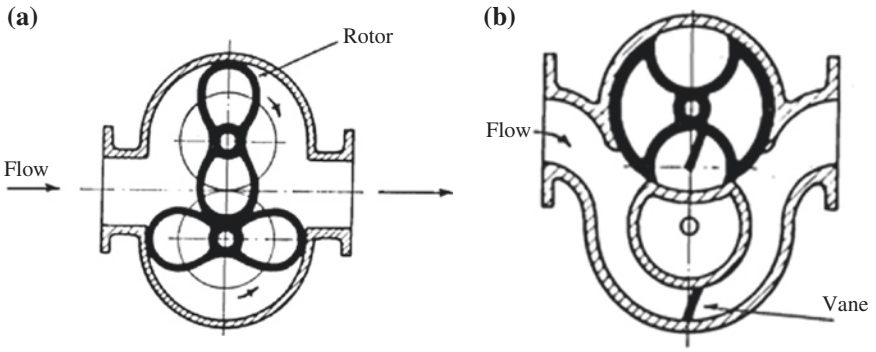


Fig. 5.27 Schematic diagram of lobe type (**a**), vane type (**b**) positive displacement matters

The meter's parts such as casings, impellers, rotors and shafts are of cast iron/steel, aluminium for impellers and if required material coating can be used. They achieve almost constant differential-pressure at low speeds, low throughputs and low pressures. A 'hand slip calibration' is possible for obtaining the relevant correction slip curve by passing air through a blocked meter and measuring it. The accuracy of the meter depends on how specification designs where followed, on the density and viscosity of the gas since they contribute to slip effects, on the loading and on outside influences such as pipeline configuration associated with resonance effects with specific loadings.

As stated before high-pressure meters are available through distortion of casing and assembly must be prevented. Thus, case increased thickness, and ribs to strengthen the casing are used together with pressure relieving slots. Sealing is vulnerable at high-pressure operations. See also Table 5.14 for more details for these meters.

5.3.8.2 Diaphragm Meters

Another type of positive displacement meter is the diaphragm meter shown in Fig. 5.27 which has a gas pressure difference that has a global impact. It is the commercial meter used for gas consumers, both domestic and industrial. The readings from this meter determine the charge made to the consumer and thus the revenue of the gas company. Due to its importance, it must meet national and international standards.

Contemporary designs have been proposed for the reduction of pressure absorption resulting from breakdown of lubrication from the meter's moving parts, such as valves resulting in an erroneous meter registration. Different casings are available according to the pressures required. Remote reading capability is also

Table 5.14 Rotary impeller, diaphragm and wet gas meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Rotary impeller (lobed) type (roots meter)	Two impellers contra rotate at the same time, intermeshing closely at all times, and thus gas is passed forward from the inlet to the outlet through the measuring chambers. The impellers are driven round by the gas due to the unbalanced load on one or the other impeller when pressure is applied at inlet. The timing gears ensure that the other impeller will turn in the opposite direction and thus become out of balance. In this way, the cycle is continuous Range: Up to 30:1. And up to 100 bar pressure. Accuracy: ±1 % from. 10 to 100 % of maximum flow	A: Accurate and reliable. Budget affordable. Accuracy maintained throughout the ranges. Documented EEC 71/318 and BS1179. Pulse count method suitable for remote data processing. No issues with gas supply security and safety. Problems may be reduced by regular inspection, spin tests and by the use of flow straighteners D: Moving parts thus friction. Viscous drag at clearances. Pressure loss since energy is needed to rotate the impellers. Requires calibration for handling slip. Noise and wear on gears and blade damage mainly from debris. Jamming: when debris is trapped between impellers causing flow restriction, A filter in inlet and an automatic meter bypass can provide remedy. Control equipment such as governors and pilot supplies may be affected, pulsation and impeller inertia causing slower impellers response time when having rapid variable loads. This may also lead to higher meter readings. Pipe (acoustic) resonance with pipe work resonant frequency may occur. Careful design eliminates this possibility. Swirl which produces an error dependent on swirl angle	Used in oil and gas industry for many years for measuring flow due to their accuracy and reliability. The available range covers the gas operating pressures with constant accuracy even with low flows and low pressures which is the case in domestic metering. Extensively used for low-pressure measurement

(continued)

Table 5.14 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Diaphragm meters	There are four chambers in the meter which empty and fill in sequence. Since the exact volume of the chambers is known, by counting the number of displacements, the volume of the gas is measured. Entrance and exit of the gas to the compartments are controlled by sliding valves. Compartments 2–3 are encased by bellows which expand and contract	A: Available in ranges and high reliability. Can be adjusted for remote data transfers or pre-pay methods. Documented and approved rated under national and international standards: BS4161 parts 1–5 for specifications, BSCP part 2 for installation, IGEM also produces recommendations	For domestic and commercial selling of gas due to the wide range available that suits the nature of different customers and also due to the strength and durability of the meter
		D: Pressure absorption causing diaphragm to balloon out	
	R: Tinplate: pressure: up to 5 kPa and up to 35 kPa special range (with up to 170 m^3/h capacity). Steel case: Pressure: up to 25 kPa. Capacity: up to 160 m^3/h. Aluminium die cast: Pressure up to 7 bar Capacity: up to 280 m^3/h		
Wet gas meter	It consists of a rotating drum semi-bathed in a fluid such as water or oil. The drum is divided into chambers of known volume, and thus, as the drum rotates, the gas is charged and discharged from inlet to outlet following the designed route through the chambers and the liquid. In order to achieve the specified accuracy, the speed of the rotating drum must be kept low in order not to disturb the water level which is critical. The meter always must stand level. Range: 10:1. Capacity: 0.5 m^3/h. Accuracy: up to 0.25 %	High accuracy. No problems with sealing moving parts. low capacity meter. Vulnerable to humidity. If dry air or gas passes through meter, they will become saturated and in order to avoid this suitable oil instead of water is used	Used as a reference standard for calibrating other types of meter, for laboratory use

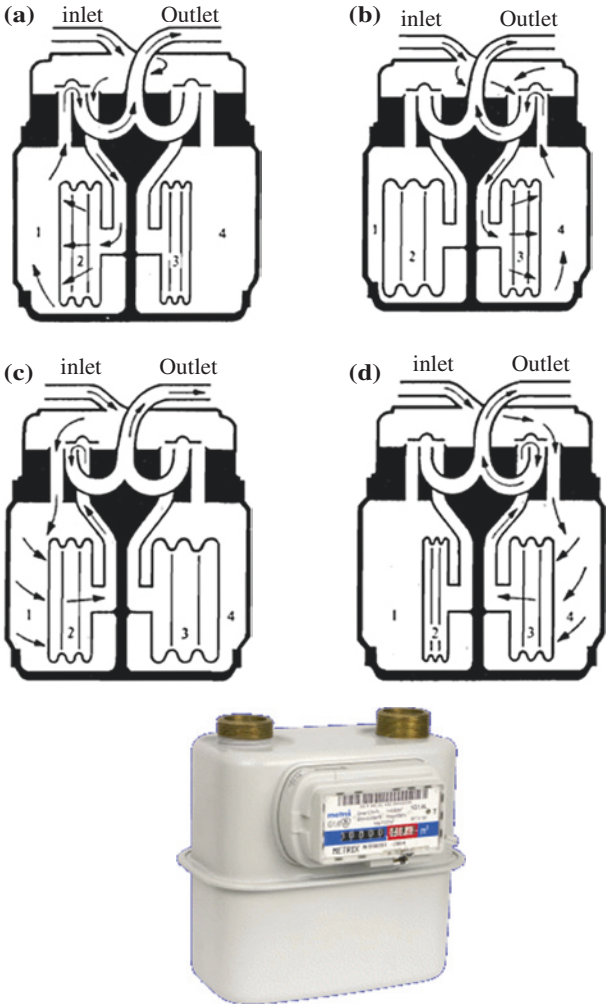


Fig. 5.28 Schematic operative features of diaphragm meter [10]. **a** Chamber *1* is emptying, *2* is filling, *3* is empty and *4* has just filled. **b** Chamber *1* is now empty, *2* is full, *3* is filling and *4* is emptying. **c** Chamber *1* is filling, *2* is emptying, *3* has filled and *4* has emptied. **d** Chamber *1* is now completely filled, *2* is empty, *3* is emptying and *4* is filling

available in the present days if any remote reading is required. Also the latest feature is the meter can accept prepaid electronic tokens such as the British Gas Quantum System that accepts a reusable gas card (see also Table 5.14 for more details) (Fig. 5.28).

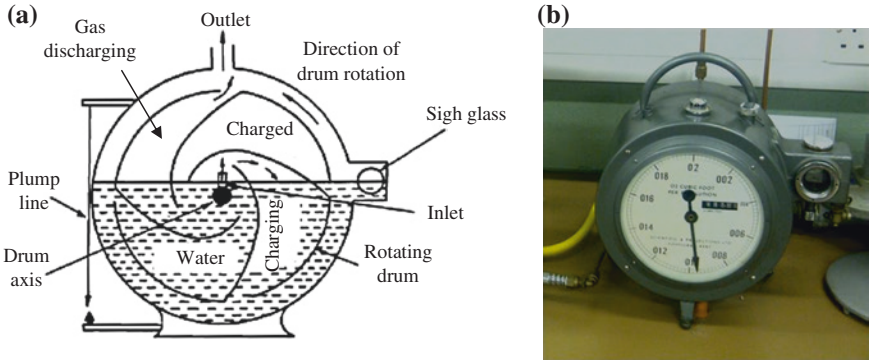


Fig. 5.29 Schematic view of wet gas meter (**a**) and typical illustration of the meter (**b**)

5.3.8.3 Wet Gas Meters

The last member of the positive displacement meters is the wet gas meters, which solves the sealing of the moving parts through its principle of operation. Figure 5.29 and Table 5.14 provide more details for these meters.

5.3.9 Rotary Inferential Meters

5.3.9.1 Axial (In-Line) Turbine

An axial turbine is a velocity device for measuring volume as shown in Fig. 5.30. The direction of flow is parallel to the rotation axis of the rotor and the speed of

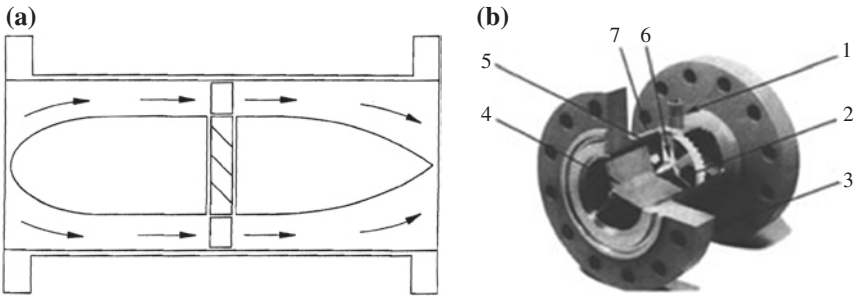


Fig. 5.30 Schematic diagram of axial turbine (**a**) and typical-type illustration of the meter (**b**) [11]. *1* Pickoff, *2* Bearings, *3* End connections, *4* Supports, *5* Rotor blade, *6* Rotor hub, *7* Meter housing

rotation is proportional to the flow rate. For gas measurement, the meter must have low non-fluid retarding torques so that rotor percentage slip is acceptable. Retarding torque is due to: (a) non-fluid forces such as bearing friction and mechanical loading from various instruments used whose contribution in the torque can be assumed constant, (b) friction drag due to turbulence can generate a constant percentage slip on the flow range of the meter (see also Table 5.15).

Table 5.15 Axial, insertion turbine and roating-vane meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Axial (in-line) turbine	A meter consists of a body, a rotor mechanism with inlet and outlet diffusers, a mechanical index and a magnetic drive or pulse counter. The flow of gas causes the multi-blade wheel to rotate at a rate that is proportional to the volume flow. Thus, the meter's driving energy from the kinetic energy from the gas that is caused to increase as it passes through the meter by reducing the cross-sectional area of the flow stream through diffusers. The rotor is constructed with the maximum effective radius to create the highest torque for a given force. Bearing friction is an important parameter; thus, meters for gases are less linear than liquid ones. Flow rate, bearing design, blade edge sharpness, roughness and nature of the flow have different effects on linearity; thus, calibration is needed for establishing meter factors. The maximum capacity rating is established through the pressure across the meter, the end thrust of the rotor shaft and bearing caused by the differential-pressure and the rotational frequency of the rotor assembly. Capacity: 50 mm /14–113 m^3/h. Up to 1015 mm /45–300 m^3/h. Pressure: 2.5–100 bar. Accuracy: Up to ±1 %	A: They have small size, large volume capacity and range ability suitable for measuring large flows. Compact and light in weight; thus, movable do not need large foundations. They can be supplied with correcting devices. Self-adjusting models such as the Rockwell type where effects of bearing wear are compensated through self-correcting and self-checking	Their advantages make them suitable for transmission distribution or industrial applications
			The self-adjusting Rockwell type is attractive for offshore applications where flow-metering space is limited. This is because it is less affected by flow disturbances such as swirl
		D: Greatly affected by any debris or matter in the metering gas or fluid. Bearing wear causes pressure loss Large errors in case of two-phase flow	

(continued)

Table 5.15 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Insertion turbine	The main parts of this type of meter are the rotor cage that is subject to the application with the leg. The insertion tube with the pressure relief chamber and at the top the M 700 connection with the wiring that runs all over the length of the meter. Sizes of rotor cages must be compatible with 100-mm-diameter pipe assembly. The calibration must be unaffected by the range of operating densities and insensitive to small misalignment. Required velocity range: 0.3–15 m/s	Ease of insertion Portability and robustness. Simple to use and operate. Meter gives direct reading or signal output for a wide range of mains	Used to measure directly the rate and direction of gas flow in individual mains. Used to validate network analysis models Used to establish boundary conditions, Used to identify blocked or unconnected mains
Rotating-vane gas meter	The meter consists of cast iron housing with an aluminium anemometer fan. As it rotates, it transmits the drive to a counter-mechanism that integrates the flow. Capacity: 4–200 m^3/h. Pressure: up to 1.7 bar About ±2 % over 10:1 range	It is a rugged construction	Suitable for most non-corrosive gases
Rotating-vane air meter (anemometer)	In this type of instrument, the vanes must be light or have a large surface area. The rotor may be of two kinds: In the cup anemometer, it consists of three light conical-shaped cups made of aluminium attached to arms on a vertical spindle geared to a counting mechanism. In the air meter type, the rotor has a multi-vane assembly made of aluminium alloy carried on a shaft with its other end attached to a counting mechanism. Velocity: 0.13–30, 0.3–25 m/s	Used in meteorology. Moving parts	Various

5.3.9.2 Insertion Turbine

It was developed to measure directly the rate and direction of gas flow in individual mains as shown in Fig. 5.31.

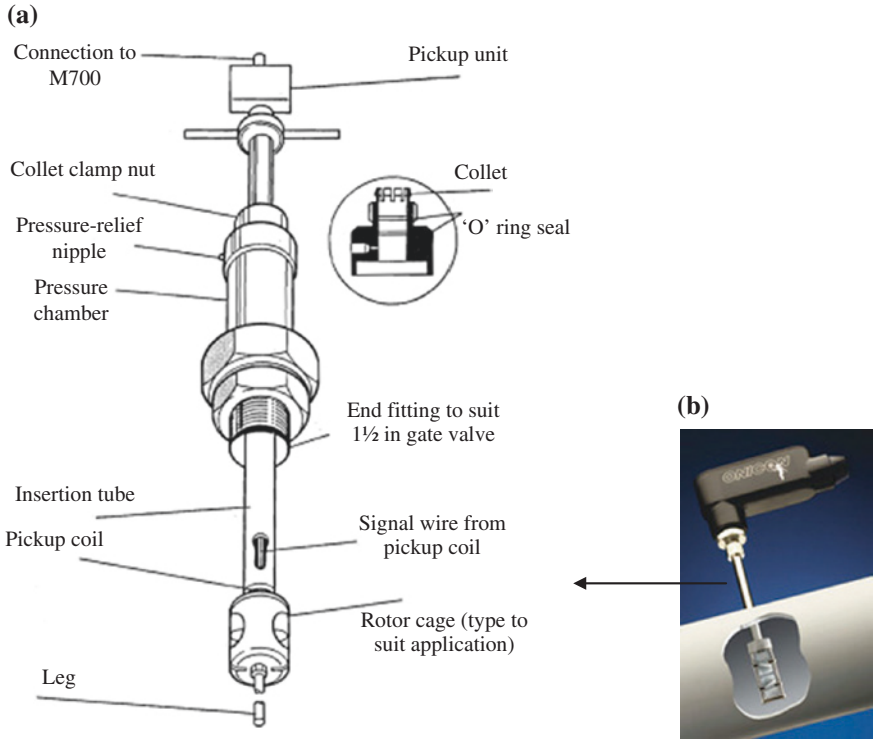


Fig. 5.31 Schematic of insertion meter (**a**) and typical illustration of the meter (**b**)

5.3.9.3 Rotating-Vane Gas Meter and Air Meter (Anemometer)

The rotating-vane gas metre is shown schematically in Fig. 5.32, which is basically an aluminium anemometer fan rotates on a vertical shaft as fluid flows and

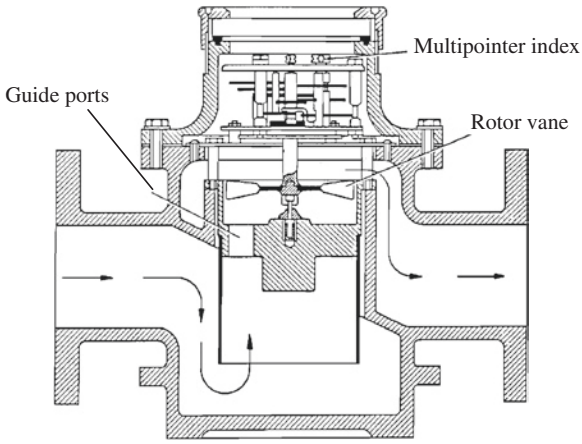


Fig. 5.32 Schematic diagram of rotating-vane gas meter



Fig. 5.33 Typical rotating-vane air meter

a counter-mechanism subsequently integrates the flow. Table 5.15 also provides more information on this instrument.

The rotating-vane air meter (see also Fig. 5.33) is an instrument which can measure velocity, but it can also be designed to measure flow rate of air and gases directly. This can be done by attaching a capacitance transducer on the outer ring of the anemometer where as the vane rotates an electrical pulse output will be taken up by the indicator unit producing a direct reading. The rotor and the whole assembly must be balanced and the bearing selection must minimise friction as much as possible. The vanes of the assembly must be light or have large surface area (see also Table 5.15).

5.3.10 Fluid Oscillatory Types

5.3.10.1 Fluidic Meters

This instrument detects flow from the oscillatory movement of a jet of a gas within the meter. The instrument produces the oscillatory movement by design since it is a fluidic feedback oscillator based on the Coanda effect, i.e. the tendency of the fluid to be attracted to a nearby surface as shown schematically in Fig. 5.34. The stream is guided to oscillate from side to side at a frequency proportional to the flow rate. Table 5.16 provides more detail of the various types of fluid oscillatory meters. These meters are also summarised in the following sections.

5.3.10.2 Vortex Type

The phenomenon of vortex shedding provides the operating principle of this relatively new type of instrument, as shown schematically in Fig. 5.35 together with Table 5.16 which provides further details for the meter.

5.3.10.3 Swirl Meters

This type of meter produces swirl, and with a combination of a venturi and a diffuser, a flow rate relationship can be established as shown in Fig. 5.36 (see also Table 5.16).

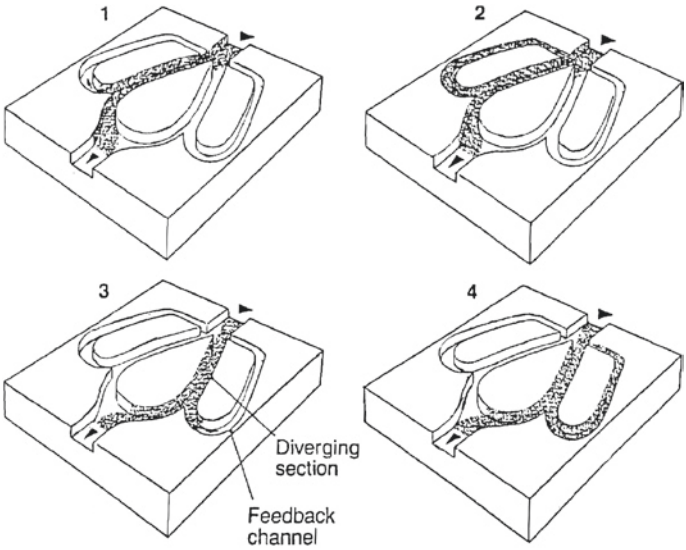


Fig. 5.34 Schematic diagram of fluidic meters. *1* Gas entering. *2* Diversion of gas to inlet. *3* Gas moves to opposite side. *4* Diversion of gas to opposite wall

Table 5.16 Fluidics, vortex and swirl meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Fluidic meter	As gas enters the upstream port it is attached to one wall. A small part of this stream is diverted into a loop back to the inlet where it pushes the main flow onto the opposite wall, and thus, another feedback is generated through an opposite route. This is how the stream oscillations are produced which has a frequency proportional to the flow rate. Oscillations can be detected by a thermistor or a strain gauge transducer placed in the feedback routes. One output pulse will be read for each cycle and thus the calculation of flow rate	A: No moving parts thus no pressure loss due to friction of moving components. The pulse signals from the detecting device can be transmitted to the gas company through existing telephone lines	Strong candidacy for future domestic gas meter. The main issue is that their size must be brought around 'brick' size so that they can be installed in house walls
		D: Unable for a single meter to cover the required flow range-a separate low flow sensor will be needed. Efforts are being made to develop one sensor suitable for both low flows and oscillations	

(continued)

Table 5.16 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Vortex meters	A bluff body is placed in the flow. At very low flows the streamlines pass smoothly round the object but as the fluid velocity increases the fluid separates from the body forming vortices in the low-pressure region behind the body. The frequency at which these vortices are shed is proportional to fluid velocity, and thus, the flow rate can be calculated. In gas flows sensitive thermistors are used to detect vortex frequency. Range: 50:1, Accuracy ±1 %	A: No moving parts thus high reliability. No maintenance ire required and their cost is similar to an orifice plate [8]. Frequency is almost independent of the fluid used so calibration to establish meter factor can be done using water. Less sensitive to dirty gas and upstream disturbances than orifice plates	An alternative to pressure differential meters. Retrofix vortex meter under development by British Gas allows an orifice to be converted into vortex shedding improving flow capacity of the system
		D: Interaction of vortices with acoustic waves which have negative impact on accuracy. Dependent on bluff body size limitation	Insertion meters version can be used as a single point velocity tester through a small hole cut in the pipeline
Swirl meters	Fixed inlet guide vanes force the entering gas into a swirling motion. Then the gas accelerates through a Venturi and then decelerated through a diffuser. This sequence causes the axis of the swirling gas to process about the pipe centre line. The frequency of precession is proportional to the gas flow rate. A thermistor in the throat casing detects the passage of the coils of the helix. As it leaves the meter the flow is expanded and diswirled by a set of vanes	A: Insensitive to upstream conditions.	
		D: Quite high pressure loss thus not suitable for low-pressure metering	

5.3.11 Ultrasonic Meters

5.3.11.1 Doppler Type

If a sound of a known frequency is reflected from a moving object, the frequency of the reflected beam is altered by an amount that is proportional to the velocity of the moving object. Thus, the flow rate is calculated through the Doppler shift frequency. Figure 5.37 shows the principle of operation of the industrial ultrasonic meter. Table 5.17 also provides more details with regard to this instrument.

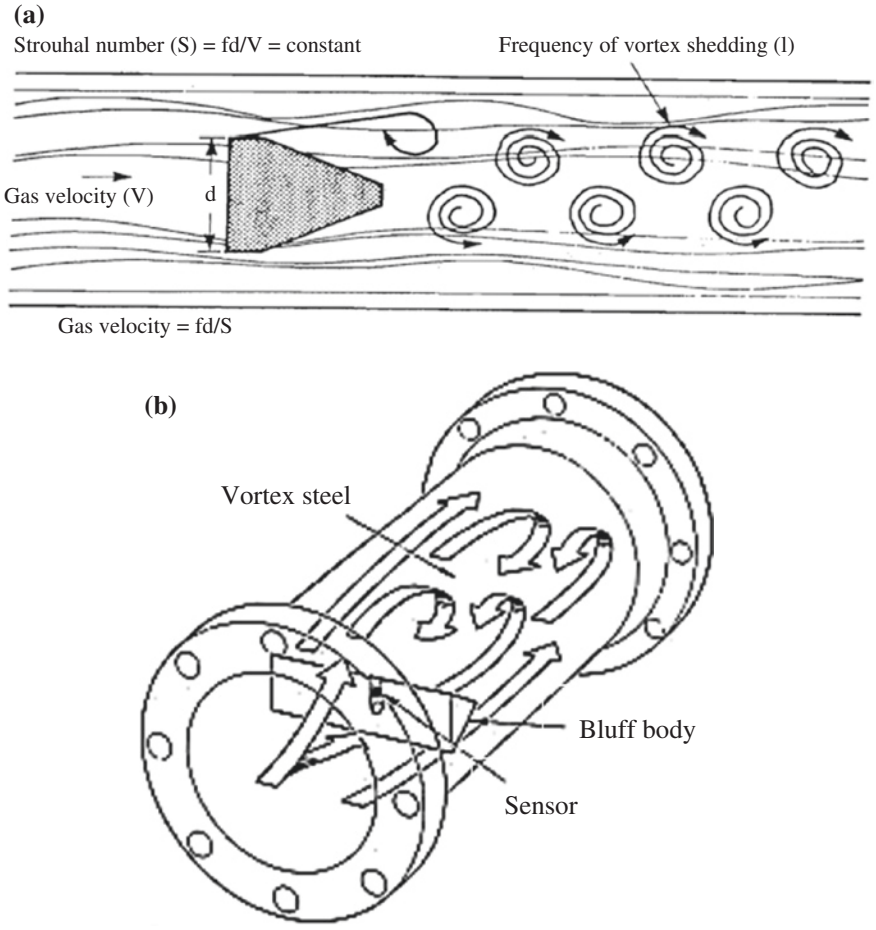


Fig. 5.35 Schematic of flow characteristic of (**a**) and typical diagram of the vortex meter (**b**)

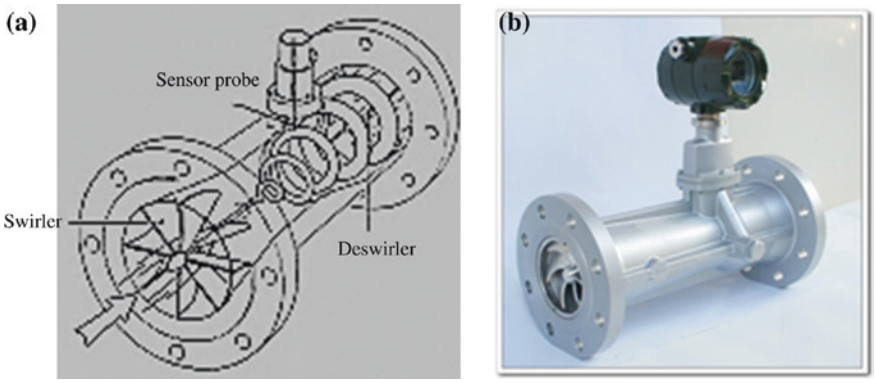


Fig. 5.36 schematic diagram of swirl meter (**a**) and a typical illustration (**b**)

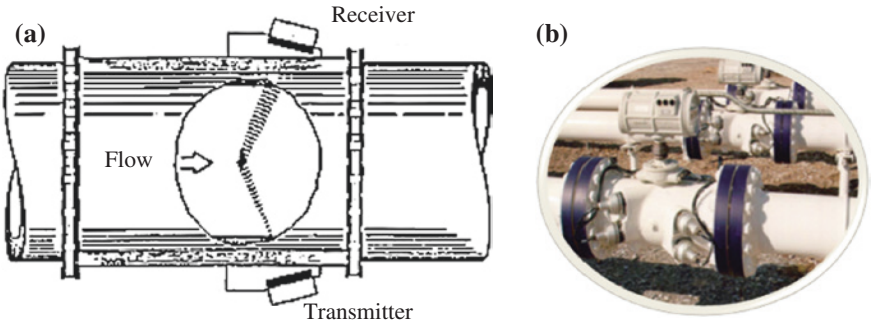


Fig. 5.37 **a** Principle of operative. **b** Typical industrial ultrasonic meter

Table 5.17 Ultrasonic meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Doppler type	The instrument consists of a transmitter and a receiver of ultrasonic waves. The scattering elements which in the case of gas can be either gas particles or eddies deflect the ultrasonic energy from the transmitter. This provides a frequency shift between the deflected energy and the energy from the transmitter; thus, the velocity of the fluid can be calculated. Since velocity profiles are present, the received signal will be a range of frequencies proportional to the prevailing flow profile conditions; thus, meters are designed to process the strongest signal through conditioned amplitude modulation and demodulation in order to give an output proportional to flow [8]	A: High reliability and low maintenance costs. Can be attached easily and externally without disturbing the pipeline flow operation; thus, personnel do not come into contact with the flowing medium that can be hazardous. In modern days, ultrasonic transmitters and receivers are cheap, and thus, the meter is relatively inexpensive	Useful or measuring the flow of corrosive fluids and slurries
		D: The flow profile pipe-wall thickness constancy number, size and distribution of the particles in the flow. Thus, there is great accuracy variance and to avoid it a carefully and rigorous calibration must be carried out	

(continued)

Table 5.17 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Time-of-flight type	The instrument consists of a pair of ultrasonic transducers mounted at an angle of about 45° to the axis of flow. Each transducer is a pair of transmitter and receiver and is usually piezoelectric elements. Fluid flow in the pipe causes a time difference in the reception of the acoustic beams which is measured as a difference in transit times and is proportional to the velocity of the flowing fluid. If the effect on the transmission properties of ultrasound of various gas compositions is known, boundary operational parameters can be established. In small bore pipes, a single beam is transmitted axially and the domestic meter proposed for this type utilises this where the gas velocity is calculated from transient times measured along and against the flow. Range: 100:1. Accuracy: ±1 %	A: Clean from debris trapping. D: Sensitive to velocity profile and is remedied by having four paths of beams. Power consumption	Measure the flow rate of clean liquid and gases Domestic versions have been proposed
Coriolis mass meter	The flow meter main component is a pair of parallel vibrating tubes or a single vibrating tube that is formed into a configuration that has two parallel sections anchored at 2 points. The Coriolis forces are generated by this vibration which must be at a distance from the fixed points. The measurement of the Coriolis force is related to the mass flow rate by Newton's second law. Considering a U-shape and points A at inlet, B exactly at the bend and C at the outlet of the tube. Between A and B, gas molecules experience increasing forces at right angles to their motion where as from B to C decrease since vibration is maximum at B and zero at C. The above motion generates a couple which causes the tube to twist relative to the horizontal plane (motion plane) and the maximum angle of twist θ is proportional to the mass flow within the tube. The angle is sensed by displacement transducers located at the 2 points of maximum deflection. Range: very low flow ranges. Accuracy: very high accuracy ±0.2 % [8]	A: High accuracy. Low maintenance requirements. D: Expensive. Failure possibility of tube from fatigue. High pressure drop across the meter. Low flow ranges	Primarily used for measuring mass flow of liquids. The high accuracy makes it a suitable high precision laboratory instrument especially with gas at high pressure (accuracy improves with high density Low maintenance requirements

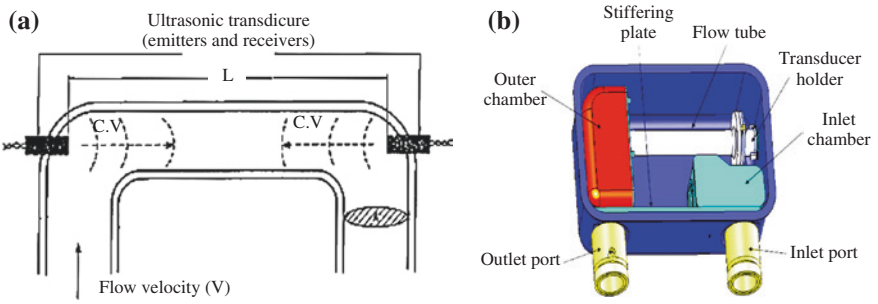


Fig. 5.38 Ultrasonic meter **a** and **b**

Figure 5.38 also shows an illustration of a domestic ultrasonic meter. A pair of transducers are mounted at opposite ends of a length of tube carrying the gas flow. The gas velocity can be calculated from a pair of transit times measured with and against the flow together with some geometrical constants. The accuracy of this calculation is not significantly affected by changes in gas properties or temperature. Almost all these types of meters incorporate sophisticated hardware and software.

5.3.11.2 Time-of-flight Type

This type relies on the measurement of the time difference between an ultrasonic pulse travelling with and against the flow in a pipe as shown in Fig. 5.39 (see also Table 5.17 for further details).

5.3.12 Direct Mass Type

5.3.12.1 Coriolis Mass Meter

Coriolis mass meter is the only representative from the direct mass-type family. The principle behind this instrument is Coriolis force. Such forces are generated

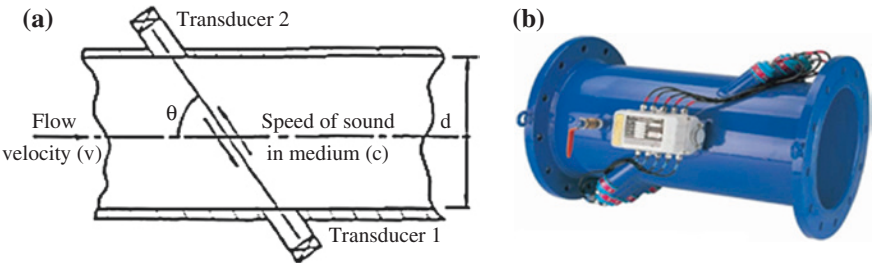


Fig. 5.39 Principle of time-of-flight flow meter (**a**) and illustration of the meter (**b**) [12]

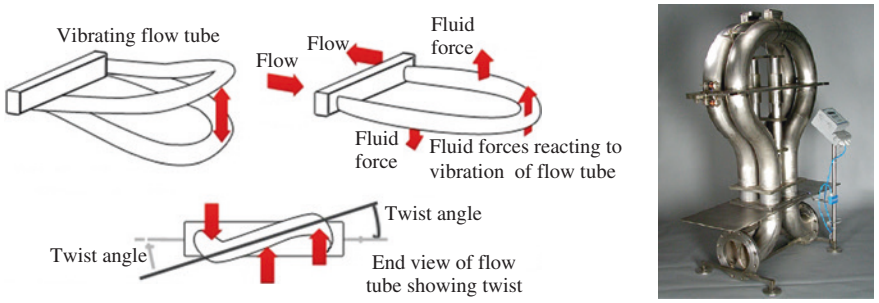


Fig. 5.40 Features and principle operative of coriolis mass meter

whenever a body, which is rotated about a fixed point, undergoes a change of position relative to the fixed point as shown in Fig. 5.40 (see also Table 5.17).

5.3.13 Thermal Types

5.3.13.1 Hot-Wire Anemometer

The physical principle behind this type of meter is that when a fluid flows over a heated surface, there is heat transfer from the surface and a consequential reduction in temperature, which is related to the rate of flow. Figure 5.41 shows the principle of the operation the hot-wire anemometer with further details provided in Table 5.18.

5.3.13.2 Probe Type

This type of meter senses changes in flow and associates them with a change in DC output. Figure 5.42 shows the typical illustration of instrument and Table 5.18 provides further details.

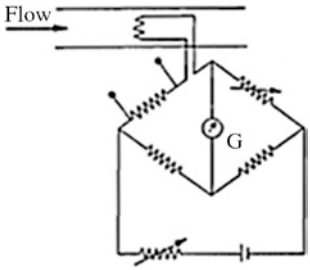


Fig. 5.41 Schematic diagram of hot-wire anemometer

Table 5.18 Thermal-type flow meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Hot-wire anemometer	Heat is supplied electrically to a wire placed in the flow stream. The temperature of the wire is obtained by measurement of its resistance using a Wheatstone Bridge. Current adjustment is required so that temperature remains constant and measuring the heating current so that the bridge remains balanced. A potentiometer is used to measure the potential difference across a standard resistance in the appropriate arm of the bridge. At equilibrium, heat supplied equals heat lost, and thus, if resistance and temperature of the wire are kept constant, a relation is established with the velocity of flow and the fourth power of the current. Wide range ±2 % [8]	Low flow rates	Suitable to measure air or gas flow Useful for laboratory and research applications
Probe type	The instrument consists of a small hand held probe made of stainless steel and containing the two sensing elements, the thermocouples. Alternative current is used to heat the elements and a change in flow results in a change in direct current (DC) output. A third unheated thermocouple (identical to the other two) is added to the dc output and this is to compensate the transient heat generated when ambient temperature changes. Range: Up to 30 m/s	A: Measuring low velocity flows 0.05–0.5 m/s due to its dense spaced scale at lower velocities	Measurement of air velocities through supply grills and ceiling diffusers in air-conditioning and ventilating systems



Fig. 5.42 Typical probe-type flow meter

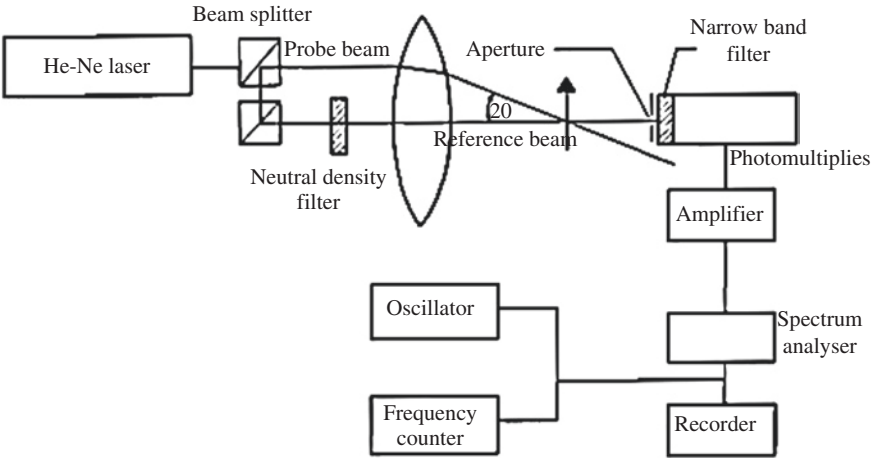


Fig. 5.43 Schematic diagram of laser doppler flow meter

5.3.14 Miscellaneous Techniques

5.3.14.1 Laser Doppler Meter

Measures the Doppler shift of laser radiation scattered from particles moving within the gas stream. Figure 5.43 shows the principle operation of the instrument, and Table 5.19 gives further details.

5.3.14.2 Cross-Correlation Meter

An ultrasonic arrangement that correlates data from two pairs of transmitters and receivers as shown schematically in Fig. 4.44 (see also Table 5.19) (Fig. 5.44).

5.3.14.3 Tracers

They inject a substance into the pipeline gas stream and then a characteristic related to the flow rate is measured. For the application of this technique, two methods are available. The first one is the dilution method where the flow rate is determined by injecting a known quantity of tracer into the pipe and measuring the dilution ratio at a downstream point. The second one is the transit-time method where a pulse of tracer is injected and a measure is taken for the time for the tracer to travel over a known volume section. Figure 5.45 shows schematically the operation features of the meter. Table 5.20 also gives more details of the instrument.

Table 5.19 Miscellaneous flow meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Laser Doppler meter	The main instruments of the meter are a laser source, beam splitter, filters, an optical system and a series of photomultiplier, amplifier, analyser and recording devices	A: Needs no calibration. Linear frequency/velocity relationship over high range. High accuracy, reproducibility and does not disturb the flow, thus no pressure loss due to the meter device D: Expensive and fragile complex and a need for skilled operation. High performance is subjected to 'clear windows' in pipeline	Fluid and combustion research Detect water vapour in gas flow or other debris A standard on which other metering devices can be calibrated
	The laser beam is split into two components, and using an optical system, they are focused to intersect the gas stream. Any particle in the gas will be trapped between the two beams, and a phase shift will be recorded in frequency which is associated with the particle's velocity. Small enough particles are selected in order to follow gas flow with great accuracy. The optical system is designed to allow velocity profile across pipes to be obtained, and thus, the total flow can be calculated. Frequencies output linear with velocities from 0.1 mm–3 Mach. Accuracy: ±0.5 %		
Cross-correlation meter	Two transverse beams of ultrasound are sent across the pipe diameter. If no flow occurs, the beam's signals are the same as the ones transmitted. In the case that flow exists, the signals received will be modulated because of the existence of particles in the flow stream or eddies in the case of turbulent flow. These modulated signals will be fed into a data processor called a cross-correlator where by comparing the two signals, the transit time will be obtained and hence the velocity. In reality, meters are designed to give the average velocity across the diameter	Very expensive by industrial standards	Two-phase and flare gas applications

5.3.15 Flow Meter Systems

5.3.15.1 Metering Large Industrial and Commercial Loads

Since gas industries have developed, there is an increase in large measurement systems (>300 m^3/h volumetric flow rate) as well as having an increase in measurement at high pressure. Concerning statutory requirements and standards, these are covered by national standards organisations, officially recognised and backed by the government such as the British Standards (BS) or the German Din, and in addition, industry-orientated national bodies also exist such as the Institution of

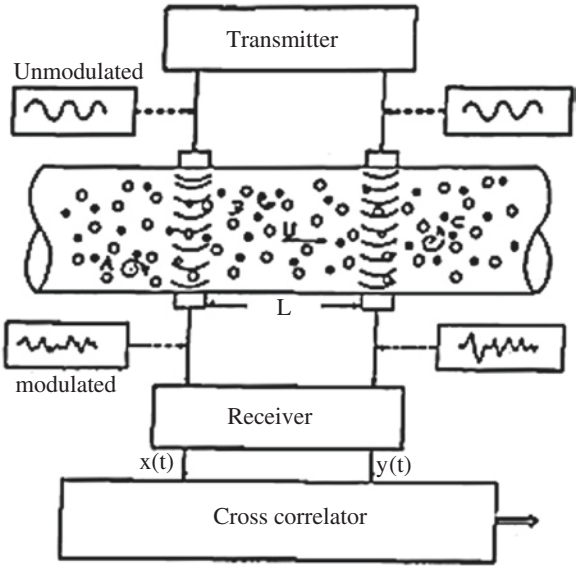


Fig. 5.44 Schematic diagram of cross-correlation flow meter

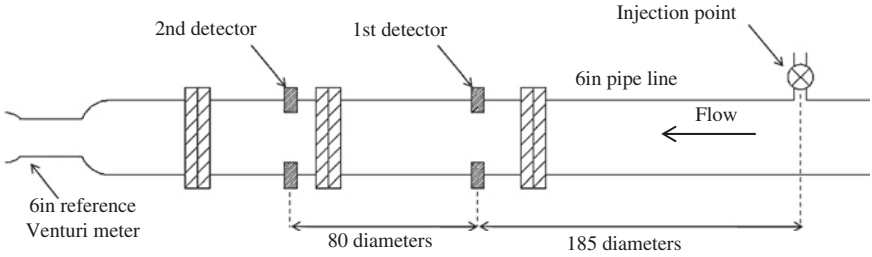


Fig. 5.45 Schematic features of tracer flow meter

Gas Engineers and Managers (IGEM). At international level, the following organisations issue standards relating to gas metering:

OIML The International Organization of Legal Metrology
ISO The International Standards Organization
IEC The International Electro-technical Commission
EEC European Economic Community

In the UK, legal requirements exist for metering systems concerning safety and accuracy. The safety aspect is covered by Gas Safety Management Regulations 1996 (GS(M)R), whereas the accuracy is covered aspect by the relevant Gas Act and the Gas (Meter) Regulations. Meters used for charging must be stamped or 'Badged' except when special terms are agreed between supplier and customer. Stamped

Table 5.20 Traces flow meter

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Tracers	*Constant rate injection method*	Mobile systems and accurate	Suitable for detecting significant deviations in meter accuracy
Dilution techniques	A tracer of known volume is injected into the stream at constant rate, and after stream and tracer concentration is fairly mixed, a measurement is taken at a downstream point. The volumetric flow rate is determined by relating background, injection and mixed flow tracer concentrations		Also can be used for calibration and proving
	Pulse injection method		
	The difference is that the tracer is suddenly injected into the gas stream. After complete mixing, the concentration profile for the total cloud of injectant is measured downstream. The volume flow rate is obtained by integrating on the concentration profile. Accuracy: ± 1 %		
Transit-time techniques	Gas is suddenly injected into the gas stream and the time for tracer cloud to pass between two detectors is measured. Due to turbulent flow, the transit time is the same as the mean residence time of the tracer in the pipe volume between detectors. The tracer gas can be radioactive or chemical (Ethyl bromide and Argon 41). Multi-channel UV recorders with crystal oscillator timing pulses are used to time the tracer pulse travelling the known distance between detector points. The volume of the measuring section is a very important parameter and BS5857 recommends its determination before assembly, external measurements of the pipe geometry such as wall thickness with X-ray or ultrasonic methods can be deployed in determining the volume. Accuracy: ±0.5 %	Three small diameter tappings are required, and the exposed length is the least for any tracer method. Short half-life isotopes and the great dilution of the tracer gas in the pipeline ensure negligible radioactive hazards towards consumers. Scintillation counters can be strapped to the pipe externally, and therefore, pipe needs no modification. Use of graphical output from counters in addition with optimised statistical analysis to achieve maximum accuracy. Use of symmetrical array may reduce need for complete mixing since sensing is present to all parts of the pipe's cross section. Also length between injection point and first detector can be reduced. Difficulties when using radioactive fluorescent or chemical as tracer gas since fluorescent requires, a transparent window in the pipe and chemical tracers require identical sampling points in order to have identical response times. Isotopes need to have powerful γ rays to penetrate the steel pipe, so detection can be feasible	Gas industry and over each laboratory

meters must be accurate within ±2 % and maximum pressure drop is specified for diaphragms meters, whereas no statutory pressure drop is cited for rotary displacement meters. International and national bodies cover meter specification as follows:

OIML	Produces standards for most types of gas meter
BS4161	Diaphragm, rotary displacement, turbine meters and also mechanical volume correctors
BS1042 and ISO 5167	Orifice plates

For gas supplied under the terms of published tariffs, the following must be met:

1. Compulsory use of stamped meter either by the relevant authority or examiner
2. Meter register serves as evidence of quantity supplied (where supply is done through meter)
3. ±2 % meter deviation from actual quantity supplied.

Special agreements between suppliers and consumers are permitted by the Gas Act. Also in the agreement, quantity supplied may be determined other than by the meter register by recording agreed parameters often in remote mode such as:

1. The maximum hourly or daily demand rate
2. The total demand each day during a charging period
3. The hourly demand, to permit the use of tariff terms for charging for differently priced blocks of gas through a single meter
4. The period of interruption and its proof.
5. Gas pressure and/or temperature at the meter.

Recommendations based upon metering pressure are available for metering system design:

- BSCP 331 and IGE Comm.750: if metering pressure does not exceed 7.5 kPa (75 mbar)
- IGE GM/1: if inlet pressure is between 7.5 kPa (75 mbar) and 7 MPa (70 bar)
- IGE/TD/9: if inlet pressure at installation exceeds 700 kPa (7 bar)

The above recommendations cover installation of gas meters in industrial and commercial facilities for all distributed gaseous fuel where the gas temperature at the meter is between −5 and 40 °C. For other gases or outside the previous limits, recommendations must be consulted which are limited to the following type of meter: positive displacement diaphragm, rotary displacement, turbine, orifice.

5.3.15.2 System Specification

The meter's design specifications include the following:

1. Performance targets

- Continuity of measurement
- Security of gas supplied being monitored

- Flow rate accuracy
- Integrated flow accuracy

2. Maximum flow capacity
3. Operating load profile
4. Upstream and downstream control equipment
5. Maximum allowable pressure drop
6. Operating gas pressure and temperature ranges
7. Gas composition limits
8. Read-out requirements, including time data for interruptible loads
9. Available power supplies for their specifications
10. Space available.

5.3.15.3 Measurement and Correction of Gas Flow Parameters

Only the orifice meter from the three types considered requires a secondary instrument for a flow reading. Secondary devices can also be used as correctors to reduce the measured volume to the base volume. For this purpose, the following instrument classes are available:

- Differential-pressure/pressure transmitters
- Temperature transmitters
- Density meters
- Correctors.

5.3.15.4 Differential-Pressure/Pressure Transmitters

These are pressure sensors measuring pressure of gases or liquids. They can also indirectly be used to measure other variables such as fluid/gas flow, speed water level and altitude. Different technologies are available with different design and cost; thus, selection is made according to the intended application and budget.

The accuracy is usually specified as a percentage of output. Careful selection is needed and ranging since uncertainty proportion to output increases as the output decreases. Recommendations are provided by BS6174 where calibration is also cited which is affected by long periods of operations at high differential-pressures. Figure 5.46 shows a typical instrument for a differential-pressure/pressure transmitter.

To minimise uncertainty down to 1 %, the range of an individual PPT should be limited to about 10:1 on DP (3:1 on flow). With a computer controlling switching, two or more DPTs can be used to increase single meter range ability keeping the uncertainty down to 1 % over 4:1 flow ratio. Also the switching point between DP ranges has an effect on error, but if flow rate variation is known, it can be compensated. Also temperature can cause a zero shift and span change of 0.1 % and a 0.2 % in the basic accuracy; therefore, temperature-controlled applications may be required on some applications.



Fig. 5.46 Typical differential-pressure/pressure transmitter

When using the pTZ method of density determination, the selection of the transmitter should reflect the effect on pressure measurement uncertainty on the total flow measurement uncertainty. Narrow-span transmitters with elevated zero values can be sued, or at relatively low pressure, absolute-pressure transmitters may be necessary.

Pressure transmitters with electrical output signals and with an overall measurement uncertainty of 0.25 % are recommended. (BS6447, 0.2A classification). The preferred location is upstream of the meter, and they should be as feasibly close to the tapping and protected against vibrations and environmental effects. They must be calibrated before commissioning and periodically afterwards using a dead-weight tester.

5.3.15.5 Temperature Transmitters

Accurate temperature measurement is essential for:

- Computation of flowing gas quantities when using the pTZ correction method of density determination

- In cases where operational variations in the flowing temperature will have a significant effect on orifice meter bore dimensions.
- Checking gas temperature in density meters.

Electrical resistance thermometry in accordance with BS1041 pt.3 and BS1904 is recommended. A platinum 100 Ω BS1904 class A tolerance is preferred when pTZ method is used. Class B tolerance is adequate for orifice bore and density meter calibration corrections. The calibration of the resistance thermometer should be done according to national standards. Figure 5.47 shows a typical temperature transmitter.

5.3.15.6 Density Meters

The measurement of density/relative density in gas flow measurement is required:

- For the calculation of volume flow at reference conditions from volume measured at line conditions (e.g. at orifice)
- For the calculation of volume flow at reference conditions from mass flow using relative density
- For mass flow measurement using density at line conditions of temperature and pressure
- For quality determination using relative density.



Fig. 5.47 Typical transmitter

The gas density at line conditions can be directly determined by:

- A continuous on-line density meter
- Measurement of relative density near reference conditions
- Application of pTZ method and by calculation from gas composition

For an accurate gas density measurement, the vibrating element, buoyancy or inertia types of density meter can be used. In the UK, the vibrating element is used for fiscal measurement and has an uncertainty as low as 0.2 % if installed properly. A reference chamber of pipeline gas controls the pressure of the supply to the density cell. A good thermal equilibrium exists between the reference cell and the pipeline gas. The electrical output of the density is related to the relative density of the pipeline gas after suitable calibration.

Continuous density meters contribute significantly to the cost of the system. There cost and benefit analyses for such a need should be made. They are preferred at high gas pressures and approaching critical point conditions because pTZ method will have greater uncertainty, but their use must be avoided when gas is dirty or at dew point conditions.

Density meters are calibrated by the manufacturer on a range of gases that have certified component analysis such as nitrogen or argon. Their primary objective is that the gas pressure and temperature at the point of density measurement must be as close as possible to those at the flow meter its self.

The density meter is fitted in the line using a pocket at a specified point according to standards. There must be access to the meter for maintenance and testing without interrupting the gas flow. The density meter bears the advantage that it can be externally placed.

The pTZ method can be used when it is impossible to measure the density directly at line conditions, but the density reference conditions are known. The pTZ method should be considered as an alternative to the direct on-line method where

- uncertainties in compressibility factor are small (low-pressure case)
- gas is near dew point conditions and contains liquid droplets and dirt; thus, a direct sample cannot be obtained
- pressure and temperature and other operating conditions are beyond the density meters' available ranges.
- benefit of accuracy is overturned by cost, and pTZ method gives adequate accuracy at lower cost than a density meter.

Figure 5.47 shows schematically the features of the density meter with a typical illustration (Fig. 5.48).

5.3.15.7 Correctors

Automatic correctors can achieve pressure temperature and compressibility correction. They all require an output drive from the meter either by means of direct

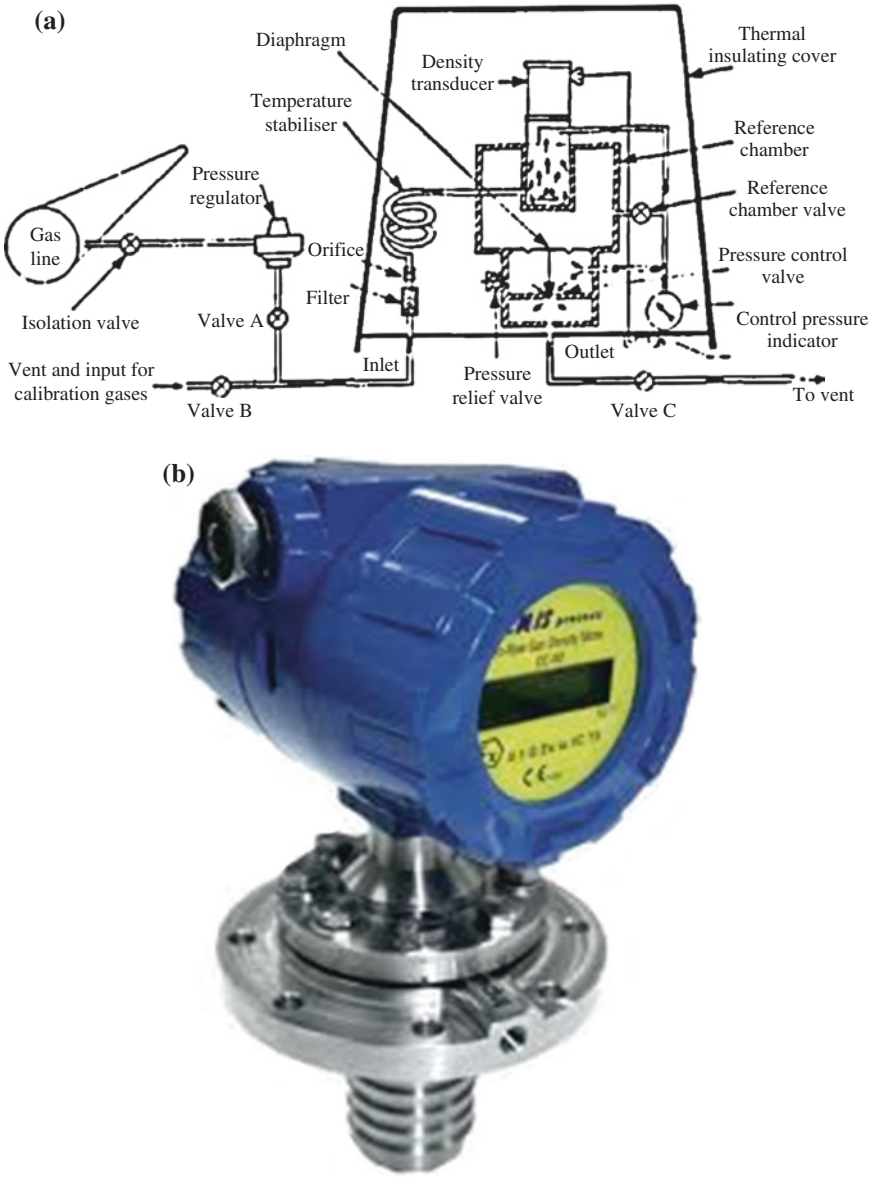


Fig. 5.48 Schematic features (**a**) of density meter with typical illustration (**b**)

mechanical linkage or by indirect electromechanical/electronic means. They are available in mechanical type with accuracy ±1 % and electronic type with accuracy ±0.25 %. They utilise the gas law correction where flow rate Q is a function of the ratio of pressure, temperature and compressibility at reference condition and

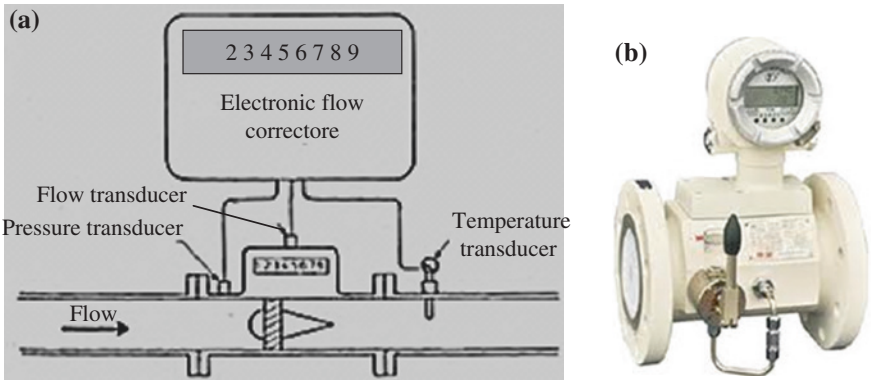


Fig. 5.49 Schematic diagram of corrector (**a**) with typical illustration of the meter (**b**)

at line condition. Figure 5.49 shows a typical corrector. There also two types of automatic pressure/temperature/compressibility correctors:

1. The trapped sample which operates by trapping a specific mass of gas in a flexible container and measures its deflection when subjected to gas-line pressure and temperature giving correction accuracies of ± 1 %.
2. The density cell which utilises a density transducer mounted in a small gas sampling line giving an accuracy of 0.2 %.

5.3.15.8 The Flow Computer

The flow computer encompasses the development of correctors. It is used in high-pressure metering, custody transfers and of course in offshore installations. The signals from the two elements (primary, secondary) are transmitted to the computing system. There the signals are processed into engineering quantities such as flow rate, and they are presented in graphical displays together with a systems diagnostic reports such as warnings or operational data such as valve position.

Possible system configuration for digital computers may include:

1. A number of meter stream computers associated with one meter task connected to a bank which concentrates information from the peripheral meters to produce the station's operational status.
2. A master computer system which receives all transmitted signals from the stations and processes them into engineering data. For fiscal measurement, a backup in case of system failure must be provided, such as a dual master computer configuration.
3. For simple fiscal systems, a single computer can be used in conjunction with an independent device displaying flow rate.

The computer system must meet the specifications and performance for real-time flow rate measurement giving accuracies of the order of 0.01 %. The Petroleum Measurement Manual, Part XV, Sect. 2 covers the specifications concerning computer system components and data integrity and security.

Meter stream computers may perform the following functions:

1. Process digital or analogue signals and convert them into engineering data and specific readings (density signals are transformed into density readings) for computation purposes.
2. Monitor these signals for any system deviations and provide alarms if limits are violated.
3. Communicating with the supervising meter bank computer for several issues including both computer's and system's health.

A meter bank computer may perform the following tasks:

1. Sum flow rates from stream computers and give totals for the connected stations.
2. Handle signals not already processed by stream computers.
3. Transmit alarm signals received from stream computers to the data logging system.
4. Calculate gas properties such as calorific value from data entered manually or acquired from an on-line chromatograph.

5.4 Gas Quality

5.4.1 Introduction

Gas quality is the key parameter which must be constantly monitored and measured. Broadly speaking, gas quality measurement is important for four main reasons:

1. To ensure the safe and efficient transportation of gas through the transmission system as the systems are designed to operate to a particular gas specification.
2. To certify that gas appliances receive the correct gas specifications as they are unable to safely and efficiently consume gases that are not correct.
3. To guarantee that consumers receive the quality of gas that they pay for.
4. To ensure that any gas that is injected into the natural gas grid (e.g. liquefied natural gas, imported gases or bio-methane) matches the quality of gas that already exists in the transportation system.

Natural gas consists of a number of different hydrocarbon gases along with some non-hydrocarbon gases and other contaminants. To put it simply, gas quality specifications define the mixtures of these gases that can be transported and conveyed to consumers. In the UK, the gas quality specifications are laid out in the GS(M)R. It is important that gas producers, suppliers and transporters adhere to

Table 5.21 GS(M)R Gas quality specification for the UK

Content or characteristic	Value
Hydrogen sulphide content	≤5 mg/m^3
Total sulphur content (including H_2S)	≤50 mg/m^3
Hydrogen content	≤0.1 % (molar)
Oxygen content	≤0.2 % (molar)
Impurities	shall not contain solid or liquid material which may interfere with the integrity or operation of pipes or any gas appliance which a consumer could reasonably be expected to operate
Hydrocarbon dew point and water dew point	shall be at such levels that they do not interfere with the integrity or operation of pipes or any gas appliance (within the meaning of regulation 2(1) of the 1994 Regulations) which a consumer could reasonably be expected to operate
Wobbe Index	(i) ≤51.41 MJ/m^3 and
	(ii) ≥47.20 MJ/m^3
ICF	≤0.48
SI	≤0.60

these specifications in order to avoid financial penalties and/or even criminal prosecutions (not conforming to them could jeopardise public safety). The Table 5.21 provides the GS(M)R Gas Quality Specification for the UK.

GS(M)R also goes on to specify that all gas with a pressure below 7 barg should be treated with a suitable stenching agent (odorant) to ensure that it has a distinctive and characteristic odour which should remain distinctive and characteristic when the gas is mixed with gas which has not been so treated (for further details on odorimetry see Sect. 4.5.1).

It is clear from the GS(M)R Specification that gas quality specification relating to the distribution and transmission of natural gas must have items relating to the integrity of both the pipeline and the flow of gas. The composition of the gas entering the network must be such that the pipeline is not damaged or affected in any way. Corrosion stressing and abrasion must be avoided.

The following sections in this chapter will discuss the various parameters that must be measured in order to accurately analyse gas quality.

5.4.2 Methanol Determination

As part of the commissioning of a new transmission pipeline, the line must be hydrostatically tested to prove that the line can withstand the operation pressures it was designed for. Subsequent to this, the residual water must be removed in order to prevent the formation of gas hydrates (white crystalline compound that looks like ice that forms when water mixes with natural gas at high pressure). Water removal is achieved by swabbing the line with pigs (pipeline inspection gauge).

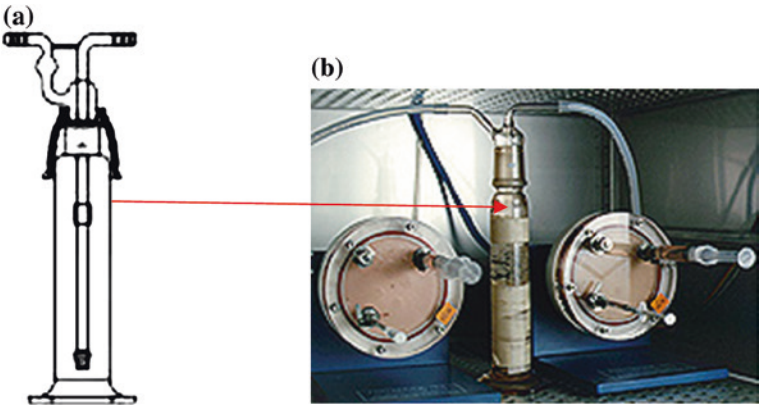


Fig. 5.50 Schematic diagram of methanol determination (**a**) and typical illustration (**b**)

However, this will not remove all the water. The next step is to lower the hydrate formation temperature below ambient temperatures using a hydrate inhibitor. Either methanol or glycol can be used (or both), which will lower the formation temperature significantly. To get the methanol (or glycol) into the pipeline, the pigging procedure is repeated with a large slug of methanol.

The foregoing paragraph has explained why methanol is introduced into the pipeline, and from this, we can accept why instruments for measuring methanol concentration in the gas and the water dew point have an important role to play in the assessment of gas quality. Figure 5.50 shows a typical methanol instrument, and Table 5.21 provides more details.

5.4.3 Water and Hydrocarbon Dew point Measurement

Due to the phenomenon of retrograde condensation, water and hydrocarbon dew point measurement is another important criterion of gas quality measurement. For example at the terminals of the UK gas transmission system, the gas has its hydrocarbon dew point depressed to about −15 °C at 70 bar which ensures that no hydrocarbon condensation can occur. If the dew point was not depressed, then the hydrocarbon dew point would rise to a temperature in the range 5–15 °C at 70 bar with the occurrence of retrograde condensation when the pressure is reduced. Refer to Dew Point section in Table 5.22 for specific instrument details

5.4.4 Oil–Fog Concentration Measurement

Along with the use of dry natural gas in distribution systems comes the problem of the shrinkage of rubber sealing rings in jointed mains. Methods for

Table 5.22 Methonal and dew point determination

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
British Gas Methanol Determination Method	There are two stages involved in this method: 1st stage—0.5 ft³ of gas is passed through 20 mls of water at a rate not greater than 1.5 ft³/h. The apparatus is usually connected to the main under test by means of a high-pressure needle valve fitted into an existing valve, or a pressure gauge can be temporarily removed. One end of a short piece of PVC tubing is attached to the needle valve and is purged with gas; the other end is then connected to the bubbler, the outlet of which is connected to a suitable meter (wet meter) 2nd stage—Injection of 1 microlitre of the wash water into a chromatograph. Calibration is done with 1 microlitre of a standard solution containing 1 ml (0.8 g) of methanol per litre. The column is of stainless steel or aluminium 3 m long and 2.5 mm wide packed with chromasorb 102 or pora pak q, 60/80 or 80/100 mesh. The column temperature is at 150 °C and the injection port is at 150–200 °C. The carrier gas is helium or nitrogen, at 28 ml/min. A flame ionisation detector is used to give the required sensitivity. The concentration of methanol in the gas in gallons/million sft³ is given by: $\left(\frac{5.6 \times h_1 \times c}{h_L \times V}\right)$} where h_1 = peak height from sample, h_2 = peak height from standard C = concentration of the standard (g/lt) V = volume of gas passed, ft³. Range: 0.01–10 gallons of methanol per million scf of gas. Accuracy: the slip is about 1 %	A: Only one component is determined without interference from hydrocarbons, and standardisation is simple	Used in the gas industry to measure methanol in a pipeline after the line has been pigged with a large slug of methanol
UGC Dewscope (Hydrocarbon dew point)	Determined at mains pressure, the hydrocarbon dew point can be measured in the form of a liquid film that appears on the mirror component of the instrument. This is achieved by gently cooling the temperature until dew appears, the dew point is often not visible until the temperature is several degrees below the dew point. The temperature is then increased gradually and the dew point is taken as the temperature at which the deposit just disappears	Portable can be used directly at pressures up to 3,000 psi	Gas industry and laboratory

(continued)

Table 5.22 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Paraffin washer (Water Dew point)	The paraffin washer is used to remove the hydrocarbon as water dew point is much lower than a hydrocarbon dew point. The instrument is connected to the main, and the gas is purged at a rate of 5 ft/h for a minimum of 10 min. The rate is then reduced to 2 ft/h in order to obtain readings. Water dew point is determined at mains pressure. Optics of the dewscope are arranged so that the dew appears on the mirror as a white deposit against a dark background. To carry out at test, the instruments are cooled at a slow pace until a mist appears. The temperature is then allowed to rise again until the dew evaporates and then slowly reduced again. By careful adjustment of the coolant, either the temperatures are noted at which the dew appears and disappears and then the values average taken, or the dew point is the temperature at which a small deposit does not increase or decrease in amount		As above
Oil impingement apparatus	Roughly between 2 and 20 ft^3 of gas is passed for this test, and it is metered at the outlet. The sample is obtained in a glass-weighing bottle and sent to a lab for analysis where the contents of the glass are dissolved in spectrosol cyclohexane. The absorbance of the cyclohexane extract is measured and compared with a calibration curve. Then, requisite dilutions of the oil are made and measurements taken again. Can detect oil quantities of as low as 5–10 mg		As above

combating this problem require introducing an oil vapour into the mains which can be absorbed by the sealing rings, causing them to reswell and prevent any leakage. Such techniques facilitated the need for the development of instruments/methods to determine the oil–fog concentration in the gas phase and the effects of the oil used on the joints. For further details of the instrument, see Fig. 5.51 and Table 5.22.

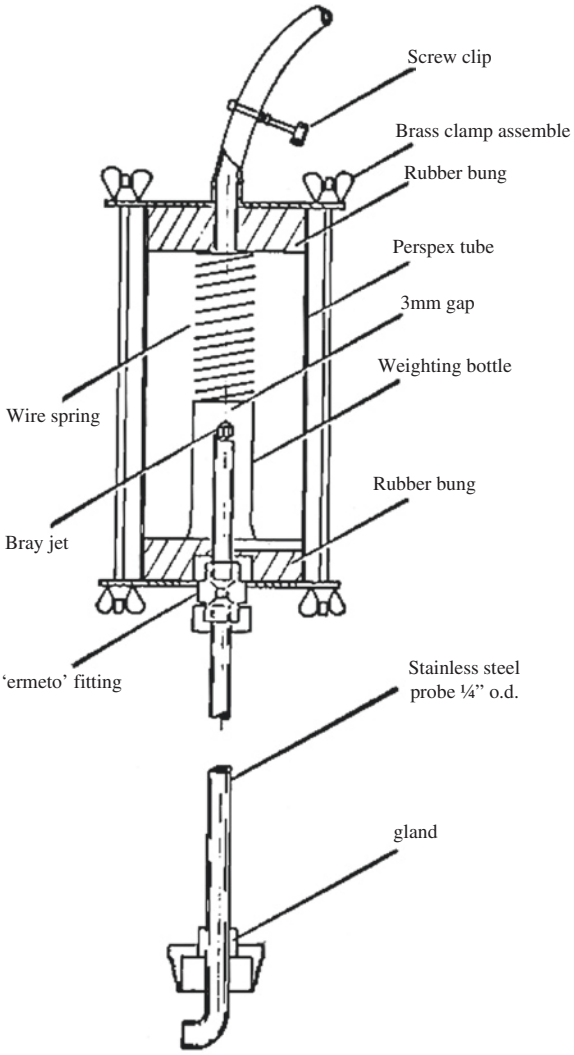


Fig. 5.51 Schematic features of oil fog concentration measurement

5.4.5 Odorimetry and Leak Detection Measurement

This section has been split into two parts: odorimetry and leak detection. Firstly, odorimetry explains why odour is added to the gas and why accurate measurement of odour concentration is essential. Secondly leak detection, leading on from odorimetry, explains why measurements of the leak are carried out.

5.4.5.1 Odorimetry

Odorosity is an extremely important criterion of gas quality. The smell in gas is man-made since raw natural gas is odourless. It is added during the gas-manufacturing/gas-refining process so as to be able to detect it. The chemical is a mercaptan. It is important that the concentration levels of the odorant are correct and consistent across the country. If the odour is too weak, then gas leaks may go undetected, and if too strong, then even the minutest of leaks will be detected miles away. One drop of mercaptan can be detected up to three miles away. For this reason, it is crucial that accurate measurements of odorisation are made. Figure 5.52 shows the schematic features of an odorisation chambers.

Odour can be measured using either subjective or analytical methods. Analytical methods work by tracing the sulphur compounds contained in odorants. They use the instruments and techniques designed for measuring total sulphur and hydrogen sulphide to assess odorant concentration.

Subjective measurement of odour is carried out by trained personnel known as rhino-analysts. Rhino-analysts smell the gas and assess the odour concentration depending on the intensity of the smell and the odour assessment scale which they are using. There are various scales of odour assessment which are used within the global gas industry: the odorisity scale, the odour level and the odour intensity scale.

The instruments used for the analytical and subjective measurement of odour are detailed in the Odorimetry section of Table 5.23. Figure 5.53 also shows the typical British Gas odorimetry.

5.4.5.2 Leak Detection

Subsequent to the detection of a gas leak, measurements of the intensity of the gas leak must be made. Broadly speaking, this is carried out in order to track down the exact location of the leak and to assess the immediate danger presented by the leak. Leak detection instruments are listed and detailed within the Odorimetry section in Table 5.23. Figure 5.53 also shows the Gasco Seeker, the typical flame ionisation and electron capture detection for gas leak detection (Fig. 5.54).

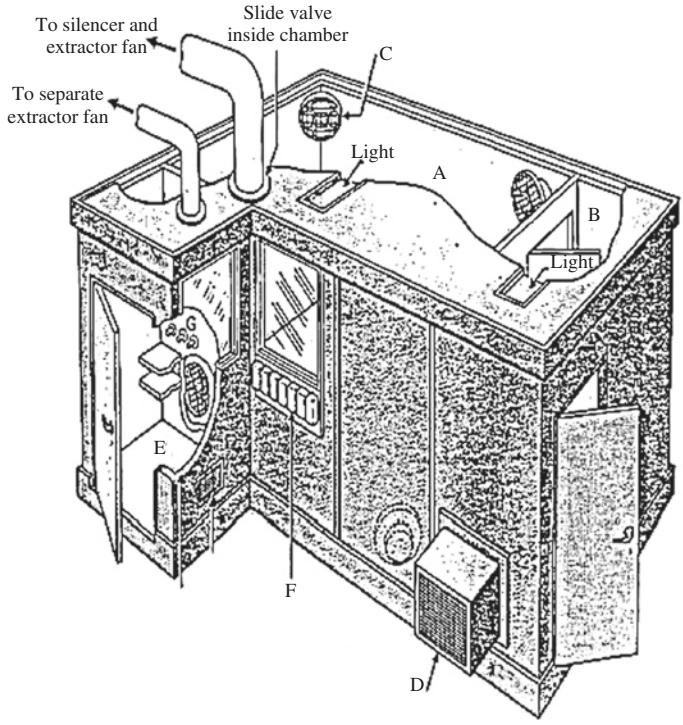


Fig. 5.52 Ordorisation chamber. *A* Test room. *B* Air lock. *C* 4 mixing fans. *D* Carbon air filter. *E* Sample injection chamber. *F* Electric fan switches. *G* Odorant injection point

5.4.6 Sulphur and Hydrogen Sulphide Concentration Measurement

In addition to water and oil removal, one of the most important parts of gas processing involves the removal of sulphur. Natural gas from some wells contains significant amounts of sulphur. Natural gas with a high sulphur content is commonly called 'sour gas' because of the rotten smell caused by the sulphur.

Sulphur exists in natural gas as hydrogen sulphide (H_2S), and the gas is usually considered sour if the hydrogen sulphide content exceeds 5.7 mg of H_2S per cubic metre of natural gas. The process for removing hydrogen sulphide from sour gas is commonly referred to as 'sweetening' the gas. Sour gas is undesirable because the sulphur compounds it contains can be extremely harmful, even lethal, to breathe. Sour gas can also be extremely corrosive. It is therefore essential that it is measured and controlled. Please refer to the relevant section in Table 5.23 for details on sulphur and hydrogen sulphide concentration measurement devices. Figure 5.55

Table 5.23 Various gas quality instruments

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
The Watson House Odorimeter (Odorimetry)	This is a subjective method whereby the measurements are made by rhino-analysts. Gas is introduced to a stream of pure air, which then passes through a funnel into the atmosphere. The observer smells the effluent from the funnel and then records his/her analysis which is recorded in accordance with the odorisity scale. Range: Flow range of 0.002–1.5 ft³/h and pressure range of 0.5–6 in water gauge	Cost, static nature of the system and training required for rhino-analysts	Gas industry and laboratory
British Gas Odorant Chromatograph (Odorometry)	This method is an analytical method. Sulphur contents are separated and passed into a hydrogen/oxygen flame, burning in a hydrogen atmosphere at the tip of a burner in a silica tube, any sulphur compound present forms species which emit a characteristic radiation in the violet and UV regions. Can determine individual sulphur compounds down to concentrations of 0.01 ppm	Extremely accurate costly	Used in offtakes and pressure reduction stations (PRS)
Gasco seeker (leak detection)	Portable device for measuring methane concentrations in gas / air mixture atmospheres. It uses a pellister (VQ3 Pellister) and Thermal conductivity detector as the second filament. A hand-operated aspirator is used to draw the test atmosphere. Range: +1 % LEL scale +1 % Gas scale +3 % reading	Gas detector is a reliable means of detecting flammable gas, and of measuring the percentage of lower explosive limit and the percentage by volume Must be regularly calibrated Does not pinpoint the exact location of the leak	Emergency gas engineers use this device on a daily basis to detect gas leaks
Flame ionisation monitor (leak detection)	Effluent gas is passed into hydrogen/air flame in which two electrodes are placed with a potential difference between them. The detector current passes through the high resistance creating a potential difference which appears across the output of the impedance converter and after suitable attenuation is measured by the recorder. Max leak sensitivity of 6	Extremely sensitive. Must handle with care	Used to detect high-risk main leakage in winter surveys

(continued)

Table 5.23 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Electron capture detector (leak detection)	Consists of an irradiation chamber and a pair of electrode. Eluted molecules with a strong electron affinity will capture electrons to form heavy, slow moving negative ions. The resulting decrease in the current flowing between the electrodes is measured. Range in 'sniffing' mode is 1 × 10*-8 mls. when sampling 1 × 10*-9 mls	High sensitivity	Used for leak testing new mains on air and pressure testing of vessels
The British Gas Sulphur Selective Detector	(Same as BG Gas chromatograph) sulphur contents are separated and passed into a hydrogen/oxygen flame, burning in a hydrogen atmosphere at the tip of a burner in a silica tube, any sulphur compound present forms species which emit a characteristic radiation in the violet and UV regions	A: good and accurate D: Sensitive	Gas industry and laboratory
The Barton sulphur titrator	Consists of three basic units, a titration cell, an electronic control circuit and a recorder. Bromine is generated by passing an electrical current through a suitable electrolyte solution. Compounds bubbled through the solution react with the reagent. The net generating current is thus directly proportional to the concentration of the compounds titrated and is continuously recorded. Range: 0–1,000 in 7 steps, Hydrogen sulphide (0.08 ppm H_2S by vol), organic sulphides (0.16 ppm) and sulphur dioxide (0.16 ppm)	Sensitive	Gas industry and laboratory

also typifies a gas pipe exposed to excessive sulphide gas concentration which led to drastic deterioration of the corresponding pipe over a period of time.

5.4.7 Component Analysis (Chromatography)

The field of component analysis is also very significant as the analyses have a bearing on the pricing structure for the natural gas. Component analysis or gas chromatography as it is often referred to is a physical method of separating the

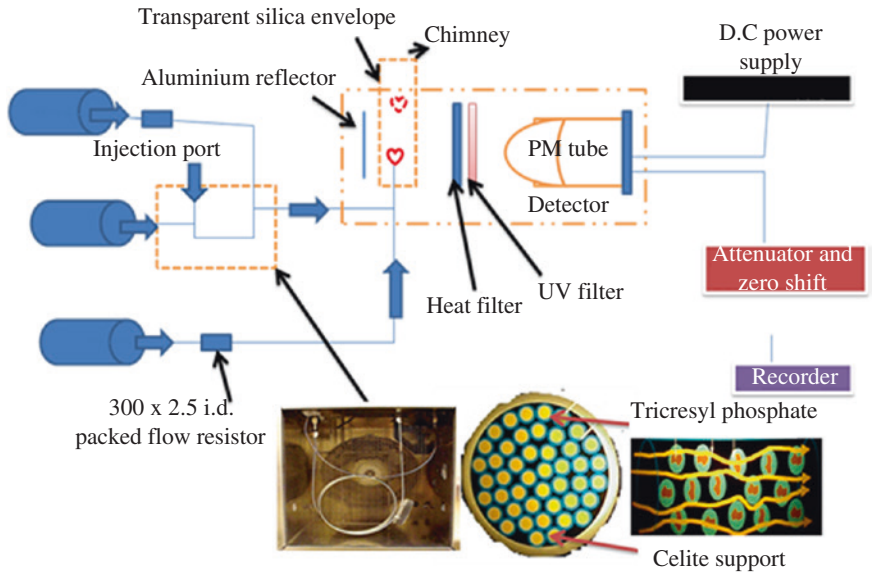


Fig. 5.53 British gas odorant chromatograph

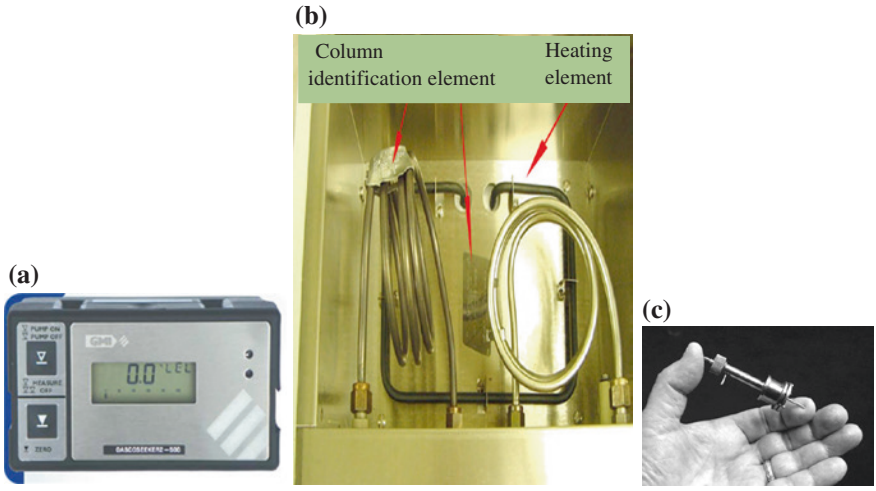


Fig. 5.54 Typical Gasco Seeker (**a**) flame ionisation detector (**b**) and electron capture detector (**c**) [13]

various components that constitute the natural gas mix (see also chromatography section in Table 5.24). Figure 5.56 shows a combination of three illustrations, two of which show the gas chromatograph recording equipment (in this case a Daniels 500) and the third illustration the sampling pointing on the pipeline:



Fig. 5.55 Typical gas pipe subjected to sulphide gas concentration

Table 5.24 Various gas quality instruments

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
Gas chromatograph (e.g. Daniels 500)	Physical method of separation. Materials are separated and partitioned into two phases. The first is stationary and the second is mobile and percolates through the stationary phase. The mobile phase is gaseous and the stationary phase solid or a solid supporting material impregnated with liquid. Depending on the affinity, the substances are retained to a greater or lesser extent by the stationary phase and issue as separated fractions at the end of the column. The gases in the flowing carrier gas stream are detected by measuring a chemical or physical property	A: Superior to any other procedures formerly employed in terms of time-consumed, separation of multi-component systems and amount of sample required D: Costly	Commonly found on offtakes or any network injection points
Perkin Elemer 8700	Same as above	Gives a full report via a computer	
The Cambridge Thomas Gas Calorimeter	This calorimeter is a modified form of the classic Rossini-style instrument in which heat liberated by the complete combustion of a discrete sample of gas is measured thermometrically in a constant pressure enclosed vessel. Accuracy: ± 0.03 MJ/m^3		Gas industry and laboratory
The British Gas Reference Calorimeter	Range: CVs in the range 30–45 MJ/m^3. High accuracy	Gives direct determination of gross CV, density and relative density. The length of time required to obtain a result can be from 2 days to a week	As above

(continued)

Table 5.24 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application
The honeywell heating value transmitter (HVT)	This instrument uses the relationship between the air to fuel ratio and the CV at stoichiometric conditions. The rotary valve motor speed is required to produce stoichiometric combustion and is then used as a basis for computing the CV of the gas being burned	HVT is insensitive to ambient temperature changes	As above
'Simmance' specific gravity bell (specific gravity)	The 'Simmance' Specific Gravity Bell consists of a small water tank in which a bell is suspended from a beam carrying a pointer. The pointer moves over a scale plate which has two lines engraved on it. These lines relate to the upper and lower heights of the bell between which the effusion time is measured. A balanced weight is fitted at the end of the beam opposite the pointer; this weight controls the rate of effusion. An orifice (0.75–0.8 mm in diameter) is mounted on the nipple at the top of the bell. A dust cap covers the orifice to protect from any dirt. The bell is connected via a cord to the beam, which rests on a quadrant arm to ensure the bell remains in the centre of the tank. Gas or air is introduced into the bell through either of two cocks which are attached to a connecting tube to a standpipe located inside the bell	Accurate and used for many substances but not in use constantly	Used in the gas industry to calculate relative density

5.4.8 Calorific Value Measurement

Calorific value (CV) is a measure of heating power and is dependent upon the composition of the gas. The CV refers to the amount of energy released when a known volume of gas is completely combusted under specified conditions. The CV of gas, which is dry and measured at standard conditions of temperature and pressure, is usually quoted in mega joules per cubic metre (MJ/m^3). Figure 5.57 shows a typical Boys non-recording calorimeter.

Gas passing through the UK National Grid pipeline system has a CV of 37.5–43.0 MJ/m^3, with the exception of Stornoway which receives liquid petroleum gas (LPG). CV information is provided daily to gas shippers and suppliers, which is then used to bill gas consumers. These data are also used by gas transportation companies

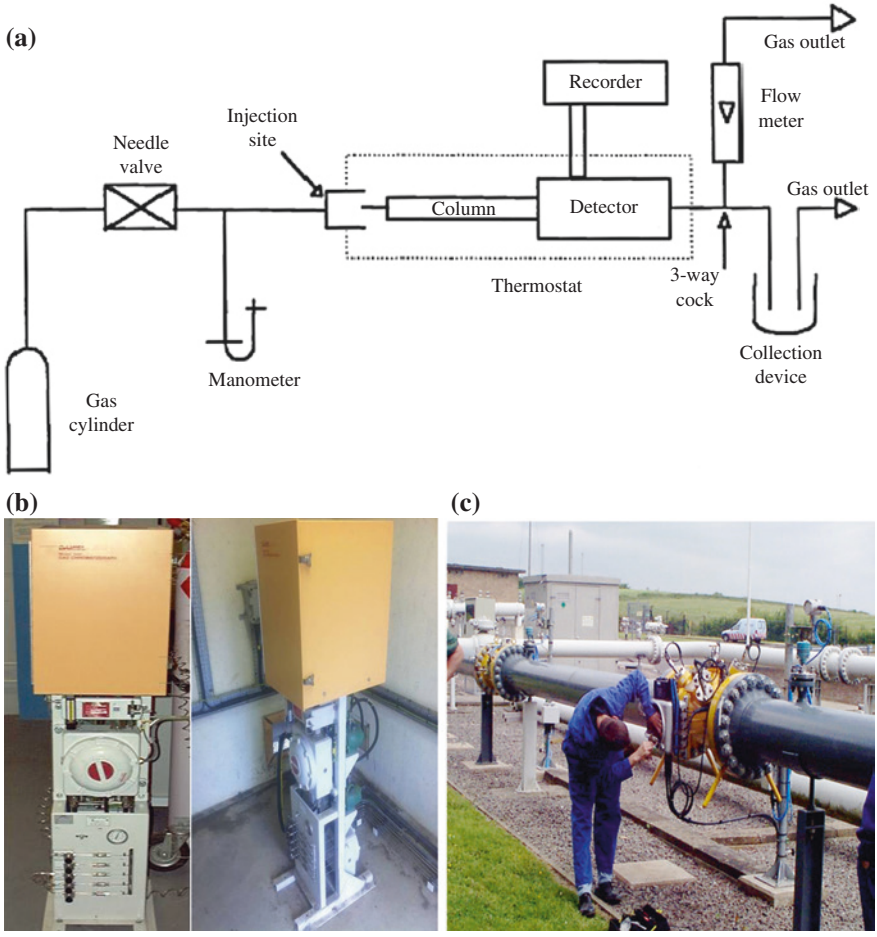


Fig. 5.56 Schematic features of chromatography (**a**) with illustration of Daniels 500 instrument (**b**) and the sampling pointing on the pipeline (**c**)

to determine transportation charges for gas shippers and suppliers. The CV of natural gas is measured continually using process gas chromatographs. Process gas chromatographs separate natural gas into its constituent compounds (i.e. methane, ethane, carbon dioxide) and measure the amount of each in the gas. The physical characteristics of each component, as defined by ISO 6976 in the UK, are programmed into the chromatograph and an overall CV is derived from the measured composition. The determination of the CV of gas is carried out in accordance with international standards and the Gas (Calculation of Thermal Energy) Regulations, as amended in 1997. These regulations stipulate when and where the CV of gas is measured and the type of instrument to be used. In the UK, the regulations are enforced by Ofgem, who also perform audit checks on the primary data. CV measurement instruments are detailed in the CV section of Table 5.24.



Fig. 5.57 Boys non-recording calorimeter

5.4.9 Density Measurement

From the gas flow equations, it is clear that if density is measured directly, then the compressibility of gas is automatically accounted for. This helps explain the reason why British Gas adopted density correction over pressure and temperature correction. To put it simply, measuring density helps measure gas flow more precisely (for further details on density measurement instruments see also the density section of Tables 5.24 and 5.25. Figure 5.58 shows a typical 'Simmance' Specific Gravity Bell used for a simple laboratory spot test.

5.4.10 Wobbe Number Measurement

The maintenance of the combustion performance of gases with regard to heating output and air requirement of burners is influenced essentially by Wobbe number. Wobbe number is simply an indicator of the quality of a fuel gas, measured from the heat produced by burning through a defined orifice under standard temperature and pressure conditions, quantitatively defined as the ratio of Btu per cubic foot to the square root of the specific gravity of the gas. Most gas bills involve a heat-value factor to correct for variations in quality; measured centrally to represent average quality fed into the distribution network, this is applied to the measured volume consumed by each customer to establish the energy charge. The factor could be the Wobbe index, but may be in common energy units or the ratio of current heat energy content to the reference value used in setting the tariff.

Table 5.25 Various industry meters

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application	Range Column
Spinner type	In this instrument, an open-ended tube full of gas is rotated and a measurement is made of the differential-pressure between the centre and the ends of the tube which is directly proportional to the gas density. Accuracy: 0.2–0.5 %	Accurate		Gas industry and research laboratory
Vibrating element	Gas is located in a cylinder inside the device, and the vibrating element is driven at a natural frequency. The resulting displacement is detected and converted to an electrical signal which is recorded. Errors are less than 0.1 %, with accuracies quoted between 0.02 and 0.1 %	A: With non-corrosive, filtered gases, the stability of the cylinder is excellent D: The necessity of having to measure the differential-pressure and maintaining the frequency stability of the electrical supply to drive the motor are major limitations		As above
Buoyant beam	This is the most fundamental of all methods. A sealed globe of glass or stainless steel is immersed in the gas and balanced by a force-balanced system. The electrical output from this system forms the transmitted reading			As above
British Gas Density Balance	The principle is to measure the apparent mass of 10-cm-diameter constant-volume glass globe suspended in an atmospheric sample of gas at an accurately known pressure. By making similar measurements using pure gas (methane) at several known densities at around atmospheric pressure, the instrument may be calibrated and the unknown density inferred. Can handle objects up to 100 g high accuracy, high precision	A: The large thermal mass of the aluminium cylinders help to damp out detrimental effects of small temperature fluctuations D: Extremely sensitive to ambient temperature changes		As above

(continued)

Table 5.25 (continued)

Instrument	Description, range and accuracy	Advantages (A) and disadvantages (D)	Application	Range Column
Ac-Me Balance (relative density)	This is a pressure balance type of instrument. The relative density is calculated from the ratio of the absolute pressures. This is obtained from a beam carrying a bulb and a counter-weight which is brought to balance successively in air and in gas by adjusting the balance case pressure. Accuracy: ±3 %	Reliable		As above
Debro recorder (relative density)	Operates on the principle that output pressure of a fan is proportional to the density of the gas being pumped. It comprises of two bells suspended from a balance beam sealed in an oil bath. Dry air and gas are introduced to the bells, and the difference in pressure between air and gas is recorded by taking the deflection of the balance beam			As above
The aeration test burner (Sigma ATB model)	This instrument gives an immediate indication of the combustion characteristics of the gas under test. It consists of a Bunsen-type burner with a fixed gas nipple operating at a standard pressure. The primary air shutter is then adjusted to manually to provide an inner flame cone to a specific height and the degree of opening the shutter is indicated on a scale, calibrated in units, which can be read of the calibration chart as ATB numbers. 0–400 ATB number	Gives an immediate indication of the combustion characteristics of the gas under test. Readings are taken by individuals; therefore, there is possibility of human error	The burner can be used in gas-mixing quality control and in appliance approval tests	

The above statements help to explain the role of Wobbe number in natural gas. From this, it is clear to see that in the use of gas and gas–air mixtures, it is necessary to monitor and control the Wobbe number as by doing so the quality of the gas being supplied will be restricted be to only minor variations.



Fig. 5.58 Typical 'Simmance' Specific Gravity Bell

5.4.11 Aeration Number Measurement

The aeration number simply gives an indication of the combustion characteristics of the gas under test. Figure 5.59 shows a typical aeration number instrument. Measuring the aeration number helps control the quality of gas entering the network, ensuring the gas is adequate for use in the various types of gas appliances. Refer to Aeration section in Table 5.25 for further details.

5.5 Concluding Remarks

Natural gas has established itself as a primary fuel choice in the twenty-first century. This is due to the fact that it is environmentally friendly and cleaner than oil and coal, hence, greater demand from consumers, i.e. industrial and residential. Throughout the years, the gas industry has witnessed numerous different measurement techniques and experimental investigations relating to natural gas engineering, consequently making it extremely difficult to keep up to speed with the evolving technology that is being implemented across the world.

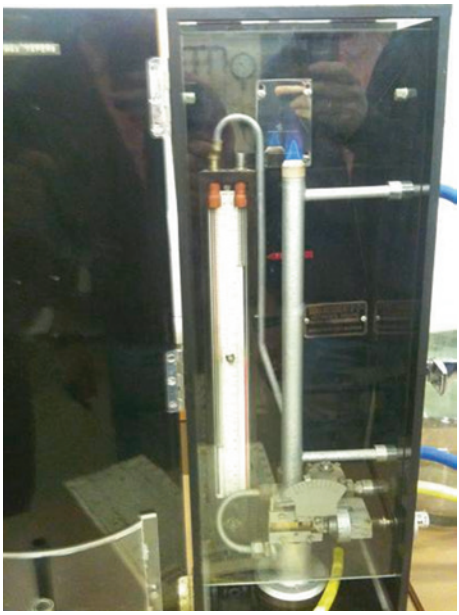


Fig. 5.59 Typical sigma aeration test burner

Contractual obligations require that the distribution and transmission companies supply gas of certain specifications in accordance with regulatory standards and the consumers ensure that the fuel supply from meets the demand requirements. Therefore, there is a need to measure the volume and determine the quality of gas supplied by the operator.

Accurate measurement of gas is influenced by the temperature and pressure of the flow. Actual volume measurement is taken by gas flow instruments, such as orifice plate, venturi meter and pitot tubes. Experimental measurements are carried out to determine the quality of gas by measuring the quantity of impurities such as hydrogen sulphide, carbon dioxide and water present in the gas. It is necessary to remove these impurities as it both poses environmental concern and affects the combustion quality.

In the light of the foregoing, this chapter has presented a comprehensive review of each of the instruments used for measurement of temperature, pressure, gas flow and also the devices used for determination of the quality of natural gas. The chapter evaluated each device in terms of range, accuracy, application and advantages and disadvantages relevant for making engineering decisions. It has also provided a framework for engineers, scientists and managers to familiarise themselves with the instruments and together with the related standards can aid their decision-making in practice.

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Chapter 6
Industrial Utilisation of Natural Gas

6.1 Burners

6.1.1 Introduction

The design and applications of the burners using natural gas have extensively been covered previously by Nasr et al. [1] and Connor et al. [2]. In the following section, however, a brief overview of various burners together with theoretical considerations, where appropriate, will be provided. This can also complement the subsequent sections in this chapter in terms of utilisation of natural gas relating to heat recovery, furnaces and elsewhere in the previous chapters.

The primary role of burners is to provide flame stabilisation, as well as other operations such as air or gas entrainment or mixing in producing the required flow pattern. On the proviso that combustion is complete, the heat released by combustion of unit quantity of gas is the same no matter how the gas is burned. The optimum performance of any industrial, commercial or domestic burner using gas stipulates a range of requirements such as flame temperature, shape, size and aeration. Burners could, arguably, be classified under six broad headings, which are controlled by mixing gas and oxidant [2]. The categories are as follows:

1. Postaerated or diffusion-flame burners.
2. Air-blast burners.
3. Machine-premix burner systems
4. Gas oxygen burners.

Other burner systems: the four mentioned above are used mostly in domestic industrial and commercial gas usage. Other systems include surface-combustion burners, pulsating combustors and catalytic combustors. Also there are special applications of types (i)–(iv), for example, radiant tubes, packaged burners and dual-fuel burners.

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6.1.1.1 Postaerated Burners

In these types of burners, all the air for combustion is obtained from the surrounding atmosphere; thus, there is no primary process. In laminar postaerated flames, with laminar diffusion, the mixing process is essentially molecular diffusion with freedom from lightback and consequently high turndown ratio, and extremely low noise levels. In turbulent postaerated flames, however, diffusion is the governing process. Moreover, compared with aerated burners, laminar postareated burners have their freedom from linting and port blockage, their compactness, the well-defined flame they produce and their relative insensitivity to changes in combustion characteristics. These advantages have enabled the postaerated burners to be the leader in the design of most domestic, commercial and low-temperature industrial appliances, particularly in the times of manufactured town gas in the UK. However, these burners give unacceptable performance on natural gas and on conversion have been replaced by aerated burners. As also noted by Pritchard et al. [2], soft flames of fluctuating shape are formed with much decreased stability due primarily to the lower burning velocity and narrower range of flammability.

There are two versions of radically different design that are capable of burning natural gas satisfactorily. These are (i) *pinhole burners* and (ii) *matrix burners*.

6.1.1.2 Pinhole Burners

Early design of pinhole burners were developed by the VEG Gas Institute in the Netherlands [2]. These consist essentially of a centre-fed pressed-steel flat-topped box with a row of small holes, typically 0.3–0.4 mm diameter which form the main flame ports. The vertical cylindrical main flames are stabilised by adjacent small auxiliary flames as shown in Fig. 6.1.

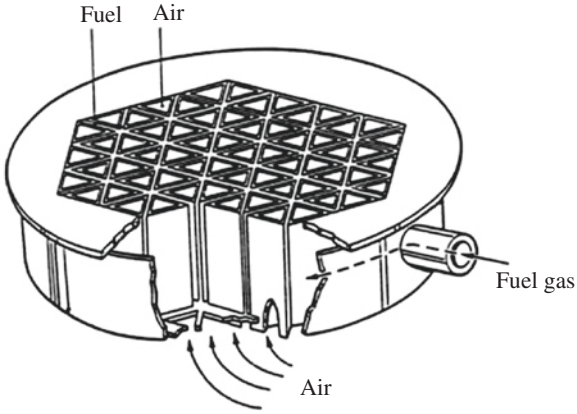


Fig. 6.1 Schematic of postaerated burner [2]

6.1.1.3 Matrix Burners

The principle of matrix burners was originally developed at the British Petroleum Research Centre, and the first version of a matrix burner comprised of a symmetrical array of short hexagonal-ended tubes. Gas fed into the free space between the tubes emerges through the very small interconnected slots formed between hexagonal tube ends on the burner top, the bottom ends being sealed. The exit flow resistance is higher than in the transverse direction so that a uniform gas distribution at the burner face is achieved. Air flow takes place through the tubes under the influence of flue draught or from a low-pressure axial-flow fan similar to that shown previously in Fig. 6.1. These burners burn satisfactorily with any fuel gas ranging from propane to hydrogen with a flame height of only a few millimetres.

Gases with low burning velocity, such as natural gas, have the separation between flame base and burner top which allows some time for gas/air intermixing and the resultant flame is blue without yellow-tipping. The pinhole burner also suffers from burn-back sooting. This problem has prevented the development of both matrix and pinhole types. Furthermore, as with other postaerated burners, matrix burners have turndown of about 40 % due to the low-gas-orifice pressure drop, although placing a restrictor upstream of the burner could significantly reduce this effect.

6.1.1.4 Atmospheric Aerated Burners

The operation of atmospheric burners is based on the primary air which is entrained by momentum sharing between the gas jet and the surrounding air in which the induced air is some 50–70 % of the stoichiometric air requirement. Two types of atmospheric burners may be distinguished: those in which the gas is used at normal supply pressures of 6.25–20 mbar equivalent to 0.064–0.204 mm of water (i.e. Bunsen burner) and those in which the gas is supplied from a compressor or high-pressure supply. Low-pressure atmospheric burners are used on nearly all domestic appliances and have very many low-temperature industrial applications, whereas high-pressure atmospheric burners are restricted more to industrial utilisation. Figure 6.2 shows the essential features of an atmospheric burner.

Air Entrainment

Referring to Fig. 6.2, at (1) gas emerges at high velocity (60) m/s (say) into stationary air. Shear forces between jet and air causes air entrainment into the jet of gas which increases in diameter until it reaches the wall at point (2), with small pressure change between (1) and (2), close to atmospheric pressure. For the mixture to overcome flow resistance in the burner, it is necessary to develop a positive pressure in the burner by slowing down the mixture in a divergent diffuser. Here, part of the kinetic energy of the mixture stream is converted to pressure energy.

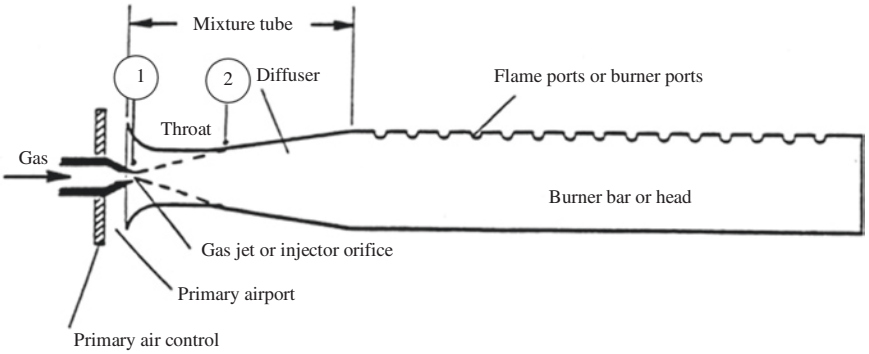


Fig. 6.2 Schematic diagram of atmospheric burner [2]

A homogeneous mixture forms in the diffuser and mixing tube, and this is distributed evenly to the flame ports (or burner ports).

Mathematical analysis is complex since throughout the burner, there are flow resistances that are dependent on flow rate, aeration level and on the local velocity profile. Rigorous treatment is outside the scope of this section, but the analyses of Jones [3] derived from Prigg [4] resolve in their simplest form to:-

$$r = \sqrt{\mathrm{RD}}\left[\sqrt{K\frac{A_P}{A_j}} - 1\right] \tag{6.1}$$

where r is the air/gas ratio, a function of the relative density of gas, d, the flame-port area A_p and the injector orifice area A_j. Constant K accounts for the friction losses and discharge coefficients in the burner, K, are usually 0.6–0.7. For maximum entrainment, the throat area, A_t, should be about 0.7 A_p. The diffuser should have an included angle of 5°–8° and should be about 12 × throat diameters in length. Often, practical constraints result in shorter mixing arrangements with some loss of primary aeration.

The Gas Jet or Injector Orifice

The integrated form of the Bernoulli equation governs the discharge of gas through the injector orifice. This may be written:

$$V_g = A_j C_d \sqrt{\frac{P}{\mathrm{RD}}} \tag{6.2}$$

where V_g is gas rate (m^3/h), A_j is orifice area (mm^2), C_d is discharge coefficient, p is static pressure at orifice (mbar) and RD is relative density of gas (air = 1).

Multi-hole injectors are widely used in the UK to reduce injector "hiss". Primary air ports are large to minimise resistance. May have form of aeration

control but increasingly in the UK they are non-adjustable to give "fixed aeration". Any dimensional changes likely to affect aeration in particular injector/throat misalignment, e.g. for cooker burners on natural gas about 4 % reduction in aeration % for each 1 mm off centre and about 2 % for each 1 % of yaw. Injector jet diameter tolerances are important, e.g. 6 % change for each 0.1 mm departure from design value. Burner pressure for different gases are normally town gas 6.25 mbar, natural gas 20 mbar, propane 35 mbar and butane 27.5 mbar. For conversion from one gas to another, gas pressure is decided on the basis of the gas modulus, as defined in the following section.

Gas Modulus

For the same heat input and the same primary aeration, virtually the same volume of primary air needs to be entrained whatever the fuel gas. Volume of gas which does the entrainment, however, is proportional to $1/C_g$, where C_g is the gross calorific value of the gas. Thus, on conversion from town gas to natural gas, about half of the natural gas needs to entrain the same amount of air. So this smaller flow must be given the same momentum by using a higher pressure. The gas modulus ($\sqrt{}$P/Wobbe No.) thus ensures a constant equal momentum. So to convert from town gas of Wobbe No. 27.8 MJ/m^3 and gas pressure 6.25 mbar to natural gas at 50.7 MJ/m^3 for equal gas modulus, the required pressure is 21 mbar. This is the basis on which conversion pressures in the UK were calculated, and service regulators are all set to give this pressure.

Effect of Gas Pressure and Burner-port Loading

For a given gas, the aeration equation shows no dependence on gas rate and hence *gas pressure*. Thus, a plot of primary air–gas ratio versus gas pressure would be expected to be like curve (a) in Fig. 6.3. In practice, the form of the graph is as shown in (b), the decrease in aeration at low flow rates being caused by changes in friction factor and the discharge coefficient of the flame ports at low Reynolds number.

The degree of primary aeration is determined by the total area of the flame ports for a particular duty, air/gas ratio increases with values *Ap/Ao*. The burner-port area for a particular gas rate is conveniently expressed as the burner-port loading or flame-port loading, which is the heat input to the burner divided by the total flame-port area ($V \times Cg/Ap$ (MW m^2)).

$$\text{BPL} = \frac{\text{Heat input}}{\text{Total flame port area}} = \frac{VC_G}{A_P} \tag{6.3}$$

where BPL is in MW/m^2, V is the theoretical gas rate *(*m^3/h*)* and Cg is the gross calorific value (CV) of the gas (J/m^3).

Also, altering the burner-port loading will also have an effect on flame stability. As also noted by Pritchard et al. [2], reducing the burner-port loading

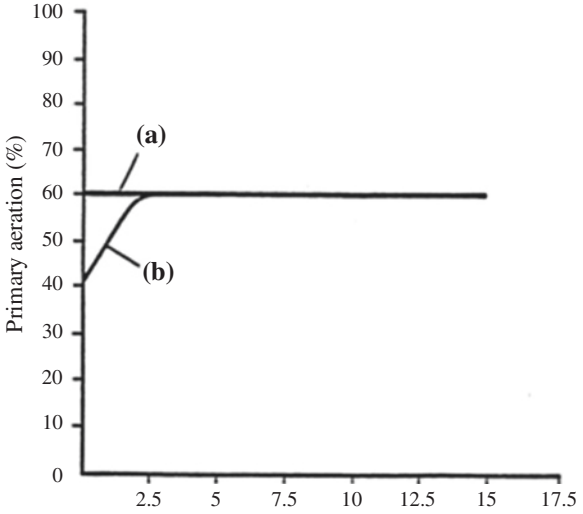


Fig. 6.3 Variation of air/gas ratio with pressure for typical aerated burner [2]

(i.e. reducing the mixture velocity at the flame ports) will, in principle, increase the resistance of a burner to blowoff and decrease its resistance to lightback particularly if the reduction in burner-port loading is achieved by providing larger burner ports. Atmospheric burners designed for town gas normally operated at 40–60 % of primary aeration (r = 2–2.5), and this is achieved with burner-port loading of 18–32 MW/m^2 with Ap/Ao = 25–30. Thus, if the flames set to lightback with high aerations and large flame ports, say, on turndown, then the required flame stability is achievable. Natural gas blowoff occurs at these high BPLs which need to be reduced to 9–13.5 MW/rn^2. Burners with low BPLs widely used in USA, but in the UK, high BPLs are retained and flame prevented from blowoff by an added flame stabilisation device. This can be done by flame-port design and flame retention. For flame-port design, either low efflux velocities resulting from large flame-port areas are used or some external means of flame retention must be provided. Figure 6.4 shows various port configurations. Gauzes are also fitted behind flame ports to provide flow smoothing and to flameproof burners, i.e. prevent them from lighting back. The problem is build up of fluff or "lint" which reduces effective, A, and therefore aeration. Most burners now produced are also multi-gas or "universal" burners. They can burn Family 1, 2 or 3 gases if the injector orifice is changed and the operating pressure is in accordance with gas modulus [1, 2].

6.1.1.5 Air-Blast Burners

There are basically three types of air-blast burners (i) premix (ii) Nozzle-mixing and (iii) tunnel-mixing burners.

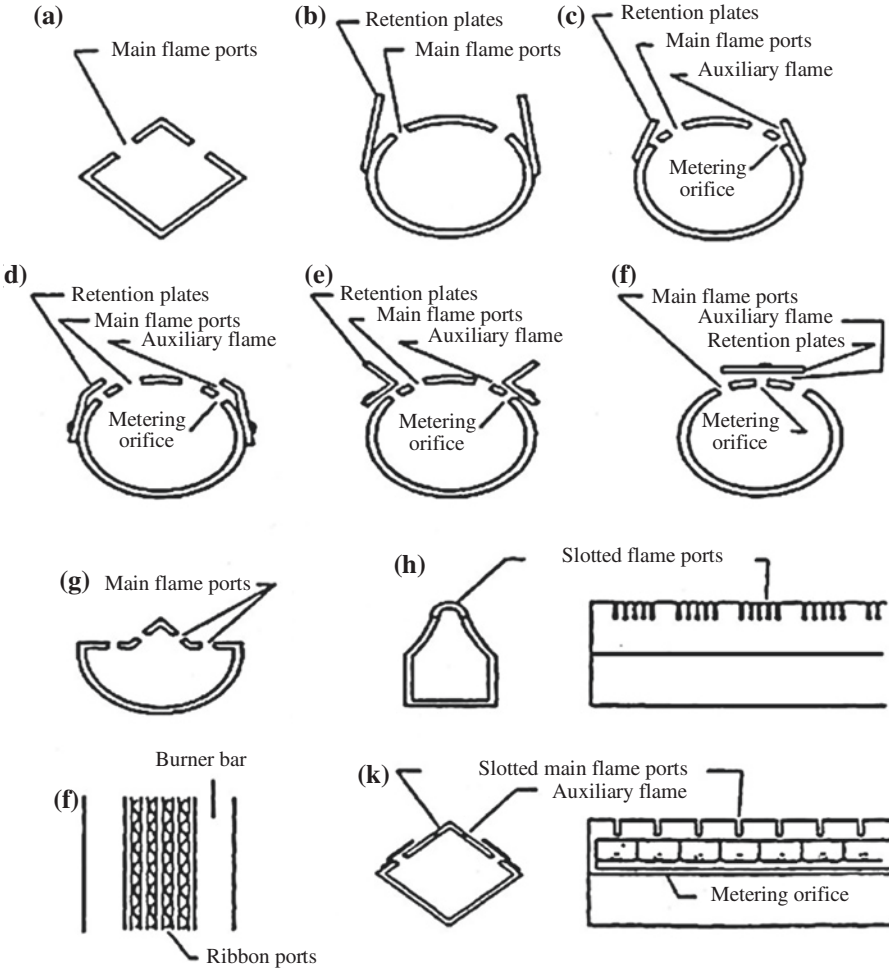


Fig. 6.4 Burner-port configurations [11]

1. Premix types

Figure 6.5 shows schematically a *forced-draught premix burner,* which should follow a set of criteria that have been adopted widely in industrial practice. These criteria are as follows:

- Attainment of Stoichiometric proportions.
- Precise control of air–gas ratio.
- Comprised of an inherent self-proportioning facility, if the gas is entrained from a supply at atmospheric (or combustion chamber) pressure.
- Use a single control valve on the air supply for varying throughput.
- Achieve high mixture pressures using air-supply pressures resulting in high turndown ratios.

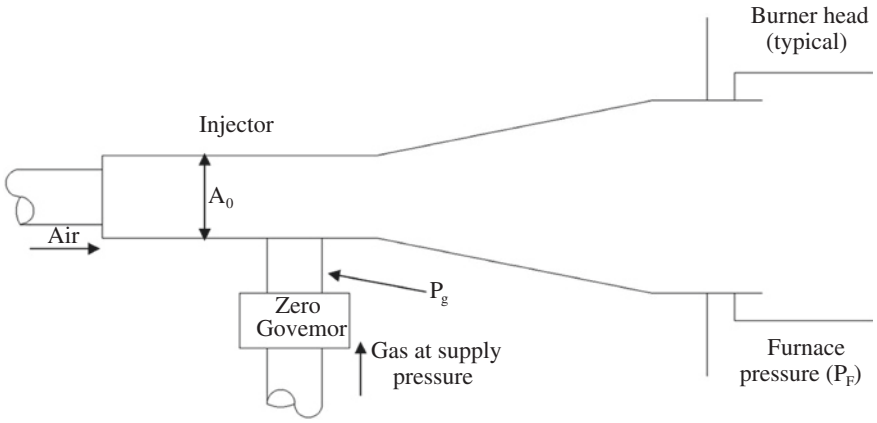


Fig. 6.5 Schematic diagram of a premix burner

- High burner-port loadings may be obtained.
- Positive hot-gas distribution in the combustion chamber should be produced by giving the combustion products sufficient momentum. Thus less reliance on natural draught than low-pressure systems.
- Produce good air–gas mixing which minimises flame length and combustion-space requirements.
- Interchangeability of fuel gases should be achievable.

If $P_g = P_f$ (see Fig. 6.5), then radius, r, is independent of theoretical gas flow rate V_a ($= r \times V_{\text{gas}}$). There is an inherent self-proportioning action, a constant air–gas ratio is given for all throughputs and throughput is controlled very simply by a single control valve on the air supply. Furnace pressure *Pf is* commonly atmospheric, and if air–gas ratio control is required, P_g *is* reduced to atmospheric via a zero-pressure governor.

2. Nozzle-mixing and Tunnel-mixing Burners

With these types of burners, air and gas are simultaneously mixed turbulently during the combustion process at the burner nozzle (burner head). If a tunnel is used, mixing and combustion occur simultaneously in the tunnel and thus the name tunnel-mixing burner. Figure 6.6 shows a typical illustration of this type of burner. The advantages these burners possessed over premix air-blast systems are as follows:

- Lightback is eliminated.
- Utilisation of lower air pressure.
- Preheated air may be employed.
- With lift, flame stability is improved and burners are capable of operating at high excess air levels.
- Capable of conversion from burning town gas to natural gas by changing the air/gas ratio.

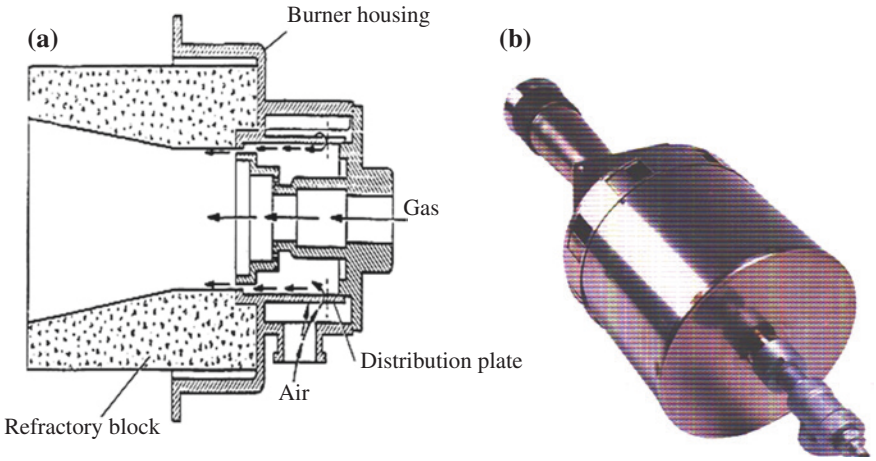


Fig. 6.6 Schematic diagram of nozzle-mixing burner (**a**) [1] and a typical illustration (**b**) [2]

The popular variant integrates the burner with throughput, ratio, ignition and flame-detection controls into a packaged system which are termed packaged burners; these will be described in Sect. 6.1.1.8.

Nozzle-mixing burners can operate on preheated air. These burners provide good combustion, if the mixing is rapid, that is by means of a high relative velocity between the air and gas streams. Alternatively, provision of a large surface area for mixing can also provide satisfactory combustion for the burners. This can be carried out by discharging the air through an annulus or a series of jets. One major drawback with these types of burners is that they are not inherently self-proportioning and control schemes of varying complexity are necessary to provide control of air/gas ratio on turndown. Such control is important since deviation from stoichiometric proportions leads to serious fuel wastage.

6.1.1.6 Machine-Premix Systems

In these, air and gas at atmospheric pressure are drawn in through a proportioning valve to the inlet of a positive displacement compressor or less commonly a fan. The mixture is then compressed and distributed to the burners. Their widest applications are found in supplying precisely controlled mixtures to controlled atmosphere generation plant or to large numbers of small burners in, for example, glass-working machinery. Because stoichiometric mixtures are distributed the safety requirements are severe, and for this reason, the system is not widely used. A recent low-pressure variant is the fully aerated domestic burner, which will be considered in Other Burner Systems in Sect. 6.1.1.8.

6.1.1.7 Gas and Oxygen Burners

If oxygen is used instead of air, high flame temperatures and rates of heat transfer are obtained. The maximum theoretical flame temperature for oxygen/natural gas flames is around 2700 °C or more, compared with about 1950 for natural gas/air flames and heat transfer rates are typically about twice as great. Premixed gas and oxygen are used in small-scale burners below about 30 kW and are used for welding, glass working, etc., but since the burning velocity is very high (four to five times as great as gas/air flames), lightback is the limiting consideration. Larger oxy/gas burners are nozzle-mixing and are increasingly used in high-temperature industrial processes. The cost of oxygen limits the application of gas/oxygen burners, but recent developments in on-site oxygen generation using pressure-swing adsorption or membrane air-separation techniques are likely to widen considerably the applications of gas/oxygen flames.

6.1.1.8 Other Burner Systems

Radiant (Surface-combustion) Burners

In these, a high proportion of the energy supplied in the fuel is converted into radiant energy. Such burners normally consist of a radiant refractory plaque either of a porous material or of a dense material with a large number of small (approximately 1.5 mm) perforations fed either from a premix air-blast mixer or from a low-pressure atmospheric injector. Combustion takes place at or just below the surface and the plaque quickly becomes incandescent. Very low burner-port loadings are possible without lightback. Surface temperatures range from about 800 °C to around 1350 °C. Figure 6.7 shows typical arrangements. Common applications are in space heating (the familiar free-standing LPG heater) and in industrial drying processes. The low flame temperatures in radiant burners result in much lower emissions of oxides of nitrogen (NOx).

Catalytic Radiant Burners

Gases can be burnt without visible flame at the surface of a catalyst at very low temperatures (about 450 °C for natural gas). Space heaters working on

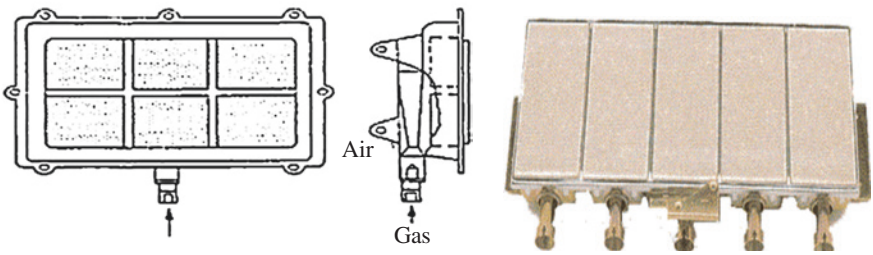


Fig. 6.7 Typical surface-combustion burners

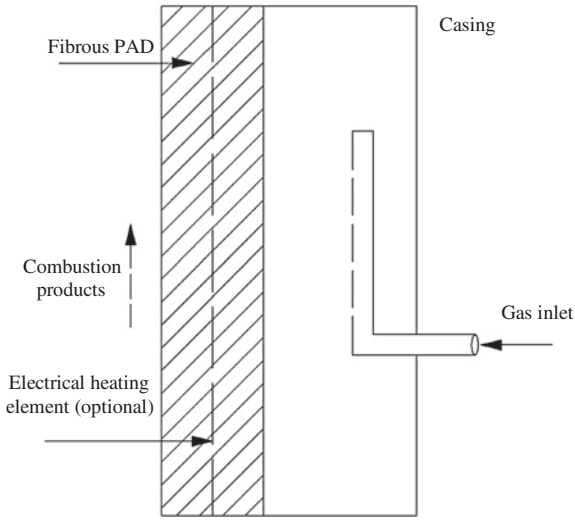


Fig. 6.8 Schematic of typical catalytic radiant burner

this principle using LPG have been commercially available for many years. Combustion of natural gas has, until recently, proved difficult because of the nature of the fuel and problems with the catalyst. Figure 6.8 shows a schematic cross section of a typical diffusive catalytic heater. Combustion air is supplied by diffusion from the surrounding atmosphere into the front layer of the pad. Combustion needs to be initiated by preheating the pad to about 250 °C. Catalysts usually include precious metals such as platinum or rhodium or transition metal oxides, for example, those of cobalt and manganese. Catalytic systems have the advantage that since carbon monoxide is not a combustion intermediate as is the case with flames they can work under conditions of air starvation and vitiation without danger. NOx levels also are very low.

Pulsating Combustor Burners

In these types of burners, combustion and heat release occur sporadically and the pressure waves produced drive the combustion products through the system at high velocity. This also gives rise to high sound levels at the frequency of the combustion oscillations, (typically 50 Hz). It is possible to obtain very high rates of heat transfer in systems powered by these burners, and in recent years, they have come into fairly widespread use, particularly in USA and Japan, for industrial, commercial and domestic applications especially air and water heating. Figure 6.9 shows a simple system consisting basically of a resonance or exhaust tube several metres long and gas and air inlet arrangements. The inlet end forms the combustion chamber into which air is drawn from the atmosphere through a flap valve. The system operates on the basis of flowing gas and air into combustion chamber ignite by spark, combustion occurs, pressure increases and the inlet valve is closed, and burnt gas goes out through the resonance tube. Inertia creates a low pressure in the combustion chamber which opens the inlet valve, and more air and

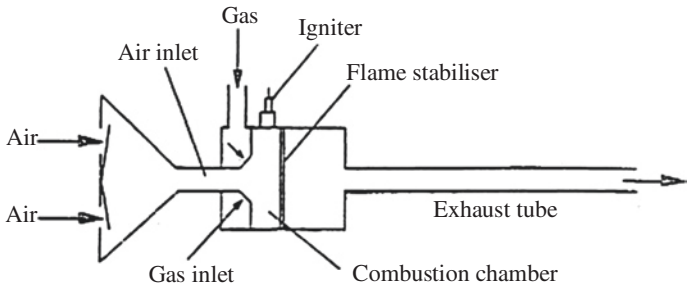


Fig. 6.9 Schematic of typical pulsating combustor burner [2]

gas enter and ignition occurs from the previous cycle and the process continues. The ignition spark is not required after this stage. Noise is the fundamental disadvantage of pulse combustors, but modern silencer systems have reduced noise to acceptable levels for all but the most demanding requirements. Pulse combustors produce lower NOx levels than conventional burners.

Fully Aerated Domestic Burners

Recent developments in compact high-efficiency boilers in which the combustion products are cooled down below their dew point have focused attention on the need for burners which produce short intense flames and which can be fired downwards or sideways as well as upwards . Fully aerated burners in which air from a low-pressure fan is premixed with gas are coming into use for this type of application. Such burners may be formed of rows of corrugated steel ribbons or may be ceramic plaques operating either in radiant or in lifted mode. Downward firing is popular since this avoids problems of condensate dropping onto the burner. The burners typically operate at 20 % excess air levels, and boilers operating at 90–95 % thermal efficiency are feasible.

Packaged Burners

In the firing of boilers and other low-temperature plants, a wide range of forced-draught automatic packaged burners have been utilised since the 1980s and are still evolving, in terms of design and or efficiency. The burner, fan, throughput and air–gas ratio controls and ignition and safety systems are integrated into a single unit. Ratings range from around 0.5 to 18 MW or more. Packaged burners are almost exclusively of the nozzle-mixing type and use metallic burner heads incorporating a wide variety of stabilising and mixing arrangements as shown typically in Fig. 6.10.

Dual-Fuel Burners

Interruptible gas allows the gas industry to optimise the operation of its transmission and distribution system. Provision for interruptible operation may be made by the installation of separate burners for the two fuels but is much more commonly achieved by the use of dual-fuel burners, i.e. burners designed to burn gas or oil

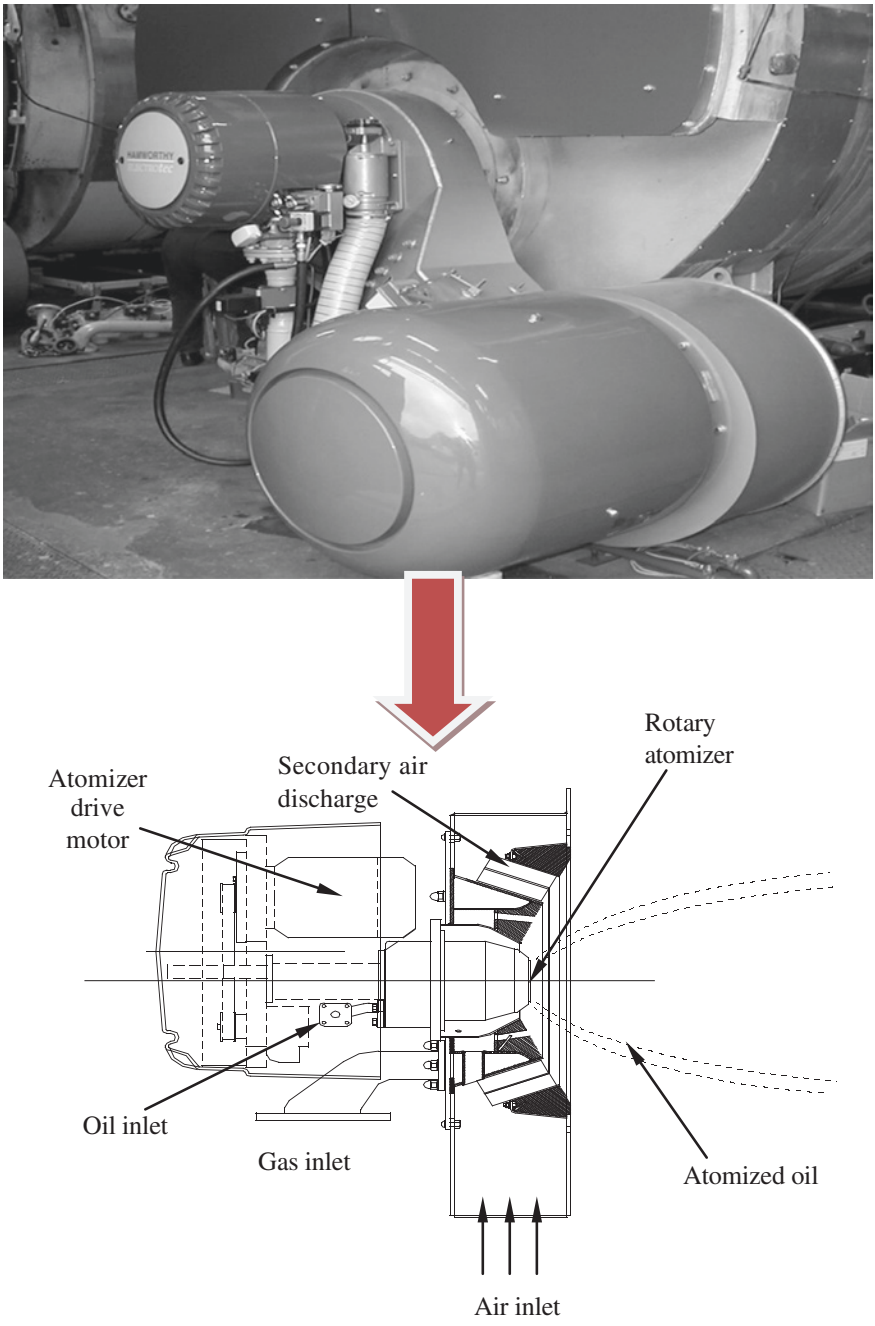


Fig. 6.10 Typical package burner [1]

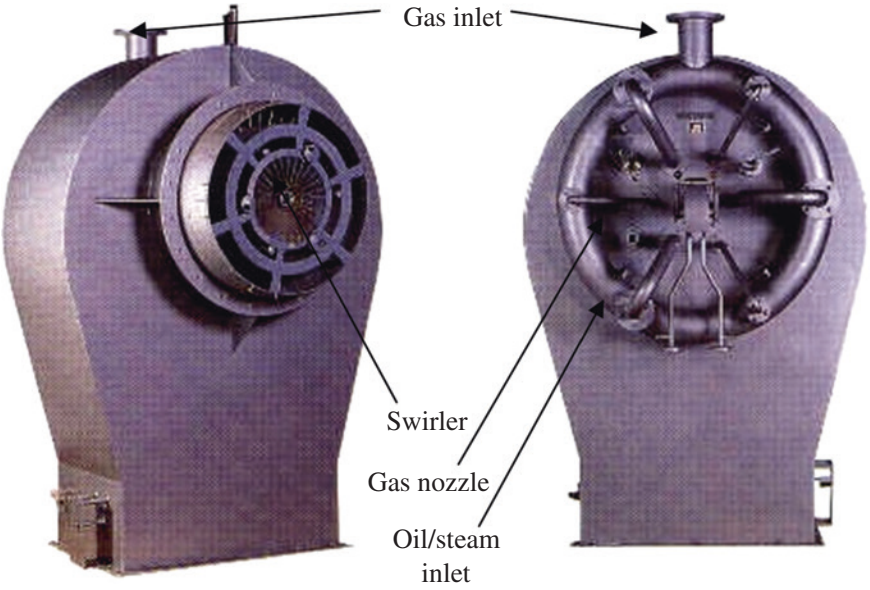


Fig. 6.11 Typical dual-fuel burner [1]

(or less commonly gas or coal) separately, the equipment being capable of rapid and efficient switchover to the alternative fuel. Dual-fuel burners have found their widest application in industrial boiler firing.

The common components for both systems include the air-supply duct which in the case of packaged burners is attached to the burner body, vanes or plates to control the air direction, an air manifold or windbox and a refractory tunnel or quarl. In addition, a pilot burner or burners must be provided and carefully located so as to ignite both oil and gas reliably. The pilot burner will be integrated into the control programme of both fuel systems when automatic control is used. Flame detectors also need to be particularly carefully located so that the pilot, main gas and oil flames are detected. The fact that the two fuel supply systems are incorporated into one burner housing leads to some degree of mutual interference. The gas system is much more tolerant with regard to mixing and combustion than the alternative system and the effect of introducing a small oil atomiser on the centre line of the gas supply system presents few problems. Conversely, the introduction of a relatively much larger gas supply into an oil burner may interfere with the mixing pattern and the solution has generally been to dispose the gas inlets out towards the periphery of the burner. Figure 6.11 shows a typical dual-fuel burner.

Heat Recovery Burners

In high-temperature processes, a very large proportion (often over 50 %) of the energy entering the process is wasted in the sensible heat content of the hot waste gases. Some of this waste heat may be recovered by using it to preheat the load

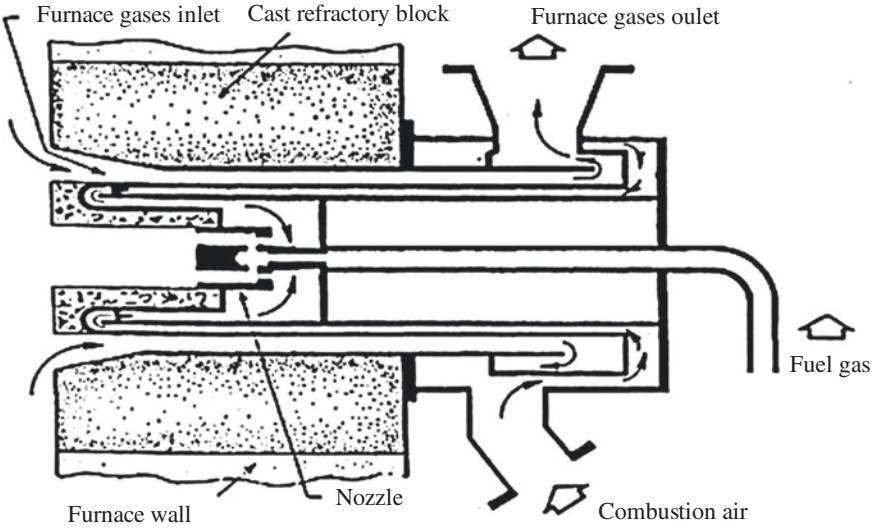


Fig. 6.12 Schematic diagram of recuperative burner [2]

(restricted to continuous processes) or by preheating the combustion air in some form of heat exchanger. Commonly, nowadays this heat exchange is incorporated in the burners themselves to provide high-efficiency compact units. Two types are available: recuperative burners shown in Fig. 6.12 and regenerative burners shown in Fig. 6.13. In the former, the combustion products are drawn counter-current to the incoming air and heat is transferred through a metallic tube. Air preheats of up to 600 °C are feasible.

Regenerative burners on the other hand are cyclical and operate in pairs. Burner "A" fires and its combustion products are drawn through burner "B" heating up the bed of ceramic spheres in its body. The system then reverses, and air is passed through burner "B" where it is preheated. Gas enters and combustion takes place, the waste gases being drawn through burner "A". The upper part of Fig. 6.13 shows the principle of regenerative firing and the lower part the application of regenerative burner to a steel coil annealing furnace. Regenerative burners can be used at very high temperatures (1,600 °C or more) since there is no metallic interface between combustion product and air. Both recuperative and regenerative systems are capable of providing enormous energy savings in high-temperature process.

Low-Nox Burners

This is a relatively new burner classification resulting from recent intense interest, particularly in Japan, USA and Germany in the reduction of emissions of nitrogen oxide from fuel-burning equipment. Nitric oxide (NO) is formed by nitrogen fixation in high-temperature flames. In the atmosphere, it converts to nitrogen dioxide (NO_2) by a photochemical reaction. NO_2 is a major contributor to smog formation.

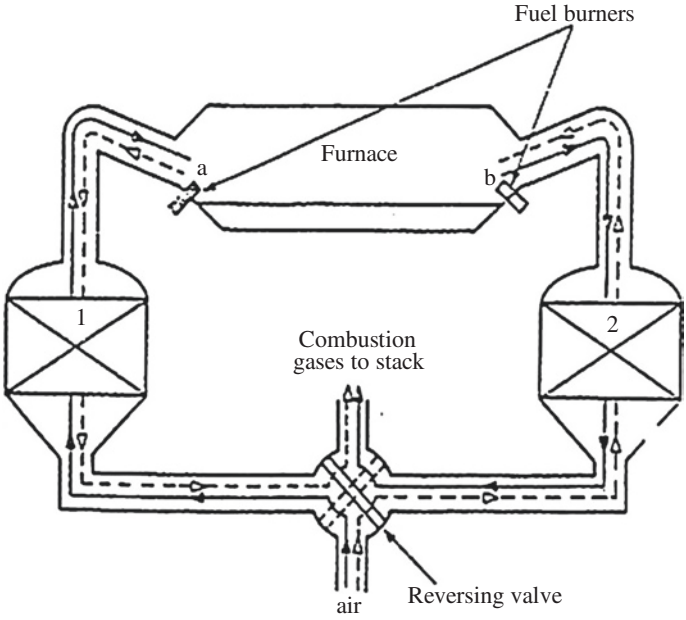


Fig. 6.13 Schematic diagram of regenerative burner

NO and NO_2 are normally lumped together as Knox and considered as a single pollutant. Unfortunately, NOx formation is favoured by those combustion conditions generally regarded as ideal from other viewpoints, i.e. near-stoichiometric, intense, high-temperature combustion. Various techniques are employed in low-NOx burners to reduce emissions including:-

- Recirculation of cooled flue gases into the combustion zone
- Two-stage combustion (fuel-rich primary combustion followed by secondary combustion after heat exchange or with dilution).

Figure 6.14 shows a Japanese low-NOx burner incorporating both these principles.

6.2 Heat Transfer and Heat Exchangers

6.2.1 Introduction

There are many published books and articles dealing with heat transfer and heat exchangers [5–10]. However, throughout downstream gas processing, and as highlighted throughout this book, heat exchangers are regularly utilised and they play an important role. It is beyond the scope of this book to give detailed analysis of these subject areas since they can be treated as stand-alone. To this end, an overview of the modes of heat transfer and the conceptual design of heat exchangers has been provided. The intention here is thus to provide an appreciation of the

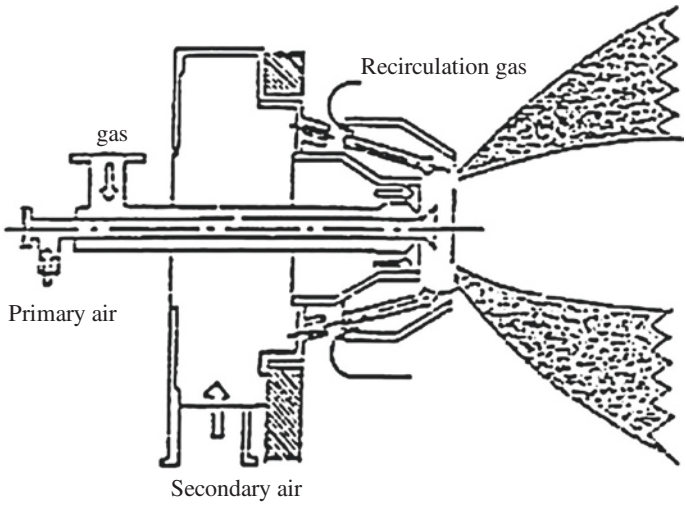


Fig. 6.14 Schematic diagram of low-Nox burner

subject matter, which in turn formulates a better understanding when dealing with heat exchangers in the gas processing and production environment. Readers are also advised to consult various publications, some of which have been given in the reference section of this chapter.

6.2.1.1 Modes of Heat Transfer

There are basically three modes of heat transfer, namely conduction, convection and radiation.

i. **Conduction**: this is primarily concerned with heat transfer in solids and takes place by one molecule transferring heat to adjacent molecules.
ii. **Convection**: this is primarily concerned with liquids and gases and occurs due to the movement of molecules. There are generally two types of convection: the first being natural convection and the other being forced convection.
iii. **Radiation**: radiation heat transfer takes place via wavelength and is transmitted by waves, in a similar way to radio waves using but different wavelengths.

For heat transfer, of any kind to take place there must be a temperature difference.

Conduction

Fourier's law of conduction states that the heat flux through a solid is proportional to the infinitesimal temperature gradient per unit length or more formally for one dimension:

$$q = -k\frac{\theta T}{\theta x} \tag{6.4}$$

where k is a constant of proportionality called the thermal conductivity W/mK, q is the rate of heat transfer $(\mathrm{W}/\mathrm{m}^2)$, θT is temperature difference $\left(\overset{\circ}{K}\right)$ and ϑ_x is the thickness (m).

Heat Transfer Through a Flat Plate

Fourier's law can be exploited to evaluate the heat transfer through a plate. If it is assumed that the heat transfer is steady, the temperature distribution is a function of x only and the conductivity is temperature independent, then by integration Eq. (6.4) reduces to:

$$\dot{Q} = \frac{kA(\Delta T)}{x} \tag{6.5}$$

where $\dot{Q}$ is the heat transfer per unit time (W).

Heat Transfer Through a Composite Flat Plate

It should be noted that the term $\frac{kA}{x}$ in Eq. 6.5 is a constant for a material of known thickness. Furthermore, since the heat flow is steady, the value of Q will not change over any linear distance x. So writing Eq. (6.5) in a more convenient form gives $\Delta T = \frac{x}{kA}\dot{Q}$ or $\Delta T = R_{\mathrm{th}}\dot{Q}$, where R_{th} can be expressed as:

$$R_{\mathrm{th}} = \frac{x}{kA} \tag{6.6}$$

Equation (6.5) is analogous with Ohm's law and can be used to calculate the temperature drops across many plates with different thermal conductivity. Equation (6.5) is termed the thermal resistance for a flat plate.

Consider the heat transfer through a number of flat plates shown schematically in Fig. 6.15. The concept of thermal resistance can be exploited here to calculate the total temperature drop across the plates. That is, since the heat flow is constant across each of the plates, $(\dot{Q} = \dot{Q}_a = \dot{Q}_b = \dot{Q}_c)$. Furthermore, the total temperature change is equal to the sum of the individual temperature changes; thus, $(\Delta T = (\Delta T)_a + (\Delta T)_b + (\Delta T)_c)$.

Applying Eq. (9.5) to each of the plates gives:

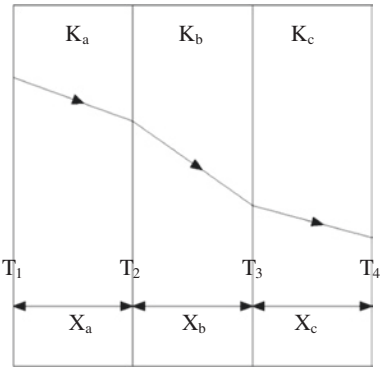


Fig. 6.15 Heat transfer through a number of plates

$$R_{\text{th}_T}\,\dot{Q} = R_{\text{th}_a}\,\dot{Q} + R_{\text{th}_b}\,\dot{Q} + R_{\text{th}_c}\,\dot{Q} \tag{6.7}$$

Hence,

$$R_{\text{th}_T} = \sum_{i}^{n} R_{\text{th}_i} \tag{6.8}$$

where R_{th_T} is termed the total thermal resistance. Application of Eq. (6.6) to the figure will give:

$$R_{\text{th}_T} = \frac{x_a}{k_a A} + \frac{x_b}{k_b A} + \frac{x_c}{k_c A} \tag{6.9}$$

Substitution of this into Eq. (9.9) renders:

$$\dot{Q} = \frac{A\Delta T}{\frac{x_a}{k_a} + \frac{x_b}{k_b} + \frac{x_b}{k_b}} \tag{6.10}$$

Note the main assumption here is that the values of k_i are independent of temperature. If this is not the case, then methods of integration should be employed. Note also that so far we have only considered the heat transfer through solid wall surfaces. It is possible to exploit the thermal resistance concept to consider heat transfer in a fluid system. This is achieved by considering heat transfer from the solid wall to a thin boundary layer of the fluid through which there is a temperature change. These regions are often referred to as films of fluid and we must consider the film heat transfer coefficients.

Heat Transfer Through a Composite Flat Plate (Including Film Coefficients)
The diagram shows a typical steady heat transfer system from the hot fluid through a composite wall to a cold fluid. In the regions adjacent to the wall, the fluid temperature changes through the film. Newton's law of cooling states that the heat flux through a film is proportional to the temperature difference that is,

$$\dot{Q} = A\lambda\left(T_w - T_f\right) \tag{6.11}$$

where T_w and T_f are the wall and fluid temperatures, respectively, (K) and "λ" or "h" is the heat transfer coefficient $\left(\text{W}/\text{m}^2\text{K}\right)$.

Figure 6.16 shows schematically the composite flat plates, and then in terms, thermal resistance is,

$$R_{\text{th}_f} = \frac{1}{A\lambda} \tag{6.12}$$

Hence, the heat transfer through the fluid can just be looked on as another resistance term for this particular system which can be expressed as:

$$R_{\text{th}_T} = \frac{1}{A\lambda_1} + \frac{x_a}{k_a A} + \frac{x_b}{k_b A}\frac{1}{A\lambda_2} \tag{6.13}$$

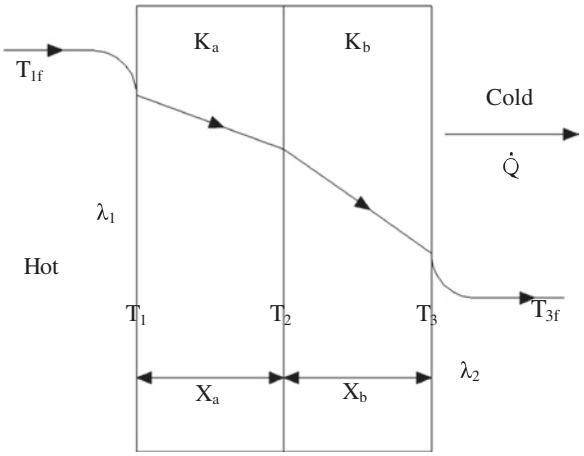


Fig. 6.16 Heat transfer from a solid to a fluid

Therefore, application of Eq. (9.13) renders:

$$\dot{Q} = \frac{A\Delta T}{\frac{1}{A\lambda_1} + \frac{x_a}{k_a A} + \frac{x_b}{k_b A} + \frac{1}{A\lambda_2}} \tag{6.14}$$

Radial Steady-state Conduction Through the Wall of a Tube and a Composite Thick

Radial Steady Flow Through the Wall of a Tube

Heat flows radially through the wall of a tube at steady-state and at a constant temperature with direction of the heat flowing normal to the axis of the tube. The rate of conduction at the steady-state condition through the wall can be estimated at a unit length rather than a surface area which makes it more expedient. The increase of area for consecutive layers with radius implies that the temperature gradient decreases with the radius. Figure 6.17 shows schematically the temperature distribution and the possible flow patterns.

If the thickness of an elemental cylindrical tube is dr and the area of flow per unit length of the tube is $2\pi r$, with the temperature gradient normal to the tube $\frac{dT}{dr}$, then using Fourier's law, the heat flow is $\dot{Q} = 2\pi kr\frac{dT}{dr}$. Also by integration, with the proviso that $\dot{Q}$ is independent of, r, the temperature difference can be expressed as:

$$\Delta T = \frac{\log_e\left\{\frac{r_2}{r_1}\right\}}{2\pi k}\dot{Q} \tag{6.15}$$

Or in terms of thermal resistance, Eq. 6.15 can be written as:

$$R_{\text{th}} = \frac{\log_e\left\{\frac{r_2}{r_1}\right\}}{2\pi k} \tag{6.16}$$

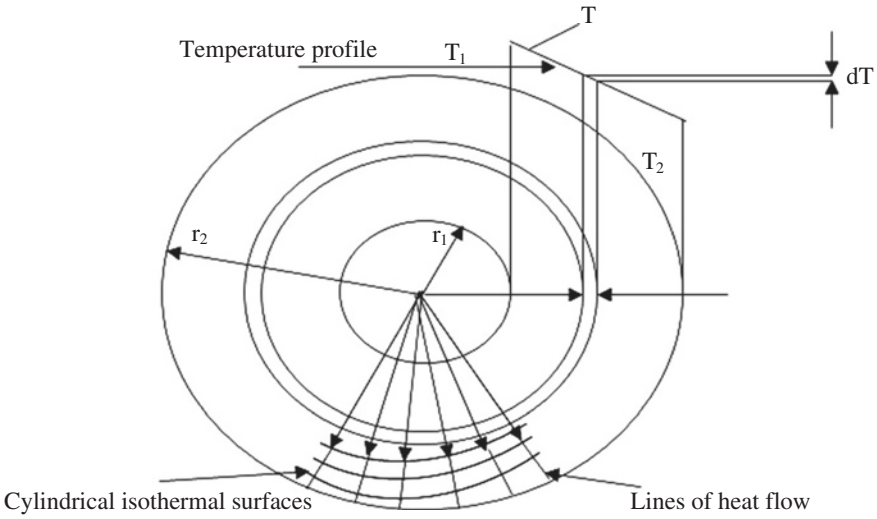
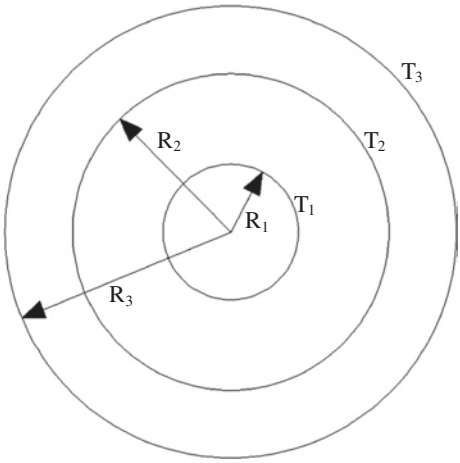


Fig. 6.17 Radial temperature profile of a tube [5]

Radial Steady Conduction Through the Wall of a Composite Thick Cylinder
Consider a thick composite cylinder as shown in the diagram for the fluid inside and outside the tube, Newton's law of cooling still applies, and hence, this can be considered as an extra resistance term, per unit length, thus in terms of thermal resistance, the radial conduction through the wall of a composite thick cylinder can be written as (Fig. 6.18):

$$R_{\text{th}_T} = \frac{1}{2\pi}\left\{\frac{1}{r_1\lambda_1} + \sum_{i=1}^{n}\frac{\log_e\left\{\frac{r_1+1}{r_i}\right\}}{k_i} + \frac{1}{r_n\lambda_n}\right\} \tag{6.17}$$

Fig. 6.18 Composite tube



Overall Heat Transfer Coefficient

Where pipes are concerned, it is also useful to show the heat transfer in terms of an overall heat transfer coefficient $U\left(\mathrm{W}/\mathrm{m}^2\mathrm{K}\right)$; thus,

$$\dot{Q} = AU\Delta T \tag{6.18}$$

Hence, using Eq. (6.17) and combining with (6.18), the overall heat transfer coefficient can be expressed as

$$\frac{\Delta T}{R_{\mathrm{th}_T}} = AU\Delta T \tag{6.19}$$

This overall heat transfer coefficient can be envisaged as an overall heat conduction term. Furthermore, since A is an area of contact with the fluid, as far as pipes are concerned, this area may be calculated with respect to an inner or outer diameter. Since the area of contact per unit length is given by $A = 2\pi r$, this leads to the following definitions.

$$U_i = \frac{1}{A_i R_{\mathrm{th}_T}}$$

where $A_i =$ the inner area; hence,

$$U_i = \frac{1}{2\pi r_i R_{\mathrm{th}_T}} \tag{6.20}$$

Likewise for the outer diameter

$$U_o = \frac{1}{2\pi r_o R_{\mathrm{th}_T}} \tag{6.21}$$

General Equation of Heat Conduction (With Internal Heat Generation)

Consider the volumetric element of the typical conducting material shown in Fig. 6.19 as follows:

Heat transfer rate into the cube in the x direction.

$$\delta Q_x = -kx\frac{\partial t}{\partial x} \cdot \delta y \cdot \delta z$$

at $(x + \delta x)$, the heat transfer rate has changed.

$$\begin{aligned}\delta Q(x+\delta x) &= \delta Qx + \frac{\partial}{\partial x}(\delta Qx) \cdot \delta x. \\ &= \delta Qx + \frac{\partial}{\partial x}\left(-kx\frac{\partial t}{\partial x} \cdot \delta y \cdot \delta z\right)\delta x.\end{aligned}$$

Therefore, the difference between the entering and leaving value:

$$\delta Qx - \delta Q_{(x+\delta x)} = \frac{\partial}{\partial x}\left(kx\frac{\partial t}{\partial x}\right) \cdot \delta y \cdot \delta z \cdot \delta x$$

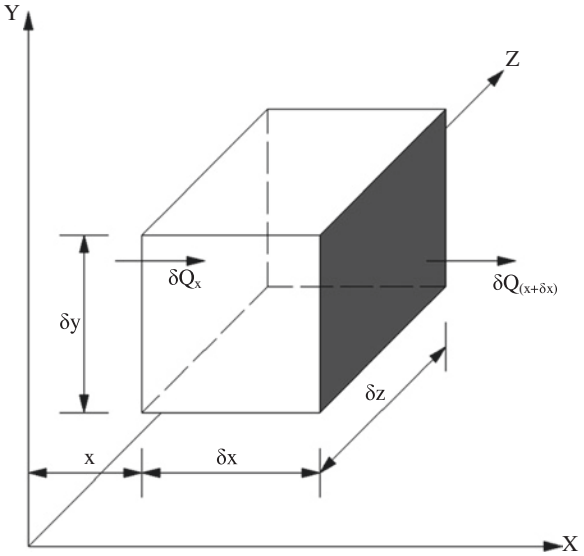


Fig. 6.19 Typical volumetric element (cartesian coordinates)

A similar expression can be obtained for the *y* and *z* directions.
Hence, the net heat transfer inflow rate is,

$$= \left[\frac{\partial}{\partial x}\left(kx\frac{\partial t}{\partial x}\right) + \frac{\partial}{\partial y}\left(ky\frac{\partial t}{\partial y}\right) + \frac{\partial}{\partial z}\left(kz\frac{\partial t}{\partial z}\right)\right]\delta y \cdot \delta z \cdot \delta x\cdot$$

Assume there is an internal heat generation Q_i*/unit-time and volume,* e.g. electrical resistance heating.

$$\text{Rate of heat generation} + \text{influx} = \text{rate of increased energy}$$

Therefore,

$$Q_i\delta y \cdot \delta z\delta x + \left[\frac{\partial}{\partial x}\left(kx\frac{\partial t}{\partial x}\right) + \frac{\partial}{\partial y}\left(ky\frac{\partial t}{\partial y}\right) + \frac{\partial}{\partial z}\left(kz\frac{\partial t}{\partial z}\right)\right]\delta y\delta z\delta x = \rho C_\rho \delta y\delta z\delta x\frac{\partial t}{\partial T} \tag{6.22}$$

where C_p is the specific heat capacity (J/KgK), ρ is the density (kg/m^3) and T is time (s).
For k constant in directions *x, y* and *z,* then Eq. (9.23) can be expressed as,

$$\frac{\partial t}{\partial T} = \frac{Qi}{\rho Cp} + \frac{k}{\rho Cp}\left[\frac{\partial^2 t}{\partial x^2} + \frac{\partial^2 t}{\partial y^2} + \frac{\partial^2 t}{\partial z^2}\right] \tag{6.23}$$

where the thermal diffusivity, α, is:

$$\alpha = \frac{k}{\rho Cp}\left(\frac{\text{m}^2}{\text{h}} \text{ or } \frac{\text{m}^2}{\text{s}}\right) \tag{6.24}$$

Equation (6.25) is the ratio of thermal conductivity to density and specific heat.
In conduction problems, there is a warm-up (Transient) period before the steady-state condition is arrived at

α is an important property in transient conduction
k is an important property in steady conduction.

The above differential equation can be reduced to suit various particular cases.
In summary, the following equations can be applied to 1, 2 and 3 steady and transient conditions.
One-dimensional transient conduction without internal heat generation:

$$\frac{\partial t}{\partial T} = \alpha \left[\frac{\partial^2 t}{\partial x^2}\right] \tag{6.25}$$

One-dimensional steady conduction with internal heat generation:

$$O = \frac{Q_i}{\rho Cp} + \alpha \left[\frac{\partial^2 t}{\partial x^2}\right] \tag{6.26}$$

One-dimensional and two-dimensional steady conduction without internal heat generation are, respectively:

$$O = \left[\frac{\partial^2 t}{\partial x^2}\right] \tag{6.27}$$

$$O = \left[\frac{\partial^2 t}{\partial x^2} + \frac{\partial^2 t}{\partial y^2}\right] \tag{6.28}$$

Three-dimensional steady conduction without heat generation is:

$$O = \left[\frac{\partial^2 t}{\partial x^2} + \frac{\partial^2 t}{\partial y^2} + \frac{\partial^2 t}{\partial z^2}\right] \tag{6.29}$$

Convective Heat Transfer

Nature of Convection
In convective heat transfer, thermal energy is carried by the motion of the fluid. As common observation indicates, high flow velocities result in high heat transfer rates as shown schematically in Fig. 6.20. In order to evaluate the convective heat transfer from a fluid to a solid surface, the wall heat transfer coefficient *h* (or "λ") is employed as

$$h = \frac{q_w}{T_w - T_F}$$
$$h = q_w / A\left(T_w - T_f\right) \tag{6.30}$$

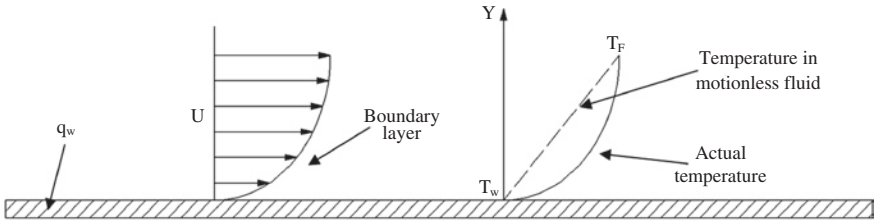


Fig. 6.20 Velocity and temperature profiles in a fluid boundary layer

where q_w is heat transfer rate per unit area (W/m^2), A is surface area (m^2), T_w = wall temperature (K) and T_F = a representative (e.g. ambient) fluid temperature away from the wall (K).

In engineering calculations, the use of the heat transfer coefficient h provides a general method for the evaluation of convective heat transfer from a fluid to a solid.

In the case of **external flows**, T_F is T, the temperature of the fluid far enough from the wall to be unaffected by its presence. In the case of **internal flows** (e.g. pipe flow), T_F is taken as the average or bulk temperature of the fluid T_B. The **precise calculation** of h *(*or *"*λ*")* requires the solution of the full heat transport equation containing convection. The convection equation can be constructed by considering the energy balance for an infinitesimally small control volume within the fluid. Readers, if required, should refer to an appropriate book on heat transfer or consult the Refs. [5, 6] given at the end of this chapter, for a derivation of convection equation as well as other related areas including dimensional analysis. A more widely used (and simpler) approach to evaluate h *(*or *"*λ*") is* from experimental correlations which are generalised using dimensional analysis.
It is, however, useful here to provide a brief explanation of two different convection heat transfers. Convection can be subdivided into two types: (i) forced convection and (ii) natural convection

(i) *Forced Convection*

In forced convection, the motion of the fluid is caused by external factors; the fluid is "forced" to move, e.g. by a pressure difference as shown in Fig. 6.21. The presence of fluid flow then also assists heat transfer. The pressure gradient drives the fluid. The fluid carries thermal energy which is transmitted to the solid wall.

(ii) *Natural (or Free) Convection*

In the absence of external forces, the temperature gradients in the fluid lead to density differences which cause buoyancy-induced motion. This motion then carries thermal energy either to or away from the solid surface, as shown schematically previously in Fig. 6.20. Under the circumstances for the heat transfer from hot surface to cold stationary air, the near-wall fluid is at a higher temperature than the fluid further away from the wall. The near-wall fluid has, therefore, a lower density than the fluid further away.

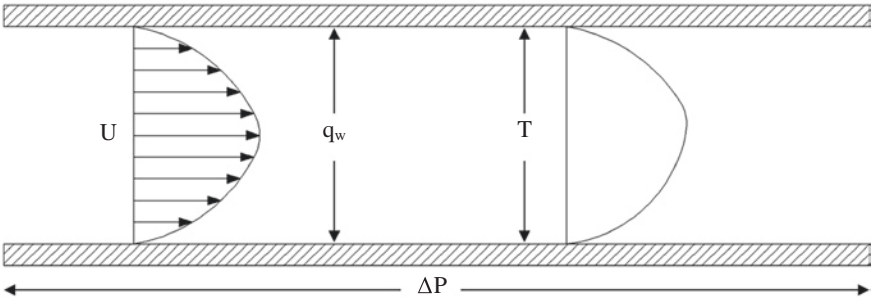


Fig. 6.21 Velocity profile in a tube

Radiation Heat Transfer

Thermal Radiation

If a heated body is completely isolated from any contacting medium, i.e. placed in a vacuum, then its temperature is observed to decrease with time, again showing a loss of energy. This mode of heat transfer is evidently quite different from conduction and convection, where heat transfer relies on contact between the heated body and a contacting medium. Thermal radiation is the transfer of energy by virtue of electromagnetic wave motion. The electromagnetic wave spectrum is wide, and common types of E-M waves are classified by their wave length (Fig. 6.22).

Thermal radiation is carried by those waves which fall within the infrared band. The infrared waves themselves result from electron transitions between electron shells within the atoms of the radiating body.

Absorptivity, Reflectivity and Transmissivity

When radiant energy falls upon a body, part or all of it may be absorbed, part or all of it may be reflected, and part or all of it may be transmitted (passed through the body). This can be shown schematically in Fig. 6.23.

Black Body Radiation

Radiation striking a body can be absorbed by the body, reflected from the body or transmitted through the body. The fractions of the radiation absorbed, reflected and

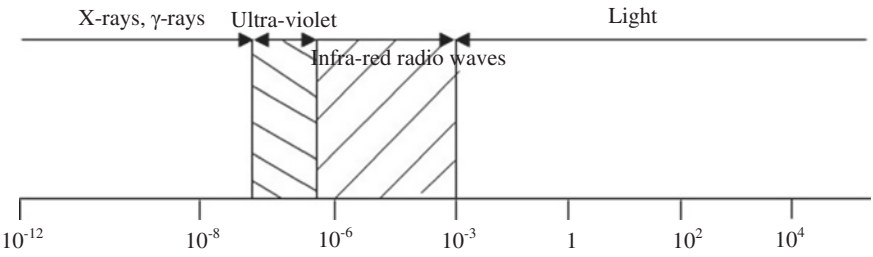


Fig. 6.22 The E–M spectrum

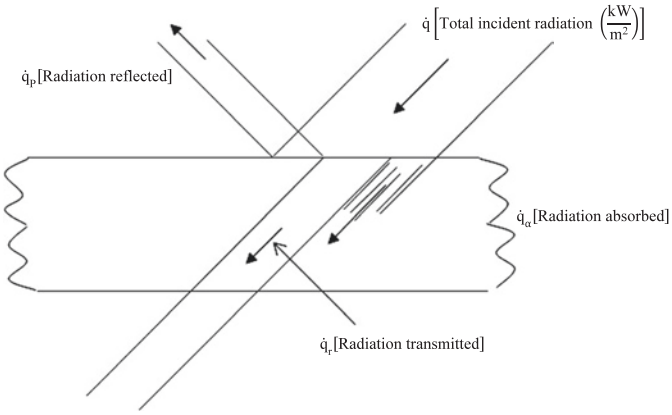


Fig. 6.23 Different radiant energy

transmitted are called the absorptivity, α, the reflectivity ρ, and the transmissivity, τ, respectively. Then, we have$\alpha + \rho + \tau = 1$. For most solids and liquids encountered in engineering, the amount of radiation transmitted through a substance is negligible. Therefore, $\alpha + \rho = 1$. An ideal black body absorbs all the radiation which falls upon it. For a black body, $\alpha =$ land $\rho = 0$. It should be noted that the term black does not necessarily mean black to the eye. *Kirchoff's law* states that the emissivity of a body radiating energy at a temperature T is equal to the absorptivity of the body when receiving energy from a source at a temperature T.

The Grey Body

In considering radiation so far, it has been assumed that the energy emitted by thermal radiation is the same for all wavelengths of radiation. This is not the case. Surfaces in practice are very often assumed to have a constant emissivity over all wavelengths and for all temperatures such an ideal surface is called a **grey body**. Then, for a grey body, $\alpha = \varepsilon$ at all temperatures where $\alpha$and ε are the total absorptivity and the total emissivity, respectively, over all wavelengths. In practice, although a suitable total value of the emissivity may be taken for a large number of industrial surfaces (i.e. furnaces, metal making) over a wide range of wavelengths, there is still a variation of total emissivity with temperature [5].

The Stefan–Boltzman Law

It was found experimentally by Stefan and proved theoretically by Boltzmann that the emissive power of a black body is directly proportional to the fourth power of its absolute temperature. If the emissivity of the body at T_1 is largely different from the emissivity of the body at T_2, then the approximate of the grey body may not be sufficiently accurate. In that case, it is a good approximation to make the absorptivity of the body (1) when receiving radiation from a source at T_2 as being equal to the emissivity of body (1) when emitting radiation at T_2 as shown schematically in Fig. 6.24. Thus, heat transfer by radiation can be defined as:

$$\dot{Q} = \varepsilon\sigma A(T_1^4 - T^4)\ (W) \tag{6.31}$$

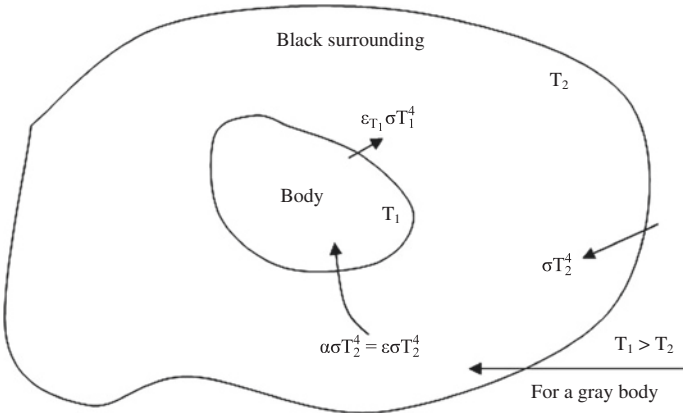


Fig. 6.24 Black body and heat transfer radiation

where ε is the emissivity, σ is Setfan–Boltzman constant (56.7×10^{-12} kW/m^2 K^4), A is the surface area (m^2) and T is absolute temperature (K).

The absorptivity, while depending mainly on the temperature of the source of radiation, also depends on the temperature of the surface itself. For most metals, this factor can be important and it has been shown that the absorptivity of two metal surfaces at T_1 for radiation from a source at T_2 is approximately equal to the emissivity of the surface when at temperature T_3 given by $T_3 = \sqrt{T_1 T_2}$. Further information on radiation heat transfer is also given in [5, 6, 11] and various published works.

It should be also noted that radiation from gases and flames are complex and need to be separately treated. Readers are advised to consult the appropriate publications. Generally speaking, gases play an important role in transferring heat, are extremely selective emitters and cannot be considered as grey [5]. The radiation from flame, however, can be estimated subject to knowing the density and distribution of the suspended solid particles and thus assume the flame is of the black nature. In the Chap. (7), the concept of combustion, luminous (yellow) and non-luminous (blue) flame is described. Effectively the yellow flame could transfer its energy to other bodies by radiation more readily than the blue flame. The analogy of this concept is gas fires for space heating in which the blue flames are surrounded by refractory materials. When these refractory materials are subjected to the corresponding heat, it enables the radiation from the fire to be more efficiently transmitted (e.g. see also previous preceding sections in this chapter).

6.2.2 Heat Exchangers

6.2.2.1 Introduction

The three modes of heat transfer that were briefly described in Sect. 6.2.2.1 can be combined and treated concurrently by considering an overall heat transfer coefficient. As the name implies, a heat exchanger is a device that aims to promote exchange

of thermal energy between two fluids. Every day experience alone (car-radiator, domestic heating radiators, and refrigerators) is sufficient to illustrate the important role of such components. There is a variety of different types of heat exchanger available. The need for different designs partly arises from the different types of fluid involved (gas-to-gas, gas-to-liquid and liquid-to-liquid) and partly because of the different applications to which heat exchangers are used. In an aeronautical application, weight and size may be the limiting factor, whereas in other situations, thermal effectiveness and capital costs may be the main considerations. In the following analysis, different types of heat exchangers are described, mainly, from atheoretical point view which can therefore be utilised in natural gas processing plant as appropriate.

Types of Heat Exchangers

1. *Double-pipe Heat Exchanger*

This is the simplest heat exchanger type. When the pipe and annulus flows are in the same direction, then we have a *parallel*-flow situation as shown schematically in Fig. 6.25. When Fluid A flows in the opposite direction from Fluid B, then we use the term *counter-flow* heat exchanger.

The thermal resistance between the two fluids is given by:

$$R_{\text{th}} = \frac{l}{hiAc} + \frac{\ell h(r_o/r_i)}{2\pi kL} + \frac{1}{h_o A_6}$$

Therefore, overall heat transfer coefficient U_{ol} based on our radius is

$$U_o = 1/\left\{\frac{A_o}{A_l}\frac{l}{hl} + \frac{A_o \ln(r_o/r_i)}{2\pi kL} + \frac{1}{h_o}\right\} \tag{6.32}$$

where h_i and h_o can be determined from correlations.

In a *parallel-flow* situation, there is a maximum temperature difference at the entrance and a minimum temperature difference at the exit (between the hot and cold fluid). In *counter-flow* situations, the temperature differences between the hot and cold fluids remain more uniform (Fig. 6.26).

2. *Shell-and-Tube Heat Exchanger*

Such heat exchangers are widely used in chemical-process industries. One fluid flows inside the tubes, and the other fluid is forced over and outside the tubes through the shell. The baffles force the shell-side fluid to flow across the tubes and thus promote heat transfer. A variety of baffles can be used in practice. A multiple number of tube passes (4, 6, 8, etc.) can also be present. A two-shell arrangement is also possible as shown in Fig. 6.27.

3. *Cross-flow Heat Exchangers*

Cross-flow heat exchangers are commonly used in gas cooling and heating applications. The fluid inside the tubes would normally be a liquid and the outside fluid a gas as depicted in Fig. 6.28.

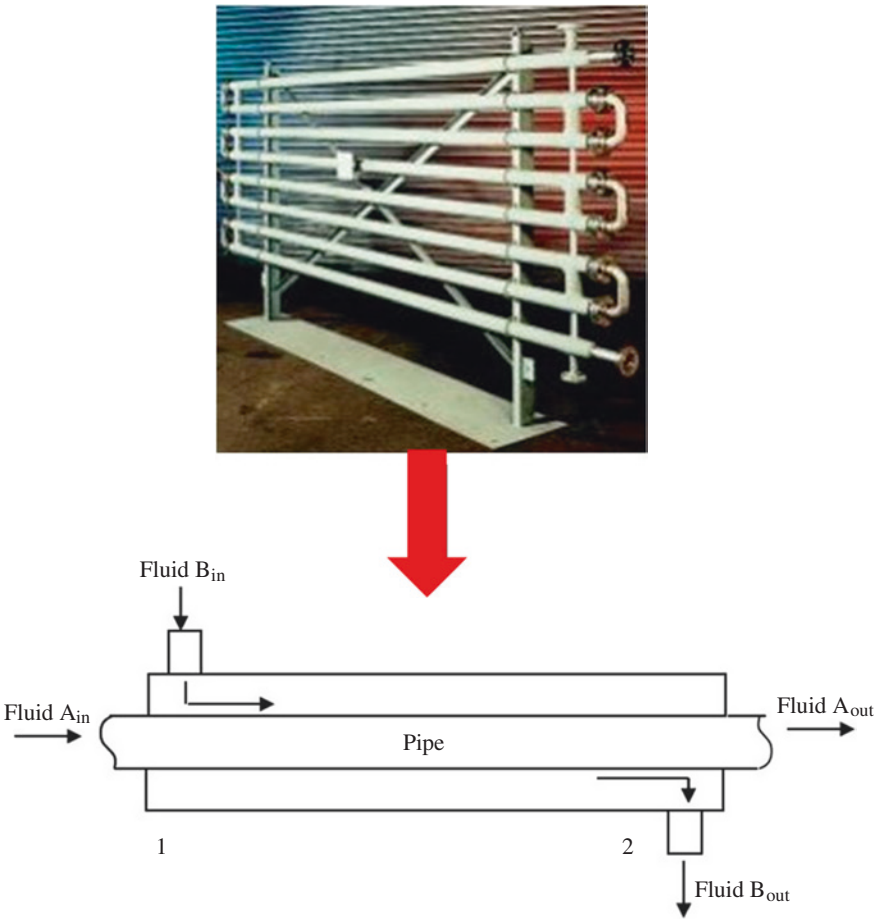


Fig. 6.25 Schematic diagram of double-pipe heat exchanger

4. *Compact Heat Exchangers*

These are basically gas-to-gas heat exchangers. Because of the lower overall heat transfer coefficient (associated with gases), gas-to-gas heat exchangers have a higher surface area per unit volume than other types.

Their examination is beyond the scope of this book. Readers should consult appropriate Refs. [7–9].

The Log-Mean Temperature Difference (LMDT)

This is one of the available calculation methods for heat exchanger design. There are a set of objectives for this method. Given a heat exchanger of known surface

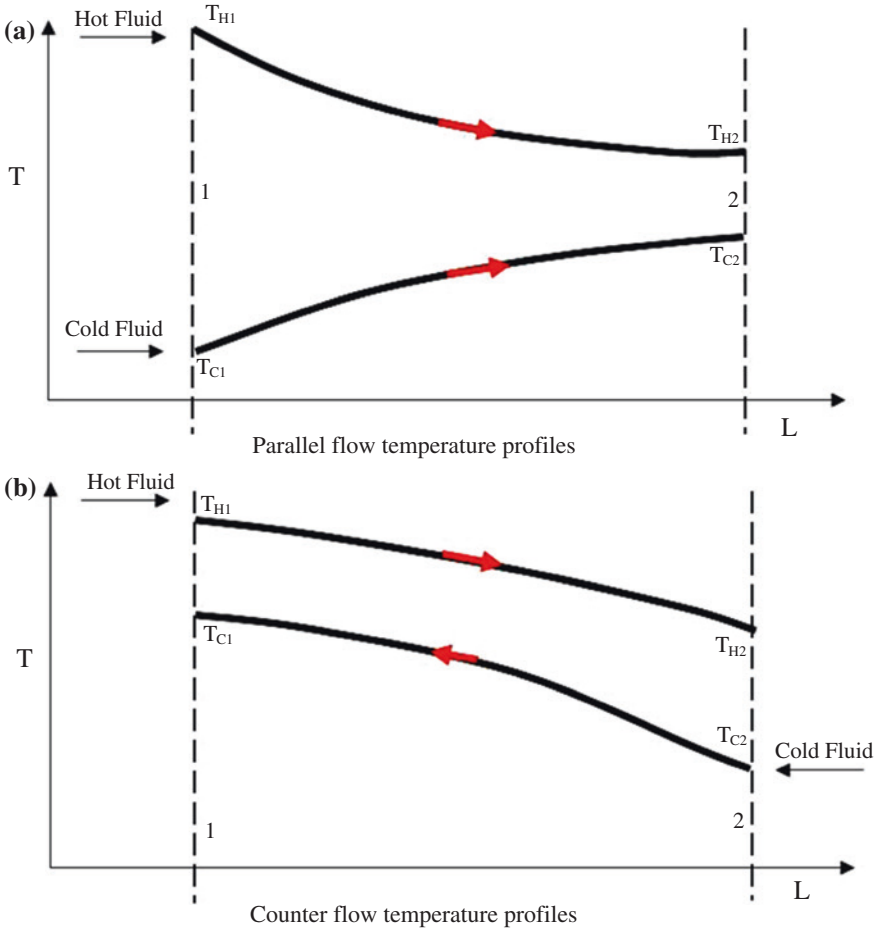


Fig. 6.26 Parallel flow temperature profiles (**a**) and counter flow temperature profiles (**b**) heat exchanger

area, A, overall heat transfer rate coefficient, U, and given the entry and exit temperatures of the two fluids a *Mean Temperature difference*, $\boldsymbol{\Delta T_m}$, is required so that

$$Q = U \cdot A \cdot \Delta T_m \tag{6.34}$$

A mean temperature difference is needed because the temperatures of both the hot and cold fluid vary from inlet to outlet. *The Log-Mean Temperature Difference (LMTD)* can be derived for *a double-pipe parallel-flow heat exchanger* as shown in the following with reference to Fig. 6.29.

Across a small element of the heat exchange of area ΔA, the heat transfer rate ΔQ is given by

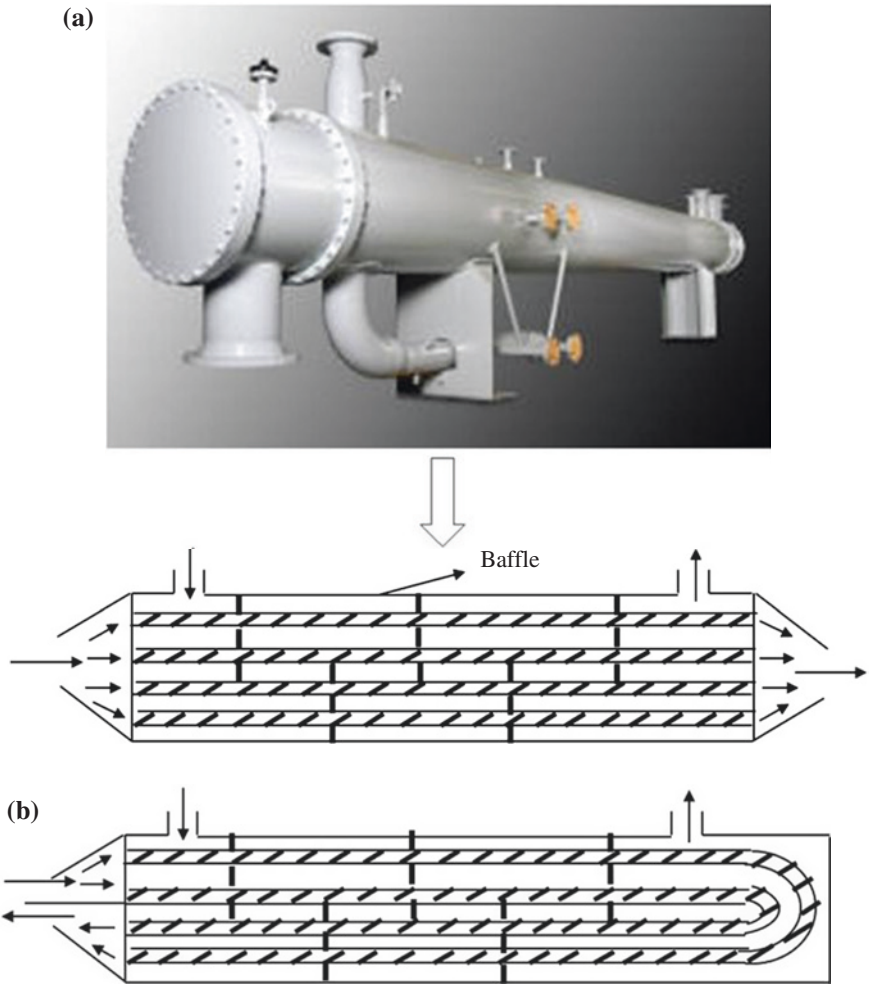


Fig. 6.27 Shell-and-tube heat exchanger **a** one tube pass and **b** two tube passes

$$\Delta Q_x = U \cdot \Delta A_x (T_h - T_c)_x \tag{6.35}$$

From the energy balance,

$$\Delta Q = -\dot{m}_h C_h \Delta T_h = \dot{m} c_c \cdot \Delta T_c \tag{6.36}$$

where $\dot{m}$ is mass flow rate (kg/s), c *is* specific heat capacity (J/kgK).

Combining 6.35 and 6.36 and by integration and when $x = x_1$, $T_h = T_{h1}$, $T_c = T_{c1}$, $A_x = 0$ and Constant $= \ln(T_{h1} - T_{c1})$, then LMTD can be expressed as:

$$\Delta T_m = \frac{(T_{h2} - T_{c2}) - (T_{h1} - T_{c1})}{\ell_n[(T_{h2} - T_{c2})/(T_{h1} - T_{c1})]}$$

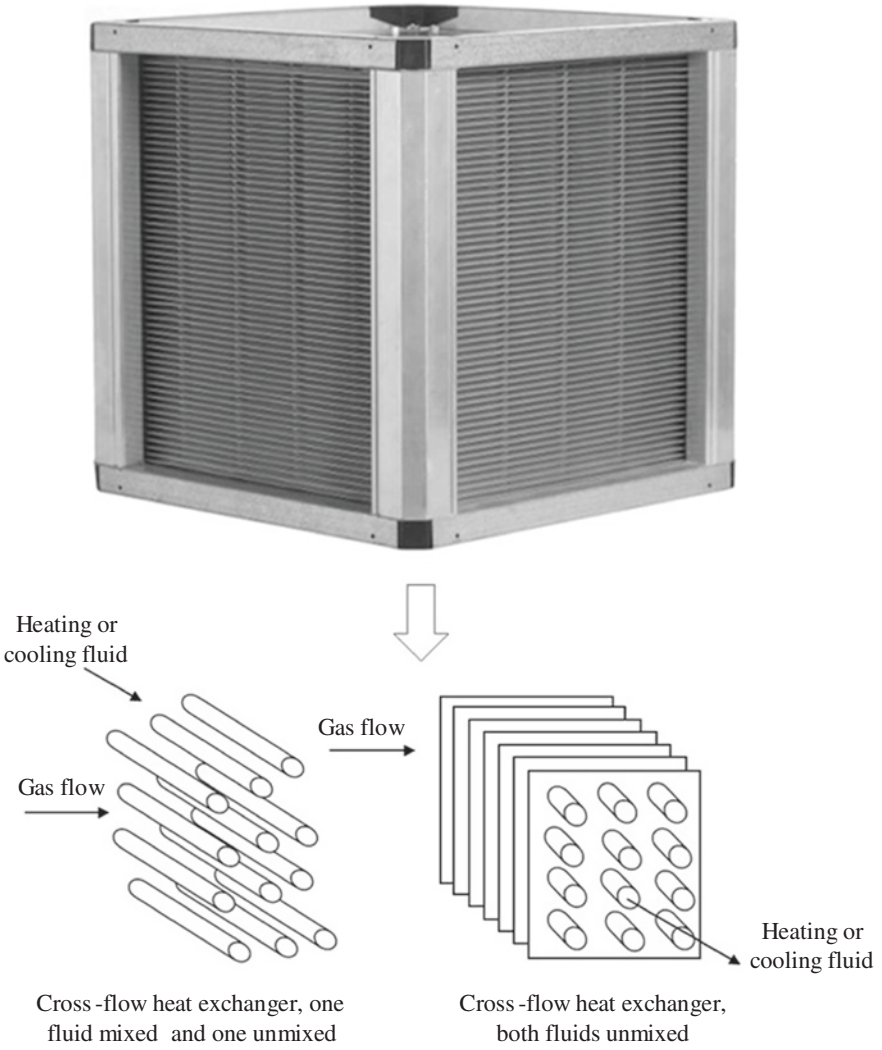


Fig. 6.28 Cross-flow heat exchangers

or
ΔT_m is called **LMDT**

$$\Delta T_m = \frac{\Delta T_2 - \Delta T_1}{\ell_n(\Delta T_2/\Delta T_1)} \tag{6.37}$$

Equation (9.37) is also applicable to a *counter-flow double-pipe heat exchanger* as shown in Fig. 6.30.

Thus, the **LMDT** for parallel flow can be expressed as

$$\Delta T_m = \frac{\Delta T_2 - \Delta T_1}{\ell_n(\Delta T_2/\Delta T_1)} \tag{6.38}$$

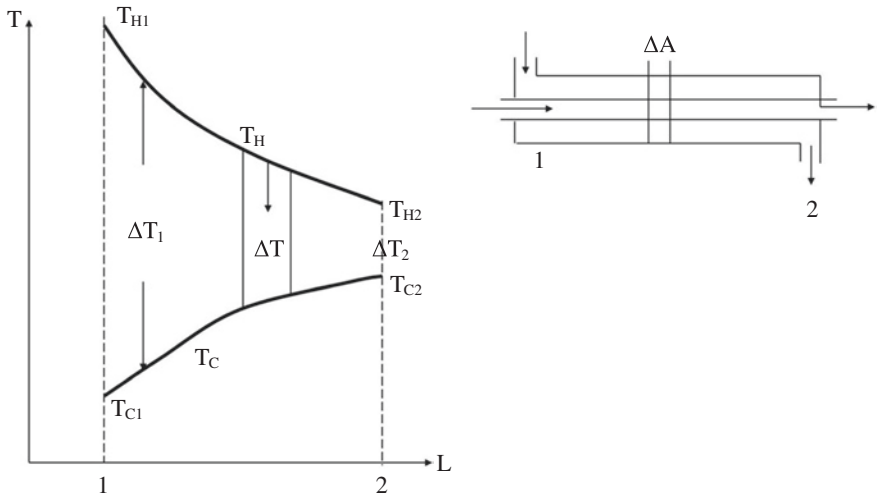


Fig. 6.29 Parallel-flow heat exchanger

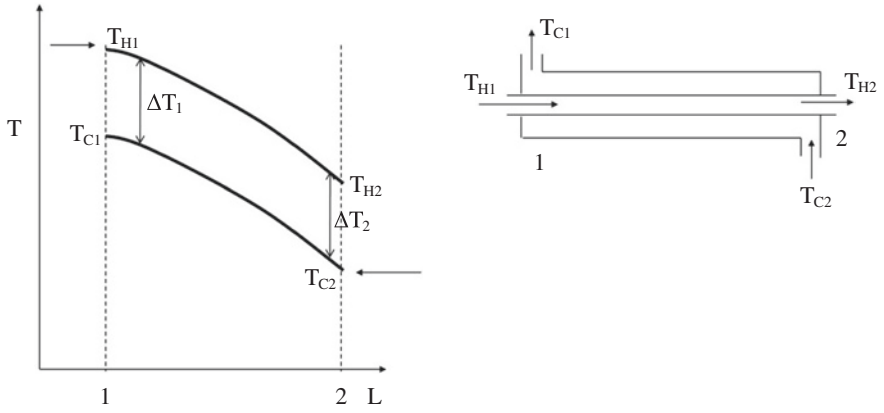


Fig. 6.30 Counter-flow heat exchanger

For other types of heat exchangers, if T_1 and t_1 are the two fluid inlet temperatures, then T_2 and t_2 are the corresponding exit temperatures as typified in Fig. 6.31.

The corresponding LMTD can thus be written as

$$\Delta T_m = \frac{F\{(T_1 - t_2) - (T_2 - t_1)\}}{\ell_n[(T_1 - t_2)/(T_2 - t_1)]} \tag{6.39}$$

where F is a correction factor.

The value of F depends on the type of heat exchanger used and also on the temperature ratios P and R. The correction factor F allows for the fact that flow may not be true parallel flow or true counter-flow. These correction factors are

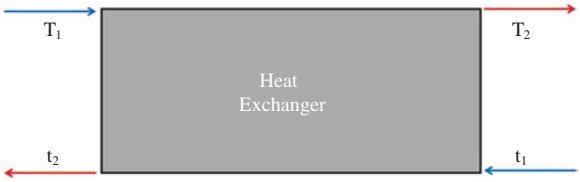


Fig. 6.31 Heat exchanger with two fluid inlet temperature

functions of both flow and temperatures and temperature differences between shell-and-tube fluids and are presented as graphs (see Fig. 6.32). These graphs are in various standards and publications such as Tubular Exchanger Manufacturers Associations (i.e. TEMA) [10], readers should refer to the standards appropriate to their respective exchangers. Note that $P = \frac{t_2 - t_1}{T_1 - t_1}$ and $R = \frac{T_1 - T_2}{t_2 - t_1}$.

The values of F for shell-and-tube and also for cross-flow heat exchangers are given in Fig. 6.31a–c.

Effectiveness and Number of Transfer Unit Method

The LMTD method discussed in the previous section is not easy to apply when all inlet and outlet temperatures are not known. An iterative solution procedure would have to be employed. A more generally applicable method for heat exchanger analysis is the effectiveness NTU method. The objective is to determine the heat exchanger effectiveness ε, where ε can be defined as

$$\varepsilon = \frac{\text{Actual Heat Transfer}}{\text{Maximum Possible Heat Transfer}}$$

The actual heat transfer Q is either the energy lost by the hot fluid or the energy gained by the cold fluid.

$$Q = m_{\dot{h}} c_h (T_{h1} - T_{h2}) = m_{\dot{h}} c_h (T_{c2} - T_{c1}) \tag{6.40}$$

The *maximum possible* heat transfer for the heat exchanger could be attained if the fluid with *the minimum mC* value was to under go a temperature change equal to the maximum temperature difference. The maximum temperature difference is the difference between the hot and cold fluid inlet temperatures T_{h1} and T_{c1}.

$$Q\text{max} = (\dot{m}\, c)_{\text{min}} (T_{h1} - T_{c1})$$

It is necessary to use the minimum mC fluid because it is physically impossible for the fluid with the maximum mC value to undergo the maximum temperature change. Therefore,

$$\varepsilon = \frac{\Delta T\,(\text{minimum fluid})}{(T_{h1} - T_{c1})}$$

Having defined ε, the problem is how to determine its value for a given situation. If $C = \dot{m}\, c$, it will be shown that ε is a function of Cmin/Cmax and of UA/Cmin.

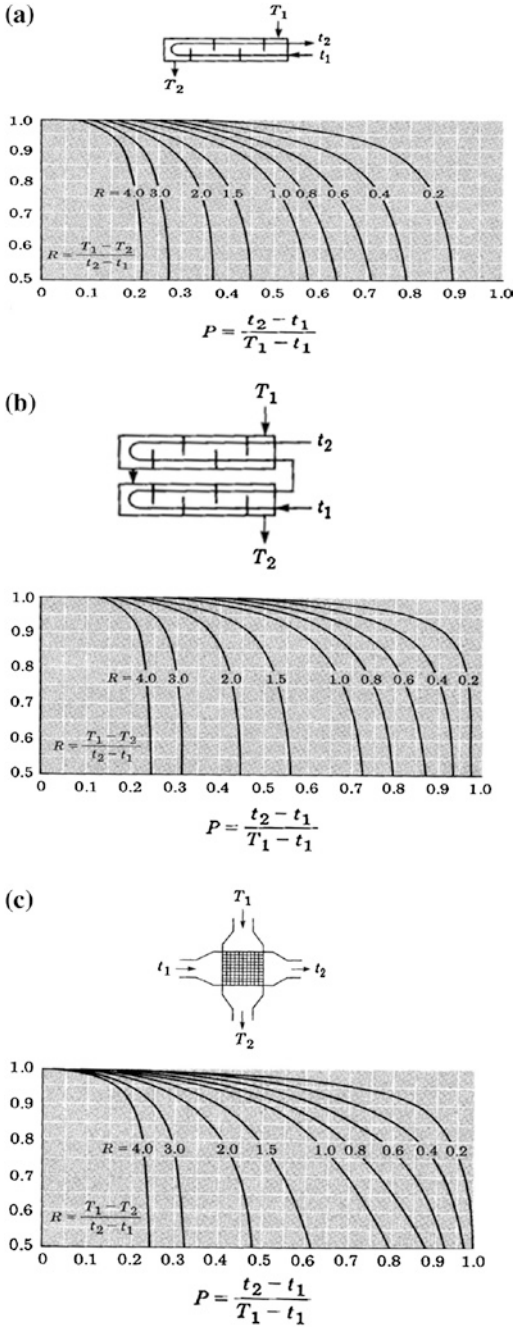


Fig. 6.32 Typical correction factor plot from TEMA for various heat exchanger configurations **a** T on the tube side, T' on the shell side. i = input, o = output. One pass on shell side, any multiple of two passes on tube side. **b** Correction factor plot for exchanger with two-shell passes and four, eight or any multiples of tube passes. **c** Correction factor plot for cross-flow exchanger with both fluids unmixed, laterally

(Note that $C \leq 1$). C is known as the capacity rate and UA/$c$min is called the number of transfer units (NTU). The NTU is a measure of heat exchanger size. This is useful since one of the restrictions of the LMTD method of analysis is that the configuration of hot and cold inlet and outlet temperatures is needed before the design can be done. If not a "guestimation" of the outlet temperatures is made, then an iterative calculation is carried out to "home in" on the final solution.

$$\text{NTU} = \frac{\text{UA}}{m_c C_{\text{pc}}} \text{ when } m_n C_{\text{pn}} > m_c C_{\text{pc}}$$

and

$$\text{NTU} = \frac{\text{UA}}{m_h C_{\text{ph}}} \text{ when } \text{m}_h \text{C}_{\text{ph}} < \text{m}_c \text{C}_{\text{pc}}$$

Effectiveness in a parallel-flow double-pipe can be expressed as

$$\varepsilon = \frac{1 - \text{EXP}[C - N)(1 + c_{\text{min}}/c_{\text{max}})]}{1 + c_{\text{min}}/c_{\text{max}}} \tag{6.41}$$

And for a counter-flow double-pipe heat exchanged, it can be shown that

$$\varepsilon = \frac{1 - \text{EXP}[C - N)(1 + c_{\text{min}}/c_{\text{max}})]}{1 - (c_{\text{min}}/c_{\text{max}})\text{EXP}[(-N)(1 - c_{\text{min}}/c_{\text{max}})]} \tag{6.42}$$

Readers should consult various references for derivations of these equations [7–10]. For heat exchangers of different types, similar expressions can also be derived (either exact or approximate). The value of $N(= \text{UA}/c_{\text{min}})$ in terms of the ε and c_{max} and c_{min} values for various types are also provided in different related publications. In boilers, condensers, and furnaces where the temperature of one of the fluids remains constant, $c_{\text{min}}/c_{\text{max}}$ is effectively zero because c_{max} tends to infinity. The effectiveness of parallel and counter-flow heat exchangers can also be found in [7–9] for example.

6.3 Overview of Combined Heat Power Using Natural Gas

6.3.1 Introduction

On a large scale, the application of combined heat and power (CHP) systems has been slow to take off, mainly due to economic considerations rather than environmental considerations. In mainland Europe, many such schemes are in operation for the supply of district heating. Combined heat and power on a smaller scale has been taken up more readily and systems up to a maximum of 150 kW heating load are commonly referred to as micro-CHP systems. Typical applications of these systems are in hotels, leisure complexes, water treatment plants, etc. Standard packages comprising engines, heat exchangers and control systems are

available from various companies, Fiat, Ford, MAN, WAUKESHA, etc. A typical package comprises an engine of capacity 13.3 L having a gas input of 338 kW and having a heat output of ~180 kW and power of 90 kW or more. The criterion for the application of these units is just the same as for large-scale CHP schemes, i.e. a sufficiently large and constant heat load throughout a large part of the day and throughout the seasons and a similar pattern of demand for electricity.

6.3.2 *Combined Gas and Steam Cycles*

Because the gases leave the gas turbine unit at 450 °C, it is possible to use some of the energy contained in them to generate steam in a Rankin cycle plant. These combined plants, where the gas turbine often uses **natural gas**, are sometimes known as COGAS plants. The boiler pressure is limited in such plants as to provide an adequate temperature difference between the turbine exhaust gases and the return water/generated steam, as shown schematically in Fig. 6.33.

In the consideration of CHP plant, for example, the process integration plays an important role which requires the utilisation of the "*Pinch analysis*" and this should be included in the design process. The term "*Pinch analysis (or Pinch Technology or Heat or Energy Integration)*" was initially introduced by Linnhoff et al. [12] followed by Shenoy [13] and Kemp [14]. The

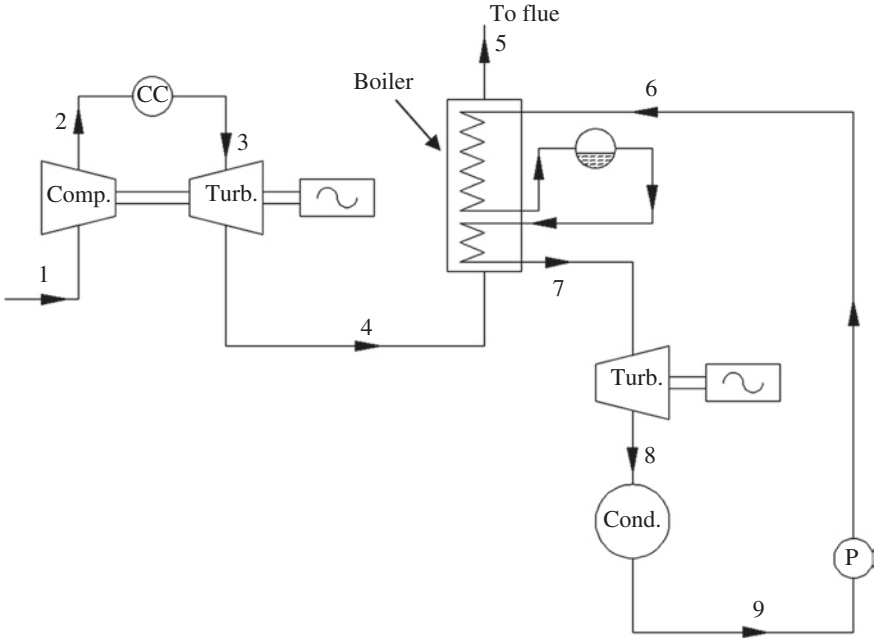


Fig. 6.33 Typical CHP plant

technique is for designing a process to minimise energy consumption and maximise heat recovery. The process integration is not constrained to the design of new plants; it can also be applied to retrofit design and the operation of current systems. Pinch analysis uses a systematic analysis relating to chemical processes and the surrounding utility systems by applying the first and second laws of thermodynamics. The first law of thermodynamics enables the enthalpy changes (ΔH) to be estimated using the energy equation whereas the second law (entropy (S)) identifies the direction of heat flow (hot to cold). A key to the technique is the Pinch temperature, which is effectively the most constrained point in the process. The application of Pinch technology is not limited to the thermal Pinch analysis; it can be equally applicable to mass-exchangers network as well as water minimisation and materials recycle. A simple explanation within a system shown in Fig. 6.33, for example, is where minimum flow rate occurs at the minimum temperature difference. Thus, for this particular plant shown in Fig. 6.33, the flow rate of steam generated in the boiler is limited by the amount of exhaust gas available, and an energy balance is thus necessary on the boiler; hence,

$$m_{\text{gas}} C_{p\text{gas}} (t_4 - t_5) = m_{\text{steam}} (h_7 - h_6) \tag{6.44}$$

where h is the specific enthalpy of steam (J/kg). Note also that the thermal efficiency (ή) of such a plant will be greater than that of the gas turbine plant or the steam plant alone.

$$Z_{Th} = \frac{\dot{W}_{34} + \dot{W}_{78}}{\dot{Q}_{23}} \tag{6.45}$$

Again, referring to the plant shown in Fig. 6.33, the heat transfer across the pinch can be illustrated as in Figs. 6.34 and 6.35 by using temperature (T) and enthalpy (S) cycle diagram.

6.3.3 *Back-pressure Turbine/Pass-out or Extraction Turbine Plant*

In many industries, there is a dual requirement for power and heat, and many systems have been devised to meet this need. These systems can include back-pressure turbine plants, pass-out or extraction steam turbine plants, and large-scale Combined Heat and Power plants (CHP) and micro-CHP plants. Diagrams illustrating the first two examples are shown in Figs. 6.36 and 6.37.

Figure 6.36 may be used when the power and heating loads are matched and steady, and this arrangement could replace two systems. A disadvantage of this plant is that there will be a lower thermal efficiency for the power plant compared with a conventional steam power plant.

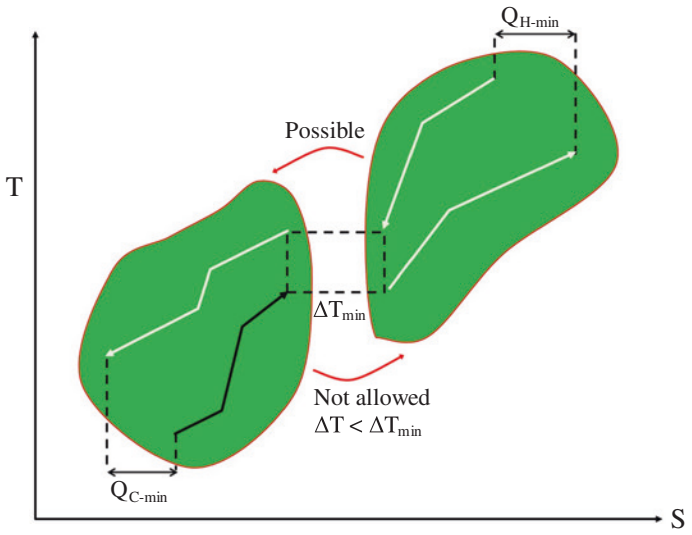


Fig. 6.34 Typical illustration of heat transfer across the pinch [note that minimum temperature difference, $\Delta T_{min,}$ occurs when $T_{Pinch\ (Hot)} > T_{Pinch\ (Cold)}$]

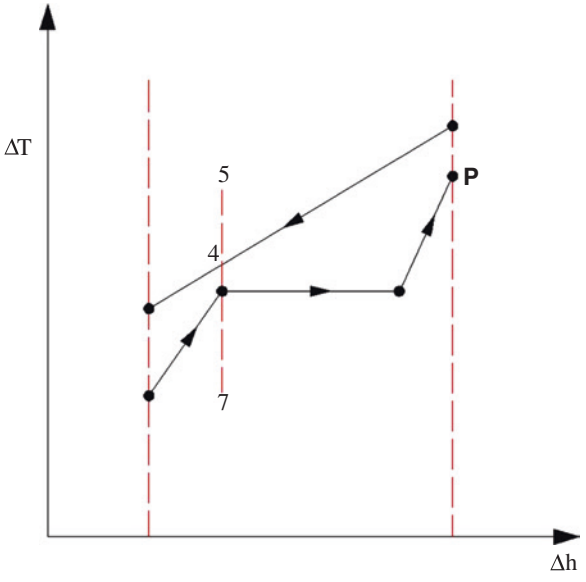


Fig. 6.35 T–H diagrams for CHP plant (Fig. 6.33)

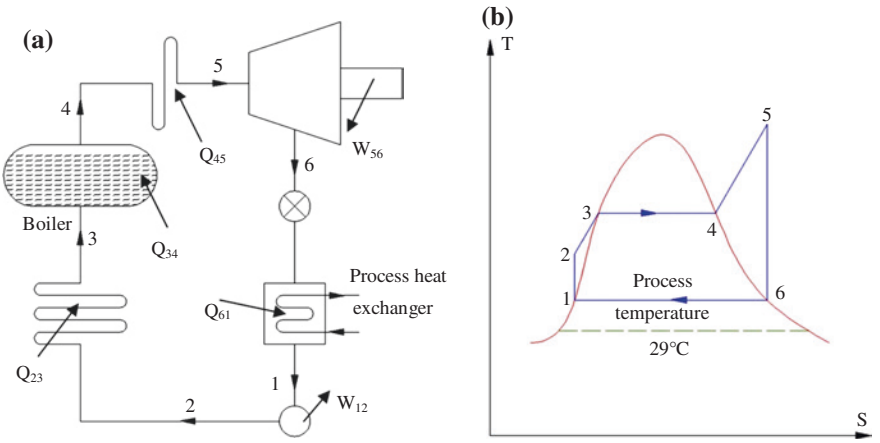


Fig. 6.36 Typical CHP plant (back-pressure turbine) (**a**) temperature (T)–entropy (S) cycle diagram (**b**)

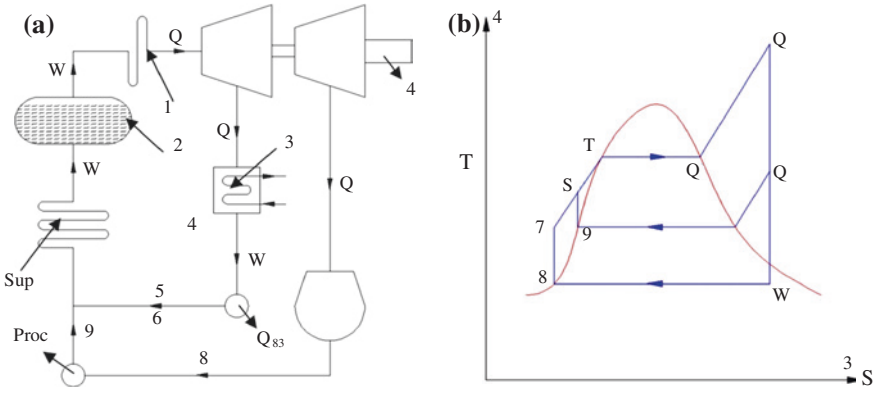


Fig. 6.37 Typical CHP plant pass-out or extraction turbine (**a**) with temperature (T)–entropy (S) cycle (**b**) [15]

Figure 6.37 shows a development of this where some of the steam is expanded "down" to a low pressure corresponding to a temperature of ~29 °C; the remainder is extracted at a pressure appropriate to the process heat requirements. This system can respond to a certain extent to changes in demand for process heat by extracting a variable amount of steam.

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Chapter 7
Fire and Explosion

7.1 Introduction

Compliance with new UK and EU gas safety legislation for chemical processing plants is today a major factor influencing its design and operation. The activities of exploration and production of natural gas are associated with gas transportation, distribution and storage. In industrial, commercial and domestic markets, there are innumerable combusting flows as gas is burned as an end product by customers. In all these activities, accurate assessment of what would happen in the event of an operational or accidental release of gas, particularly where gas dispersion, fire or explosion might be involved, is an essential part of ensuring safe operations.

Accidental release of a flammable mixture can result in a drifting vapour cloud. If ignited, damaging pressures may be generated due to interactions of the cloud with repeated obstacles contained within highly congested regions of chemical processing plants. Liquefied natural gas (LNG) is being transported in large carriers at essentially atmospheric pressure and a temperature of about −160 °C. These carriers deliver the liquid to terminals where it is off-loaded into insulated storage tanks. Should a spill from a carrier tank occur then, if the release were not ignited immediately, the heat absorbed from the water or the surface of the dike would result in the rapid evaporation of the LNG and the formation of a vapour cloud. The resultant hazard to the environment would then be realised upon the ignition and combustion of this vapour cloud. When ignition occurs, a flame will travel through the cloud, releasing and transferring energy, and thus leading to thermal damage. If the speed at which the flame travels is sufficiently high, then the combustion process can also lead to mechanical damage through the generation of pressure waves of significant amplitude. Such an event has become known as a "vapour cloud explosion". Figure 7.1 shows typical consequences of combustible gas into the atmosphere. These events can occur inside equipment or pipes, in building, in onshore or offshore chemical processing plants.

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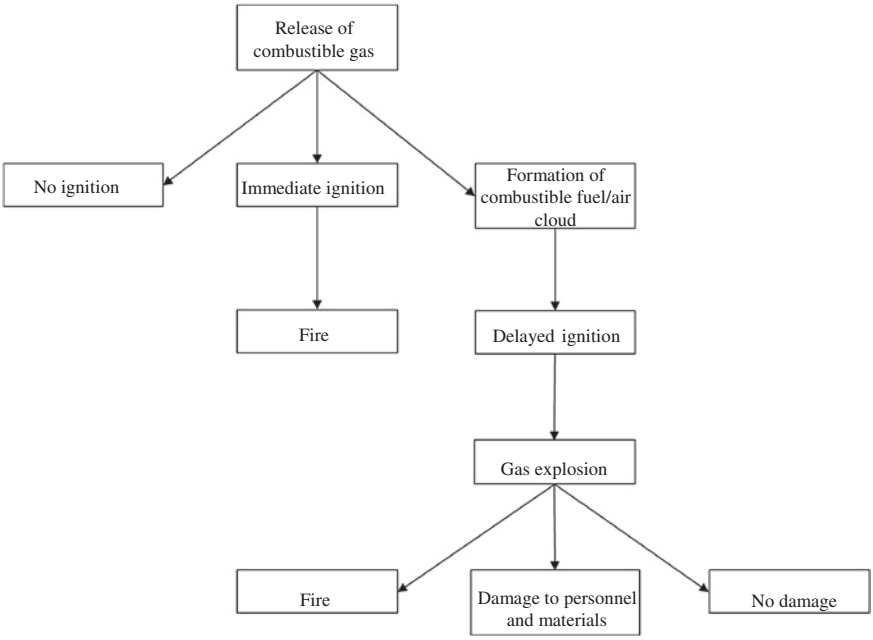


Fig. 7.1 Typical consequences of accidental release of combustible gas into atmosphere

Three elements are necessarily for fires or explosions to occur. These are:

- Fuel
- Oxidiser
- Ignition source

When these three elements are present at the relevant level, combustion will occur. In fires, energy released due to the combustion process will be released slowly, while in explosions, energy release is very rapid. It should be noted that fires can result from explosion and explosion can result from fire.

7.2 Examples of Actual Incidents of Vapour Cloud Explosions

Losses in the oil/gas process industry are mostly caused by vapour cloud explosions. Details of the individual accidental explosions show a large variety in size of the explosion and loss experience. It was learned that gas explosions have a tendency to repeat themselves in similar conditions. Example of such explosion incidents includes the Flixborough onshore plant and Piper Alpha offshore module.

Nipro Plant, Flixborough, U.K., June 1974

In the Flixborough, UK, incident, about 40 tonnes of cyclohexane vapour was released at high pressure due to the failure of a temporary pipe connection between 2 vessels. A large cloud was formed over the plant, some reactor vessels, a pipe rack and a large region of open site. The flammable cloud was ignited about one minute or so after the release. Following ignition, an explosion took place, which is estimated, from the blast damage as equivalent to about 16 tonnes of TNT. Figure 7.2 shows the aftermath of the explosion at Flixborough in the UK on June 1974.

The damage recorded that the congested region of the plant and 2 adjacent sections of the plant were almost entirely responsible for the blast. The plant was totally destroyed, and 28 persons were killed, and 36 others were injured on site. Outside the plant, 53 persons were reported injured and many houses and shops suffered damage.

Piper Alpha, North Sea, July 1988

At Piper Alpha's offshore module, a rather small gas explosion in a compressor module caused fires. The main part of the platform burned down killing 167 people. From the Piper Alpha incident, we can learn that a gas explosion can easily result in domino effects and loss of control. Installations should be designed to avoid domino effects. Figure 7.3 shows the aftermath of the explosion in the Piper Alpha offshore module plant at North Sea in the UK.



Fig. 7.2 Flixborough (UK) explosion, June 1974. (Paphoses.co.uk)



Fig. 7.3 Piper Alpha (UK) explosion, July 1988. (News.bbc.co.uk)

Buncefield Oil Storage Depot, England, December 2005
On Sunday 11 December 2005, a large explosion occurred at the Buncefield Oil Storage Depot (Figs. 7.4 and 7.5). The explosion destroyed a significant section of the depot and caused widespread damage to homes and businesses surrounding the site. Overfilling of one of the storage tanks caused a release of 300 tons of unleaded gasoline and created a very large vapour cloud (>120,000 m^2).



Fig. 7.4 Explosion ensuing fire at the 2005 Buncefield oil and gasoline terminal



Fig. 7.5 Pre-accident view of Buncefield oil storage depot

The ignition source of the cloud was likely an emergency pump house. The explosion resulted in very large overpressures (>2 bar), which caused significant damage to neighbouring commercial and residential structures.
Reviews of actual vapour cloud explosion incidents suggest that their occurrence is linked to:

1. Chemical and physical properties of the fuel. Historical evidence suggests that the probability of an explosion is higher with more reactive fuels such as propane and ethylene.
2. The way in which the fuel is stored.
3. The environment into which the release occurs. A common feature of vapour cloud explosions is that they have all involved ignitions of clouds which have engulfed sections of plant with a high degree of congestion.

Following the Piper Alpha disaster in July 1988, a group of concerned individual companies together with the UK and Norwegian regulators (UKHSE and NPD), set about organising a joint industry Project on Fire and Explosion for topside structures ("JIP Phase 1" [1]). The project ran from May 1990 to February 1991; the aim was to identify the best available knowledge worldwide and to identify significant gaps in that knowledge base leading to uncertainty in design, as well as to devise future programmes aimed at reducing the uncertainty level. The output of that project now comprises 26 reports and a "1991" guidance document [2].

7.3 Combustion and Flammability Characteristics of Gases

Fire and explosions are basically rapid chemical reactions or combustion. Good understanding of the basic chemical reactions and thermodynamics principles is essential for the assessments of the fire or explosion processes. This understanding will help improving the safety issues related to the damaging effects resulting from a fire and/or explosion incidents. When an incident involves the release of flammable gas, the gas reacts with the oxygen in air. Following ignition, heat and light are given off and products of combustion are generated.

7.3.1 Combustion Chemistry

The following example shows the oxidation of methane with oxygen in air:

$$(\textbf{Reactants})1CH_4 + 2O_2 \rightarrow (\textbf{Products})1CO_2 + 2H_2O + \text{heat of combustion} \tag{7.1}$$

In this chemical reaction processes, one molecule of methane is completely oxidised by two oxygen molecules and heat of combustion is generated as a result of the rearrangements of chemical structures from reactants to products. The resulting products are one molecule of carbon dioxide and two molecules of water vapour. It is important to note that the number and kind of atoms on the reactants side (left-hand side) are algebraically equal to the number and kind of atoms on the products side (right-hand side); in all sorts of chemical reactions, mass must be conserved, i.e. the total mass of materials before the reaction must be equal to total mass of the materials after the reaction. This can be checked as follows:

$$\text{Mass of materials “before”reaction } = (12 + 4 \times 11) + 2(2 \times 16) = 80$$
$$\text{Mass of materials “after”reaction } = (12 + 2 \times 16) + 2(2 + 16) = 80$$

When air is used instead of pure oxygen for the oxidation of methane, the chemical equation becomes:

$$1CH_4 + 2(O_2 + 3.76N_2) \rightarrow 1CO_2 + 2H_2O + 2(3.76)N_2 \tag{7.2}$$

The products of combustion are mixed with a large amount of inert nitrogen which heats up along with the rest of the combustion products.

The fuel-to-air ratio of the above reaction can be calculated from:

$$\frac{\text{Mass fuel}}{\text{Mass air}} = \frac{\sum_{\text{Fuels}} n_i M_i}{\sum_{\text{air}} n_i M_i} \tag{7.3}$$

where

n_i moles of species i
M_i molecular weight of species i

In this example,

$$\text{AFR} = \frac{1\ \text{mole}\,CH_4 \times 16\frac{g}{mol}CH_4}{\left(2\ \text{mole}\,O_2 \times 32\frac{g}{mol}\,O_2\right) + \left(3.76\ \text{mol}\,N_2 \times 28\frac{g}{mol}N_2\right)} = 0.058 \quad (7.4)$$

The air-to-fuel ratio is calculated from:

AFR = Air-to-fuel ratio = 1/ FAR = 17.2

A useful parameter often used in combustion chemistry is the fuel–air equivalence ratio, Φ;

$$\varphi = \frac{\text{AFR (mixture)}}{\text{AFR (stoichiometric)}} \quad (7.5)$$

where

$\varphi < 1$ excess air (lean mixture)
$\varphi = 1$ stoichiometric
$\varphi > 1$ excess fuel (rich mixture)

In stoichiometric combustion, all the fuel is burned and all products are fully oxidised. While controlled combustion in practical engineering devices such as boilers, gas turbine or homogenous-charged spark ignition engines may get very close to stoichiometric combustion, non-stoichiometric combustion usually occurs in fires. That is, where not enough air is available through natural convective effects to fully oxidise available fuel. In such cases, partial oxidisation of the fuel will occur, rich mixture combustion condition. The product of combustion will then be a combination of carbon dioxide, carbon monoxide, unburned fuel and oxides of nitrogen.

7.3.2 Flammability Characteristics

The main parameters that describe the ability of a flammable mixture to burn are its flammability limits, flashpoints and auto-ignition temperatures. The *flammable limits* (FL) specify the range of a concentration of a flammable mixture that will burn (or explode) if an ignition source is introduced. The limits are commonly called the "lower flammability limit" (LFL) and the "upper flammability limit" (UFL). Below the LFL, the mixture is too lean to burn, and above the UFL, the mixture is too rich to burn. Figure 7.6 shows a schematic drawing for the variation of explosion pressure with concentration of flammable mixture of fuel and air.

The flashpoint is the lowest temperature, at which a liquid gives off enough flammable vapour to ignite when mixed with air. Substances with a flashpoint below 55 °C are considered flammable. Figure 7.7 shows a schematic drawing for the flashpoint curve.

The auto-ignition temperature is the lowest temperature at which a flammable mixture will self-ignite. Table 7.1 shows values for flammability limits, flashpoints and auto-ignition temperatures for several common substances.

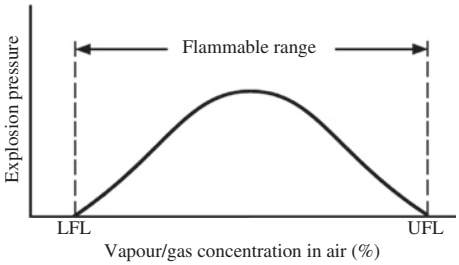


Fig. 7.6 Schematic drawing for the variation of explosion pressure with concentration of flammable mixture of fuel and air

Fig. 7.7 Flashpoint curves

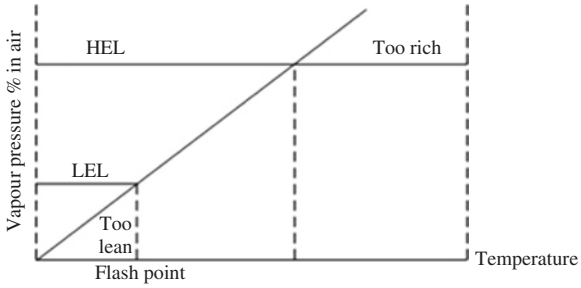


Table 7.1 Characteristic values for the LFL, UFL, flashpoint and auto-ignition temperature for common gases

Substance	Lower flammable limit %v/v in air	Upper flammable limit %v/v in air	Flashpoint °C	Auto-ignition temperature °C
Acetic acid	5	16	43	427
Acetone	3	13	−19	538
Acetylene	2	83	–	305
Benzene	1	8	−11	560
Butyl acetate	1	8	22	421
Butane	2	9	–	365
Butyl alcohol	1	11	29	345
Butyl methyl ketone	1	8	25	423
Carbon disulphide	1	44	−30	90
Cyclohexane	1	8	−18	260
Cyclohexanol	1	9	68	300
Cyclohexanone	1	9	44	420
o-Dichlorobenzene	2	9	65	648
1,1-Dichloroethane	6	11	14	458
1,2-Dichloroethane	6	16	13	413
Dichloromethane	16	66	–	600
Diethanolamine	2	13	12	662

(continued)

Table 7.1 (continued)

Substance	Lower flammable limit %v/v in air	Upper flammable limit %v/v in air	Flashpoint °C	Auto-ignition temperature °C
Diethylamine	2	3. 10	−26	312
Diethylether	2	48	−45	160
Diisobutyl ketone	1	6	49	396
Diisopropyl ether	1	21	−28	443
Dimethyl sulphoxide	3	42	88	215
1,4-Dioxane	2	22	12	180
Epichlorohydrin	4	21	31	416
Ethane	3	12	–	515
Ethanol	3	19	12	363
Ethyl acetate	2	12	−4	410
Ethylene glycol	3	22	111	400
2-Ethoxyethanol	3	18	43	235
2-Ethoxyethyl acetate	2	8	56	379
Ethylene	3	34	–	305
Furan	2	14	−36	–
Furfural	2	19	60	316
Glycerol	3	19	199	370
Heptane	1	7	−4	204
Hexane	1	8	−22	223
Hydrogen	4	76	–	–
Isobutyl alcohol	2	11	28	426
Isophorone	1	4	84	460
Isopropyl alcohol	2	12	12	399
Methanol	6	37	11	385
Methyl acetate	3	16	−10	455
Methyl ethyl ketone	2	11	−6	516
Nitrobenzene	2	9	88	482
Octane	1	7	13	206
Pentane	1	8	−40	285
Propane	3	12	–	515
Propyl acetate	2	8	13	450
Propylene	2	11	−108	460
Pyridine	2	12	20	482
Styrene	1	6	31	490
Tetrahydrofuran	2	12	−14	321
Toluene	1	7	4	530
1,1,1-Trichloroethane	8	11	−1	537
Trichloroethylene	13	90	32	420
o-Xylene	1	6	17	464
m-Xylene	1	7	25	528
p-Xylene	1	7	25	529

The limits indicated are for gas and air at 20 °C and atmospheric pressure. (Data extracted from Gas Data Book, 7th edition, copyright 2001 by Matheson Gas Products)

LFL can be estimated from the flashpoint as:

$$\text{LFL} = \text{vapour pressure at flash point}/(7500\ \text{mm Hg})$$

For mixtures, the following empirically derived Le Chatelier's (H. Le Chatelier, Ann. Mines, vol. 19, ser 8) equations may be used

$$\text{LFL}_{\text{mix}} = \frac{1}{\sum_{\text{i}=1}^{n} \frac{y_i}{\text{LFL}_i}} \tag{7.6}$$

$$\text{UFL}_{\text{mix}} = \frac{1}{\sum_{\text{i}=1}^{n} \frac{y_i}{\text{UFL}_i}} \tag{7.7}$$

where

LFL_i, UFL_i	Lower and upper flammability limits for component i in the mixture
y_i	the mole fraction of i
n	the number of combustible species

7.3.3 Flame Speed and Burning Velocity

In order for fire or explosion to take place, two conditions must be present; sufficient quantities of flammable fuel and air and an ignition source. These two components are necessary to initiate fire or explosion. Once the flammable fuel or gas is ignited, a flame front is generated, which separates the burned from unburned mixtures. Chemical reactions will take place at the flame front, and the heat released during combustion causes the flame to propagate and the gas to expand and pushed away from the ignition source due to convection effects.

The definition of the burning velocity is the speed at which the flame front propagates through the flammable mixture relative to the unburned mixture ahead of the flame. It should be noted that the burning velocity does not take in account the flame expansion relative to a stationary object, such as a wall. When such expansion is taken into account, the term "flame speed" is used. Flame speed is the velocity by which the flame travels relative to a stationary object. The flame speed associated with deflagration is much higher than the burning velocity.

In general, the flame speed is related to the burning velocity as follows:

$$S_f = Su + u_\text{g} \tag{7.8}$$

where

S_f	flame speed
S_u	burning velocity
u_g	gas velocity relative to a stationary object

Rates of flame propagation are characterised by the burning velocity (Su), i.e. the velocity of a plane flame normal to itself and relative to the unburned reactants. Tabulated burning velocities are those measured under laminar or static flow conditions and represent the lowest possible intrinsic flame propagation rates.

Examples:

- 0.43 m/s for methane
- 0.80 m/s for ethylene
- 3.50 m/s for hydrogen

In general, the rate of generation of mass of products can be calculated from:

$$\frac{\mathrm{dm}_b}{\mathrm{dt}} = S_u A_f \rho_u \tag{7.9}$$

This enables the calculation of the flame spread in any system with product confinement, at constant pressure.

Turbulence may enhance burning velocities well above the laminar value. It is usually present in explosions and may be generated by the flow of unburned gas ahead of the flame over obstacles within an enclosure. Turbulent burning velocity, S_t, should exceed the laminar value by the intensity of turbulence. A better correlation of experimental data, which also allows for the burning velocity tending to zero at the limits of flammability, gives:

$$\frac{S_t}{S_u} = 1 + \left[\frac{(T_b - T_{b,I})}{2T_u}\right] \cdot \frac{u'}{S_u} \tag{7.10}$$

where

U′ the turbulence intensity
T_b the final flame temperature
$T_{b,L}$ final flame temperature at the limit of flammability

The increased transport of heat and mass in the flame front causes increased burning rates. Quantifying the effects of turbulence is difficult, and some subjectivity is unavoidable. Two factors in addition to turbulence Reynolds number, Re, have to be introduced to fully describe flow behaviour—intensity and scale of turbulence. Small-scale turbulence increases local heat and mass transfer rates to give higher Su and $P_{\max}$. Large-scale turbulence generated by obstructions can increase in flame area, thus higher $P_{\max}$. Turbulence may typically enhance burning velocity by a factor up to 10.

Flame speeds can become high, particularly in the tube situation where displacement of the gas ahead of the flame creates pipe-flow turbulence. Flame speed without turbulence for hydrogen would be about 24 m/s but turbulence could lead to “normal” flame speeds of 240 m/s or more even at this flame speed; however, deflagrations are still “normal” flames. It is only when the flame speed exceeds the velocity of sound in the unburned gas that the flame becomes “abnormal” (See the following section). Detonation is an abnormal flame process which occurs at

many times the speed of sound in the unburned gas. For hydrocarbon–air mixtures, the detonation velocity is 2,000–3,000 m/s (NB velocity of sound in air is about = 330 m/s).

It can be shown that in the general case,

$$S_f = \left(\frac{\rho_u}{\rho_b}\right) S_u$$

i.e.

$$S_f = E.S_u \tag{7.11}$$

where E = expansion factor.

The expansion factor, E is a measure of the increase in volume produced by combustion and may be expressed as

$$E = \left(\frac{T_f}{T_i}\right)\left(\frac{N_b}{N_u}\right) \tag{7.12}$$

where:

Nu and Nb: the number of moles before and after combustion.

For most hydrocarbons mixture with air, $E \approx 8$ [3], main combustion properties are given in Tables 7.2 and 7.3.

Table 7.2 Main combustion properties of hydrogen and hydrocarbon mixtures with air [3]

Fuel	Molecular weight	A flammability %v/v gas		% gas at stoichiometric ratio	B	
		LFL	UFL		% gas at maximum burning velocity	Maximum laminar burning velocity Su (m/s)
Hydrogen H_2	2	4	75	30	54	3.5
Methane CH_4	16	5	15	9.5	10	0.45
Ethane C_2H_6	30	3	12.5	5.6	6.3	0.63
Propane C_3H_8	44	2.2	9.5	4.0	4.5	0.52
Butane C_4H_{10}	58	1.9	8.5	3.1	3.5	0.50
Pentane C_8H_{12}	72	1.5	7.8	2.6	2.9	0.52
Hexane C_6H_{14}	86	1.2	7.5	2.2	2.5	0.52
Heptane C_7H_{16}	100	1.2	6.7	1.9	2.3	0.52
Acetylene C_2H_2	26	2.5	80	7.7	9.3	1.58
Ethylene C_2H_4	28	3.1	32	6.5	7.4	0.83
Propylene	42	2.4	10.3	4.4	5.0	0.66
Butylene	56	1.7	9.5	3.4	3.9	0.57
Benzene C_6H_6	78	1.4	7.1	2.7	3.3	0.62
Cyclohexane C_6H_{12}	84	1.3	8.0	2.3	2.7	0.52

Table 7.3 Properties of hydrocarbons

Fuel	Net calorific value (W/m^2) (288 K, 1atm)	Minimum ignition energy	Auto-ignition temperature (K)	Maximum laminar flame speed (m/s)	Expansion ratio T_f/T_i $T_i = 288$ K	Adiabatic flame temperature (K)
Hydrogen	10.2	0.02	847	28.0	8.0	2,318
Methane	34	0.29	813	3.5	7.4	2,148
Ethane	60.5	0.24	788	4.0	7.5	2,168
Propane	86.4	0.25		4.0	7.6	2,198
Butane	111.4	0.25	678	3.7	7.5	2,168
Pentane	138.1	0.25	533	4.0	7.7	2,232
Hexane	164.4	0.25	498	4.0	7.7	2,221
Heptane	190.4	0.25	488	4.0	7.6	2,196
Acetylene	51	0.02	578	14.2	9.0	2,598
Ethylene	56	0.12	763	6.5	7.8	2,248
Propylene	81.5	0.28	733	5.1	7.7	2,208
Butylene	107.1	0.28	658	4.3	7.6	2,203
Benzene	134.0	0.22	833	4.9	7.9	2,287
Cyclohexane	167.3	0.24	518	4.1	7.8	2,232

7.4 Deflagration and Detonation

Following ignition, explosions from combustion of flammable gas can produce two different types of flames:

i. Deflagration (flame propagation)
ii. Detonation

The difference depends on the speed of the shock wave generated from the combustion process. In the deflagration mode, the most common mode, the flammable mixture burns relatively slowly and the generated flame propagates at subsonic speed relative to the unburned gas. For hydrocarbon–air mixtures, the deflagration velocity is typically far below 100 m/s and relatively modest overpressure. The main mechanism of flame propagation is of a flame front that moves forward through the gas mixture, i.e the reaction zone (flame front) progresses through the medium by processes of diffusion of heat and mass. The flame propagation speed and explosion-generated pressure will strongly depend on the gas cloud and the geometrical conditions within the cloud (i.e. process equipment, piping, etc.) or geometries confining the cloud (i.e. buildings etc.). A deflagrating flame may turn into detonation (deflagration to detonation transition, DDT), particularly when travelling down a long pipe but unlikely when travelling in vessels or open spaces. Acceleration to detonation appears to be unlikely or impossible in three-dimensional systems unless powerful initiators, e.g. explosive detonators, have been used (Fig. 7.8).

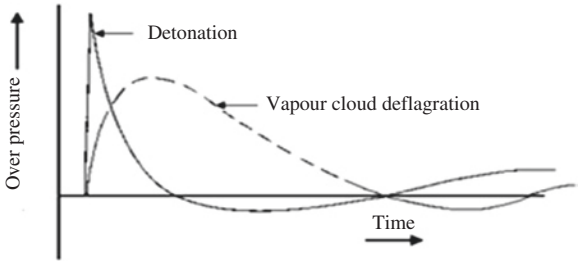


Fig. 7.8 Characteristics of deflagrating and detonating pressures

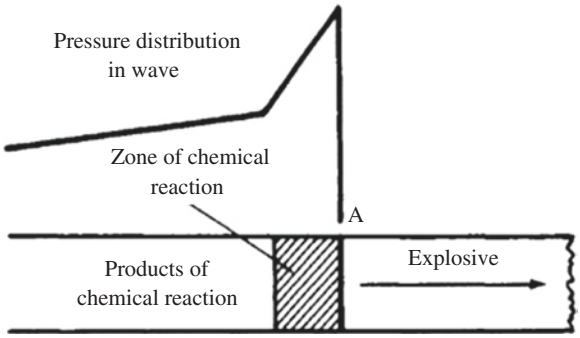


Fig. 7.9 Schematic diagram of a detonation wave: (*A*) shock wave front. The arrow shows the direction of propagation of the wave

In the detonation mode, the flame front travels as a shock wave followed closely by a combustion wave or flame, which releases the energy to sustain the shock wave. The detonation front reaches and exceeds the velocity of sound (330 m/s) typically of the order of 2,000–3,000 m/s. A detonation generates greater pressures and is more destructive than a deflagration explosion flame. Whereas the peak pressure caused by the deflagration of a hydrocarbon–air mixture in a closed vessel is of the order of 8 bar, a detonation may give a peak pressure of the order of 20 bar (Fig. 7.9).

7.5 Confined and Vapour Cloud Gas

7.5.1 Introduction

A basic distinction may be drawn between explosions which occur in some sort of containment (confined explosion) and those which occur in unconfined vapour or gas clouds (unconfined vapour cloud explosion). In general, the term "unconfined" is a misnomer. It is more accurate to call this type of explosion simply a "vapour cloud explosion (VCE)".

A confined explosion occurs in a confined space, such as a vessel or a building. On the other hand, VCE occurs in the open. This type of explosion is usually the result of a flammable gas spill. The gas is dispersed and mixed with until it

comes in contact with an ignition source. VCE are rarer than confined explosions since the explosive material is frequently diluted below the LFL by wind dispersion. These explosions are very destructive since large quantities of gas and large areas are frequently involved.

7.5.2 Confined Gas Explosions

The two most common confined explosion scenarios involve explosive vapours and explosive dusts. The nature of an explosion is a function of several characteristics that can be determined experimentally. These characteristics are dependent on the explosive material used and include flammability or explosion limits, maximum rate of pressure rise and maximum explosion pressure after ignition. The limits of flammability or explosively are used to determine the safe concentrations for operation or the quantity of inert required to control the concentration within safe regions. In general, explosions that occur in buildings or plant are initially confined but due to failure of part or whole of the structure become at an early stage vented explosions.

The level of generated explosion pressure depends on the rate at which the energy is released.

Two basic types of energy released in an explosion:

i. Physical energy: includes pressure energy in gases, strain energy in metals, etc.
ii. Chemical energy: derives from a chemical reaction usually combustion.

A completely confined gas explosion can generate a pressure higher than 8 bar. This is much greater than most structures can withstand, and normally a weaker section will fail and allow the gases to be vented at a much lower pressure. Once venting begins, the rate of pressure rise will depend on the balance between the rate of increase of volume due to combustion and the rate at which the unburned and/or burned gases are vented. If the latter is small, the pressure will continue to rise but at a lower rate. However, internal gas explosions rarely cause complete destruction because either by design or fortuitously the pressure is relieved at an early stage of the explosion by failure of weak components. Such explosions are termed vented confined explosions.

The maximum rate of pressure rise is indicative of the flame front propagation rate and thus the magnitude of the explosion. Thus, the explosive behaviour of different materials can be compared on a relative basis. It is also used in to design of relief system (rupture discs, explosion door, etc) for a vessel during an explosion before the pressure ruptures the vessel, or to establish the time interval for adding an explosion suppressant (water, carbon dioxide, or Halon) to stop the combustion process.

The maximum pressure generated in a totally confined explosion in which the mixture completely fill the enclosure depends on the energy content (calorific value, CV) and is independent of the volume of the enclosures. Although calorific

values differ widely between flammable gases, the energy content ("energy density") of most stoichiometric hydrocarbon mixtures is nearly the same.

Thus, all stoichiometric hydrocarbon–air mixtures ignited at atmospheric pressure under complete confinement will generate close to the same maximum pressure. Assuming no molar change, this pressure is given by:-

$$P_{\max} = P_o(T_{\text{final}}/T_u) \tag{7.13}$$

where :

$P_{\max}$ the maximum-generated pressure
P_o atmospheric pressure
T_{final} final temperature
T_u temperature of the unburned mixture

Neglecting heat losses, the final temperature T_{final} is higher than the mixture adiabatic flame temperature tabulated earlier due to additional compressive heating of burned and unburned gases.

Although the final maximum-generated pressure, $P_{\max}$, is about the same for all gases, the time to reach it will depend on the rate of flame propagation and therefore the reaction rate. The figure below shows an example of pressure time histories of different gases (Fig. 7.10).

The maximum explosion pressure, $P_{\max}$, is used in the pressure-resistant design. The design pressure must be higher than the anticipated explosion pressure. If economics dictates that pressure-resistant design is not warranted, relief systems will be required to drastically reducing the pressure loads.

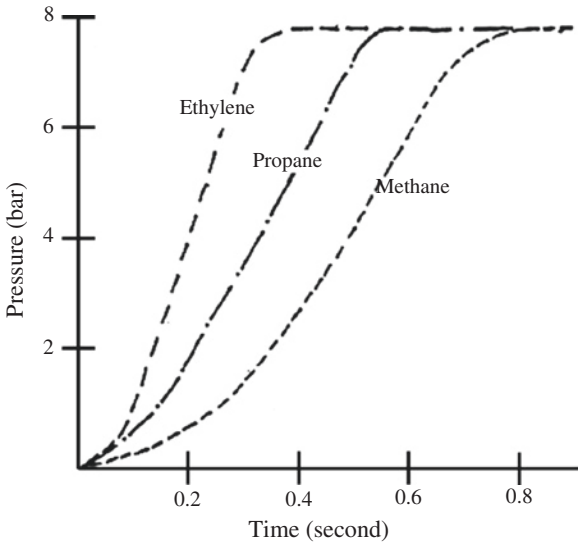


Fig. 7.10 Schematic pressure time histories for different stoichiometric gas–air mixture in a 1 m^3 vessel [19]

Throughout totally confined explosion, the pressure generated always acts uniformly on the enclosure because pressure equalises at the speed of sound (2 orders higher than flame speed). In spherical vessels starting with an initially quiescent gas mixture, and centrally ignited, it has been reported [4] that the maximum rate of pressure rise is inversely proportional to the cube root of the volume. The constant of proportionality is termed the K_G value and has the units of bar m/s:

$$K_G = (dP/dt)_{max} V^{1/3} \tag{7.14}$$

where:

K_G is a flame deflagration index
V the vessel volume

This relationship is called the "Cubic Law" and has been used to scale data from laboratory-sized vessels to plant-scale equipment. As the strength of an explosion increases, the deflagration index increases. The table below gives values of K_G for some gases (Table 7.4).

Pressure development in a vented explosion depends on:

- Mixture distribution
- Enclosure Geometry
- Point of ignition

Pressure developed in a vented explosion when mixture does not fill enclosure can be as high as when it does despite the fact that energy release is less. Because maximum pressure attainable often depends on maximum flame area achieved, maximum pressure normally occurs when flame reaches its maximum surface area. Non-central ignition reduces maximum flame area possible which may affect pressure/time profile through effect on volume of unburned mixture expelled prior to burnt gas venting. Location of the point of ignition relative to the relief vent is also important. The closer the relief vent to the point of ignition, the lower the maximum-generated pressure. Higher pressure is possible if two ignition sources present. This is due to flame kernels coalesce and increase flame area.

Figure 7.11 shows a pressure–time history in vented explosions. Following ignition, early flame propagation (stage A) is associated with a low-pressure gradient until the vent opens at P_1 before flame reaches it. This is followed with a faster flame acceleration (stage C) until maximum pressure P_2 is reached. Decay in pressure will follow (stage D) during flame venting from the vessel.

Table 7.4 Average K_G values for selected gases [4]. *NB. Values at ambient conditions and ignition energy of 10* J

Gas	K_G (bar m/s)
Methane	55
Propane	75
Hydrogen	550

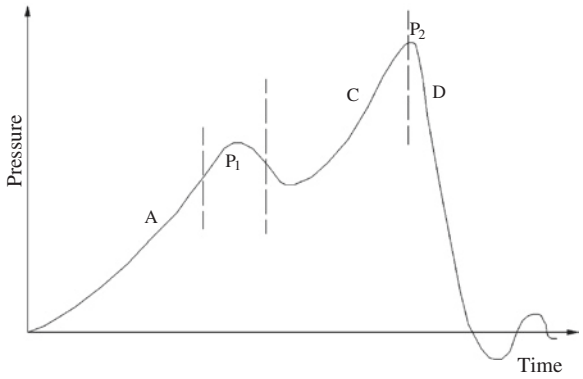


Fig. 7.11 A pressure–time history in vented explosion

7.5.3 Vapour Cloud Explosions

A VCE is the result of a release of flammable material in the atmosphere, a subsequent dispersion phase, and, after some delay, an ignition of the vapour cloud. A flame must propagate at a considerable speed to generate blast, especially for 2D (double-plane configurations) and 3D (dense obstacle) environments. In order to reach these speeds, either the flame has to accelerate or the cloud has to be ignited very strongly, thereby producing direct initiation of a detonation.

Flame acceleration is only possible in the presence of:

- outdoor obstacles, for example congestion due to pipe racks,
- weather canopies, tanks, process columns and multi-level process structures;
- high momentum release causing turbulence, for example an explosively
- dispersed cloud or jet release;

Several factors influence the strength of VCE if the cloud ignites:

- degree of confinements, turbulence and cloud composition
- strength of ignition source
- cloud configuration
- reactivity of the mixture.

The major blast source of VCE is the presence of obstruction or confinement. Reactivity of the mixture has only a secondary effect on the explosion strength. Turbulence increases the strength of the blast. It can be generated by the type of release prior ignition or by the flame propagation itself in the cloud especially when many obstructions are present down the path of flame propagation.

Large-scale release of fuel vapour followed by ignition, usually result in fire balls which produce much radiation, due to the high emissivity of pyrolysing fuel, but no blast. However, premixing with air occurs because of fast jet entrains, VCE can occur. VCE can cause devastating damage and have two unexpected features:

i. They correspond to flame speeds in excess of 100 m/s, without involving detonation.
ii. The damage is often caused at sites remote from the location of the explosion where the devastation may be much less.

VCEs have increased in number due to an increase in the use of flammable materials in chemical processing plants and operation in more severe conditions. Some of the worst disasters in the history of the process industries have been VCEs. Some examples include:

- June 1974, Flixborough, England (28 fatalities)
- October 1989, Houston, Texas, USA (23 fatalities)
- March 2005, Texas City, Texas, USA (15 fatalities)
- December 2005, Buncefield, England (no fatalities but 43 injuries and major damage)
- October 2009, Jaipur, India (12 fatalities)

Lees [5] has shown that:

i. Vapour cloud fires are more common than explosion
ii. The ignition probability increases as the size of the vapour cloud increases
iii. Ignition of the cloud at a point remote from the release increases the severity of the explosion
iv. The explosion efficiency, defined as the percentage of the combustion energy that is converted into a blast wave, is usually small—about 2 %.

It is difficult to cope with a potential VCE once an accidental release has occurred. Consequently, the best procedure to guard against a VCE is to prevent the release in the first place. Safe piping is essential to protect against VCEs. Forty per cent of all major plant losses are due to piping failures, and corrosion is one of the largest single causes of plant and equipment breakdown. Moreover, mistakenly open valves that caused mammoth emissions of hydrocarbons have resulted in two major VCEs with a total of 29 deaths in those two instances. Thus, close scrutiny regarding piping and valves is mandatory to help prevent VCEs. Commonly used methods to prevent VCE include:

- Use leak detector for leaks at low concentration
- Keep low amount of volatile, flammable materials
- Install automated block valves to shut systems down at the early stages of a spill.

7.6 Explosion Blast Loading on Structure

In VCE, the generating deflagrating flame produces a pressure/blast waves causing loading on structures which lies in its path causing destructive effects. The pressure waves move at subsonic speed and retain its destructive force longer as it moves away from the ignition source. This force increases as the distances from

Table 7.5 Provides typical blast effects

Pressure (psi)	Damage
0.15	Typical pressure for glass
0.4	Limited minor structural damage
0.5–1	Breakage of window glass
0.7	Minor damage to house structure
1–2	Corrugated asbestos shattered
3	Near total destruction to light industrial buildings of rigid steel framing; corrugated steel structures less severely damaged
3–4	Frameless, self-framing steel panel buildings demolished
5–7	Nearly complete destruction of houses
7	Loaded railroad cars overturned
7–8	Brick panels, 8–12 inches thick, not reinforced fail by shearing or flexure
9	Complete destruction of railroad boxcars
10	Probable total destruction of buildings

the ignition force increases. This is due to an increase in the flame front area due to volume expansion. Therefore, the area of most destruction will be at some distance away from the ignition source. The turbulence generated as the flame moves around solid obstructions, such as vessels, pipes, buildings, etc, can dramatically increases the surface area of the deflagrating flame and therefore increases its destructive damage due to higher generated peak pressure (Table 7.5).

Different pressure/time profiles of gas explosions result in different levels and types of damage. Therefore, the structure response to an explosion blast is time-dependent.

Structure response depends on relative ratio between the duration of the imposed load, Td, and the natural period of vibration, Tn, of the structure. Three basic types of response have been identified by Harris [19]:

i. Td > Tn: here the loading experienced will effectively be equivalent to a static load equal to the peak explosion overpressure.
ii. Td ≈ Tn: here the loading experienced will effectively be equivalent to a static loading of a magnitude greater than the peak overpressure. The equivalent static overpressure can be up to 1T 12 times the incident overpressure
iii. Td < Tn: here the pressure is effectively partially absorbed and the loading experienced will be equivalent to a static loading lower than the explosion peak overpressure, i.e. a structure can withstand a higher dynamic pressure than the static load necessary to cause failure.

Thus, gas explosions can be considered approximately as static loadings equal to the peak overpressure.

In detonation, generated pressure waves move at supersonic speeds but dissipate very quickly which make its destructive effects localised on nearby objects. At large distance, low-energy shock will be present. The blast produced by detonation usually pulverise nearby surfaces, creating a crater whose size reflect the size of and magnitude of the detonation blast.

7.7 Mitigation

7.7.1 Mitigation by Design

Mitigation by design is one of the important and cost-effective ways of reducing explosion damaging overpressure [6] gave a concise account of the best advice that current platform ventilation and explosion understanding could then give. This work has not been overtaken by subsequent research although much more experience has now been gained. A "classic" case of mitigation by design optimisation has been illustrated [7]. In the design of the BP ETAP platform against explosion, they showed how in one instance a layout with pig launchers in a north–south direction gave a computed (CFD) overpressure of over 4 bar, with the equipment in east–west direction maximum pressures of 0.5 bar were attained. The paper appears to be a good illustration of how design optimisation for explosion at an early stage results in low or no cost optimisation and can also bring design simplification. The authors also correctly draw attention to the pitfalls of failing to take into account vendors skids, cable racks, etc. Even explosion mitigation by design optimisation was not universal practice at that time work on water spray methods had been used in similar geometries.

7.7.2 Mitigation by Water Spray

The optimisation process for system design with water spray mitigation comprises getting a water spray of sufficient loading density in the right place with a droplet size amenable to forming micro-droplets by the stripping process. While many studies and most implementations have concentrated on the use of area deluge methods as implemented for fire protection, the delivery of water drops already in the fine mist format necessary for flame reduction is an obvious alternative. Over the years, a number of fine mist devices have been developed and tested. One device uses stored thermal energy in superheated water at 180 °C, svp 10 bar. On discharge, immediate flashing occurs and many droplets formed are in the 10 micron range with a tail of non-flashed material as large as 100 μ. One of these devices, Micromist [1, 8] has been tested in a medium-scaled replica of an offshore module. Tam et al. [8] used 4 devices, two on the roof discharging downwards and two discharging horizontally in a range of directions. Different ignition to discharge times (50, 100, 150 ms) were used and both central and end ignition were tried. The tests demonstrated that such devices could arrest a propagating flame and totally suppressed an explosion within a confined module. The results showed that the performance of the mist curtain lies somewhere between that of a diluent and a perfect suppressant. If the mist curtain was deployed over only the upper half of the module, then the effect was negative. There remain a number of questions to be answered, especially those concerned with the number

and position of devices to be deployed, reliability of device and trigger system, maintenance required, etc. Nevertheless, the idea of a device with localised intelligence is appealing [20–23].

7.8 Mathematical Modelling of Explosions

Following the Piper Alpha disaster and the subsequent Cullen enquiry, offshore operators now have to define the risks associated with their operations and convince the HSE that these risks are as low as reasonably practicable (ALARP). This approach embodies a vast array of design methodologies for structures, process engineering, construction, human factors and safety. Many of these procedures are well documented in standards which have been tested over time. However, when considering safety issues, it is necessary to ascribe probabilities to failures and events which may lead to a loss of flammable material and subsequently may lead to a fire or an explosion. The probability of an explosion has major implications on the safety of personnel on a platform both in terms of potential loss of life and the possibility of escalation of the process which could lead to a domino effect and more serious consequences. An understanding of the possible severity of an explosion is therefore a critical factor in establishing the satisfactory demonstration of ALARP. The mechanisms that enhance explosion overpressures, therefore, need to be established with some certainty in order to ensure that all aspects of safe design (structures and processes), and the safe protection of personnel are taken into account. Builders and operators of process plant require design criteria to ensure that the risks associated with an explosion from an accidental release of flammable material satisfy ALARP.

In recent years, considerable effort has been devoted to the development of reliable means of assessing the vulnerability of industrial plant to fire and explosion. Examples include the Commission of European Communities (CEC) co-sponsored project MERGE—Modelling and Experimental Research into Gas Explosions and the Steel Construction Institute's Joint Industry Project (JIP) on blast and fire engineering. Current explosion models are either empirical or based on phenomenological computational fluid dynamics (CFD) models and are very often tuned to particular experiments which thereby limit their range of applicability. Moreover, the effect of obstacles in the generation of turbulence and its effect on the rate of chemical reaction are inadequately treated. Since explosions are characterised by the interaction between a compressible fluid flow and rapid chemical reactions, any realistic predictive computer model must be able to deal with both of these aspects of the problem.

There is a wide range of explosion models that are frequently used in support of facility layout and risk. They range from simple empirical to phenomenological to complete CFD models. Moving from one to another means more physical representation can be simulated. The common practice is to use a model to generate solution to a defined explosion incident. Once the model is constructed, the produced results are validated against a series of experimental tests, to examine how well the model recreates the actual explosion phenomena.

Results that are often used for model validation are the profile and magnitude of the generated explosion pressure. These data are used to assess the damage that might be expected at various locations. For VCE, the gaseous fuel and its properties and concentration must be identified together with the areas of congestion and confinements.

In the following sections, brief overview of models that are used in explosion modelling is presented and discussed together with their shortcoming.

7.8.1 Empirical Models

Empirical models are related to experimental data. They represent little or no physics. Among these models are the TNT equivalency and multi-energy models.

7.8.1.1 The TNT Model

The trinitrotoluene (TNT) equivalency model relates the explosion to that equivalent of TNT. There are some essential differences between gas explosion and TNT. Among those are that the local pressure of the gas explosion is much less than that of TNT. The pressure decay from TNT is much more rapid than that of the gas explosion. The TNT equivalency model uses pressure–distance curves to yield the peak pressure. The model was extended [3] to take geometry effects in account.

The model calculates the TNT equivalent mass W_{TNT}, which is the mass of TNT that would produce the same effects as the amount involved in the explosion. The TNT equivalent mass describes the effect of the explosive at a certain space when it blasts and can be calculated from the explosive amount Wexp and the equivalency/yield factor f (kg TNT/kg explosive substance):

$$W_{TNT} = f \cdot Wexp \tag{7.15}$$

The yield factor, *f*, comes from experience and not universal. It also does not suit gas explosions since the behaviour is different than solid explosions. More specifically, it does not work for weak gas explosions. It must be taken into account that TNT equivalences often vary with distance. To determine the distance at which peak pressure occurs, the following formulae is used:

$$R = Z(W_{TNT})^{1/3} \tag{7.16}$$

where:

R the distance to peak overpressure,
Z the scaled distance (from Fig. 7.12)
W_{TNT} the equivalent weight of TNT.

It should be noted that the TNT method does not take into account the vapour cloud size or the obstacles and congestion in the region of the vapour cloud.

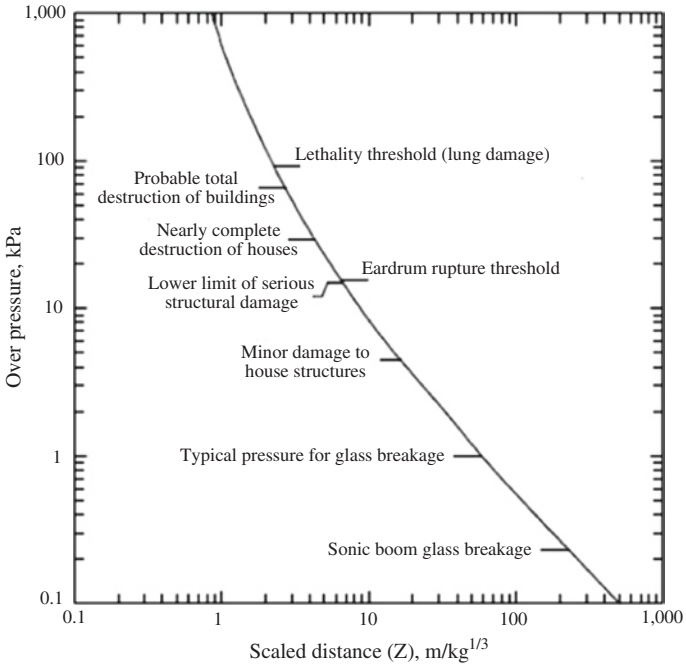


Fig. 7.12 Peak side–on overpressure versus scaled distance [11]

Nevertheless, the model has been used extensively to predict peak pressures from gas explosions due to its relative simplicity and the vast amount of data that exists from experiments.

7.8.1.2 The Multi-energy Model (MEM)

Berg [9] proposed another model based on multi-energy concept, which assumes that only the volume of gas that is within a confined or obstructed area contributes to the overpressure generated in a gas explosion. The fuel in the unconfined regions is treated as a fireball with small-generated overpressure. The model appears inaccurate when modelling a weak explosion as only small parts are confined. It also cannot model complex geometry properly. Its performance is unclear when many confined parts exist.

A hemispherical flame shape is assumed at the point of ignition with a constant flame speed. It is also assumes that the mixture within the confined region to be homogeneous and stoichiometric. The generated overpressure is calculated as a function of the combustion energy and a scaled distance from the ignition point.

The following equations are used to determine the overpressure:

$$\bar{R} = R(P_O/E)^{1/3}, \bar{P}_S = P_S/P_O \tag{7.17}$$

where:

$\bar{R}$ combustion energy-scaled distance,
R distance from the ignition point (m)
Po ambient pressure (pa),
E the combustion energy per volume times Vcloud, where Vcloud is the volume of vapour cloud in the congested area
$\bar{P}_S$ dimensionless "side-on" peak overpressure
P_S the "side-on" peak overpressure.

Considering that the combustion energy vary little with different hydrocarbon–air mixture, the combustion energy can be estimated from

$$E \approx 3.5\, V\text{cloud, [MJ]} \tag{7.18}$$

where Vcloud is measured in m^3.

The strength of the explosion can be estimated by taking into account the lay-out of the explosion source. The charge strength is given a number between one and 10, where 10 represent a detonation. Figure 7.13 shows the variation of P_S with *R, at different* blast strength values from 1 to 10.

A development of the MEM is the Baker–Strehlow methodology [10] and extended [11]. The model uses essentially the same method, however instead of specifying an explosion strength, the user can specify a flame speed, and the curves are represented as a function of the flame speed. The model takes into account several parameters such as dimension of the confined areas and potential

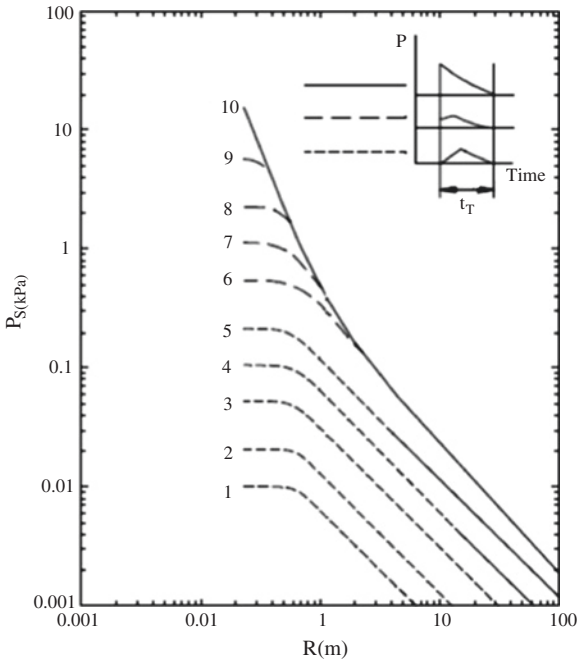


Fig. 7.13 Dimensionless overpressure versus energy-scaled distance [9]

explosion locations. It then calculates flame speed and fuel reactivity. The model is attractive since it takes into account geometry and hazardous areas and because of being easy to perform. However, it can be overconservative.

Since the models are based on experimental finding rather than on real physics, they are limited in their applicability and give only little description of the flow field.

7.8.2 Phenomenological Models

Phenomenological models are regarded as simplified physical models. They represent the main physics of the explosion. The models simplify the geometry and replace it by a vented chamber that contains several turbulence-generating grids. Simplifying geometry leads to uncertainty in the model setting as well as the simulated results. These models fall between empirical models and CFD models in terms of complexity.

Among codes that implement phenomenological model is the SCOPE [12] which is developed further [13]. The model is one-dimensional and is based on the idealised geometry of a vented vessel containing a series of obstacle grids. The code can handle vented and external explosions. However, unlike CFD-based models, it cannot describe details of the flow field. Puttock et al. [13] developed model takes into account the effects of obstacle complexity—phenomena not previously realised in earlier versions. At the same time, the authors have taken into account a revised turbulent burning velocity formulation, allowed for vents at rear walls, treatment of gas mixtures and of temperatures and humidity variations. After validation against over 300 experiments, the prediction of the data sets and the predictability in blind tests are not significantly inferior to the much more complex CFD calculations—but this type of model is so simple to run that batch running permits large numbers of combinations of ignition position geometry, etc—to be assessed.

Based on the applications to vented vessels explosions, Fairweather and Vasey [14] and Chippett [15] developed phenomenological models. The models see the confined explosions in buildings as several combustion chambers linked together. This is the case of a typical process plan congested with piping networks. The model parameters are determined from the obstacles-flow interaction. A combustion model is based on the local flow properties to determine the laminar and turbulent burning velocities. In this model, it is assumed that flow conservation equation apply on each chamber as a whole. Thus, the flame shape is predicted empirically and the ignition point can be anywhere within the chamber. The turbulent burning velocity is based on the laminar flamelet concept [16] and [17] which has been calibrated against measurements made [18]. This model can handle external explosions and can generate physics description from the geometry.

Phenomenological models, however, do not provide the same wealth of information about the flow field as CFD models as they contain less geometrical details than CFD models. Moreover, they can only deal with single enclosures.

7.8.3 Computational Fluid Dynamics Models

CFD models solve the partial differential equations, Navier–Stokes equations governing the fluid flow in the explosion process over thousands or millions of computational cells. The numerical solutions are generated by discretising the solution domain (in both space and time). The conservation equations are applied to each of the sub-domains formed by the discretisation process, generating a number of coupled algebraic equations that are normally solved by an iterative procedure.

Most models use a porosity distributed resistance (PDR) to model small-scale objects. A gas dispersion model is used to simulate the gas flow through the computational domain in and around the structures that exist. This then produces a cloud that is varied in concentration. In order to accurately model an explosion, the gas dispersion model must be able to simulate the release and flow of gas taking into account the obstacles, structures and the ambient conditions. After the cloud has dispersed, an ignition source is introduced. Turbulence is modelled using a turbulence model while the combustion chemistry is modelled using a reaction rate model.

Solutions obtained with CFD codes contain detailed information about the flow field, i.e. velocities, pressure, density, species concentrations, etc. Surface pressure data can be used for structural analysis. The data results from a CFD model are usually validated against well-controlled, well-defined and repeatable experiments in order to gain confidence in the results obtained, CFD models are powerful and useful tools. However, accuracy of the results will depend on the correct implementation of the geometry, initial and boundary conditions, turbulence and combustion sub-models.

Examples of CFD-based codes for explosion modelling include:

- EXSIM: a purpose-built explosion code developed by the Telemark Technological R&D Centre (Tel-Tek) in Norway and Shell Global Solutions in UK.
- FLACS (**FL**ame **AC**celeration **S**imulator) code has been developed at the Christian Michelsen Research Institute in Norway, now CMR-GEXCON
- AutoReaGas: developed as a joint venture, between Century Dynamics Ltd. and TNO.
- CFX: a general purpose, commercially available CFD code, developed by AEA
  – Technology Engineering Software at Harwell.
- REACFLOW: a CFD code developed at the Joint Research Centre of
  – the European Union in Ispra, Italy [24].

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Chapter 8
Business and Project Management of Natural Gas

8.1 Business and Project Management

8.1.1 Introduction

No matter where the gas is supplied from and where its destination is, the entire process, within the context of the natural gas business, should be managed in a coherent manner. This chapter thus starts by giving an overview of natural gas and its place as a viable source of future energy as well as providing information on the business elements of it. Moreover, the related legal framework of gas contract intends to give a better insight and knowledge when it comes to contract negotiation. After all we have to sell the gas once it has been produced. The chapter continues to give information on project evaluation, investment appraisal, teamworking, economic efficiency, gas chain and pricing of the gas [1].

The basics of project management can be applied to any product. This is also true in the framework of the natural gas business. What differs from project to project is the product uses, its application and the associated business. There is a need to understand the product history since this is what has shaped the product and in the case of gas how its use has developed resulting in it being regarded as the fuel of the future. There are many reasons why gas is so unusual. The high cost of gas storage and consequent need of continuity all the way from production to point of use this of course can be turned into an advantage as continuity of supply can be a factor and offered to the customer when justifying change from existing fuels, e.g. coal and oil. The fact that natural gas comes out of the ground apparently "free" and that gas operations, at times, can be very profitable. This has resulted in piped gas systems being built throughout Europe and elsewhere. The relatively risky way of handling gas at high pressure, or, after liquefaction is also another reason for peculiarity of natural gas. In addition, the frequent contact with natural gas of complete laymen, or more exactly, and more frequently, laywomen, with the resultant need for safety and high standard of work, including quality, characteristics and behaviour.

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It is interesting to compare gas industry development in Europe and Japan, on the one hand, with that in the USA, on the other. In the USA, manufactured gas never made a particular impact because of (i) the wide availability and low cost of liquid fuels, (ii) the absence of gas lighting and (iii) US cities largely managed without illumination until the arrival of electricity, the civil war and its consequences during the period of town gas development in Europe and, most importantly, the availability of associated natural gas from oil production all over the country which established a price standard difficult to meet for gas manufacturers. In Europe, Japan and a number of less important gas-consuming areas, the non-availability of natural gas had become an accepted fact. The Netherlands enabled the gas industries in Germany, France and Belgium, and to some extent Italy and Switzerland to use it as their main source of supply.

Gas transmission lines from Texas to New England, Canada to California, across Canada and from Mexico into the US, became the ancestors of major European supply mains from Russia, the Norwegian North Sea and Algeria into central and Western Europe including the UK, in addition to the new lines which had already been laid for the transport of Dutch gas. Obviously, the new transmission mains do not all supply gas across national frontiers, and large diameter high-pressure distribution grids apart from the US have also been laid in most major European countries, in the Arabian peninsula, India, South America and Australia. Interestingly, few Japanese gas utilities are interconnected by high-pressure mains, a fact which can be explained in terms of their reliance on LNG for a gas supply.

8.1.1.1 The Business of Natural Gas

Natural gas and its place in the world as the fuel of the future are becoming increasingly evident in recent years. There are constant references to the demand, customers, investors and profit in the business of natural gas. It is thus appropriate we should understand the business elements which can influence these areas and the need to ensure the right level of profit so that the future business developments can take place as well as the level of return on investment to the financial industry.

It is not easy sometimes to see how, as individuals we can support the business, after all, we are usually recruited and employed for our, hopefully, high individual skills and expertise in engineering, geology, finance, etc., but it is essential that all understand and appreciate how we can support the business of producing and selling natural gas at a profit. Figure 8.1 typified the concept of the “Gas Chain” as a model.

The gas chain concept assumes that a complete chain exists between the source of natural gas and the point of use. There are also problems of security of supply and unit fuel cost, as compared to other fuels, as customers are interested in the comparative cost of heat at the burner tip. This is further complicated by high-volume users, speciality industries, low-volume users and chemical users, where natural gas may provide part of the feedstock for the manufacture of the product.

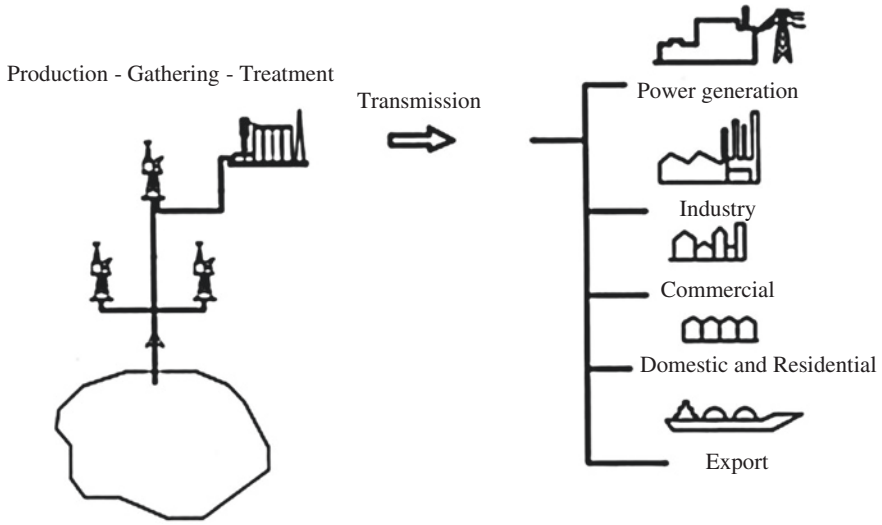


Fig. 8.1 Typical gas chain concept

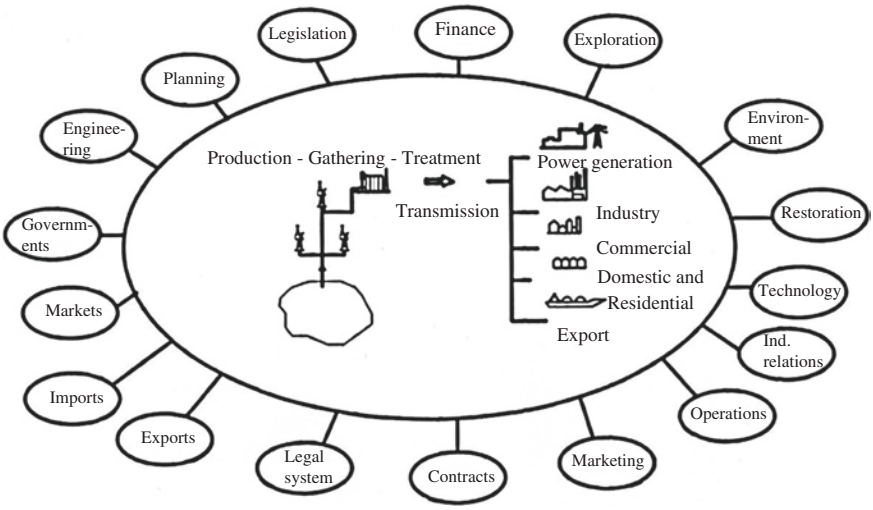


Fig. 8.2 Typical business model

Let us start to build-up a business model from this as typically shown in Fig. 8.2 and by taking each element in turn raise our awareness of its affect on the business of producing natural gas. The key factors of the gas chain model are as follows:

- There must be a complete process chain from wellhead to the customer.
- Each link in the chain must be correctly designed and sized so that there are no weaknesses (production and maintenance) so that continuous operation and availability may be achieved.

- There should be close co-operation of all participating parties, preferentially with investment participation and interest.
- All elements of the chain should be appraised and understood.

Also the key elements of the gas chain are as follows:

- The market prospects and potential value of gas in the market.
- The competitiveness of other projects and competing energy supplies.
- Transportation costs throughout the chain.
- The commitment of all organisations involved in the chain to invest and develop their part of the chain.

The process and associated technology of natural gas can be found elsewhere in the preceding chapters, what is important however is that at the end of this phase, large amount of capital has been used and there is no product to sell. Hence, the business demands sourcing the finance against predictions and estimations of volume and quality. At this stage, even with today's technology, finance is a gamble and may be sourced by governments or companies, from within their own funds. Capital funding from elsewhere will demand high level of return on the investment. However, when gas is found and sampled, then the picture changes and with volumes now predicted and quality assured the investment becomes more clear and at a lower level of risk delivery now takes over and planning of finance, construction and especially marketing become crucial to the business to ensure the necessary funding. While accurate estimates of costs of construction, operation and ongoing maintenance are crucial it is the market potential and realisation and the projected profit after taxes, royalties, license cost, etc., which will make the business viable.

In addition the market for natural gas can be segmented according to its characteristics of volume of gas to be supplied, achievable selling price, quality and end use. Users demanding high volumes and comparable heat at the burner tip will expect a lower unit price than say a chemical producer who uses gas as both source of energy and part of the feedstock for the product who may be prepared to pay a higher unit price while taking large volumes of gas.

Other customers/users may demand a clean fuel, e.g. the ceramics manufacturer, in addition to fuel economy with the resulting negotiated unit price. Ease of use and convenience may be the priority of the commercial and domestic sector which although demanding low volumes are prepared to pay a higher unit price.

All the above will require a clean, high-quality fuel and security of supply in addition to a comparable *Net Cost* of heat at the burner tip.

To provide an example of Net Cost, it may be worthwhile to refer this to the power industry in moving to natural gas. The power industry has depended on the use of coal, and conversion to a more environmentally friendly fuel is desirable. Processes for the scrubbing of exhaust gases from power stations, to eliminate sulphur products and particulate matter, exist but are not without their problems of high-operational and maintenance costs along with the production of gypsum as a waste product. The latter has, of course, decimated the production of gypsum from

natural sources, and there is an oversupply of material to the subsequent manufacturing industry which results in a low-price product. In addition to this, the use of coal produces fly ash, which although has some use, e g as an additive in the production of cement and some building products, still results in waste which is not part of the core business of producing electricity. This cost of course can be avoided if the power station converts to natural gas. In addition, the cost of the operation and maintenance of coal, limestone and fly ash processing and handling plant can be eliminated along with the corresponding reduction of staff. These costs should be taken into account when making the case for conversion to, and using natural gas to replace coal. The increased efficiency of the process is also attractive. All these must be factored into the calculation of the net cost of heat delivered at the burner tip and comparing fuel cost. Natural gas may cost more but can be cheaper, and more environmentally friendly, than using coal and oil. Similarly, it is possible to carry out an analysis of other fuels which makes the use of natural gas more attractive in terms of an overall model for a respective business. The natural gas business has more a flexible pricing mechanism, and an individual customer can bargain for its own price, unlike oil and coal, which have a fixed market price. Furthermore, the natural gas supply can be flexible, firm or interruptible, and a lower price may be possible depending on the process requirements.

Within the natural gas business model considered, the ideal supply profile should also be considerable. Figure 8.3 shows an ideal total supply profile from a suppliers point of view, where the agreed total maximum daily supply profile closely follows the daily contract quantity thus utilising the pipeline capacity to the full. In practice, this is difficult to achieve due to the vagaries of climate and activity of the customer, and careful estimation and aggregation of a large number of supply profiles is required for maximum sales of gas and the calculation of

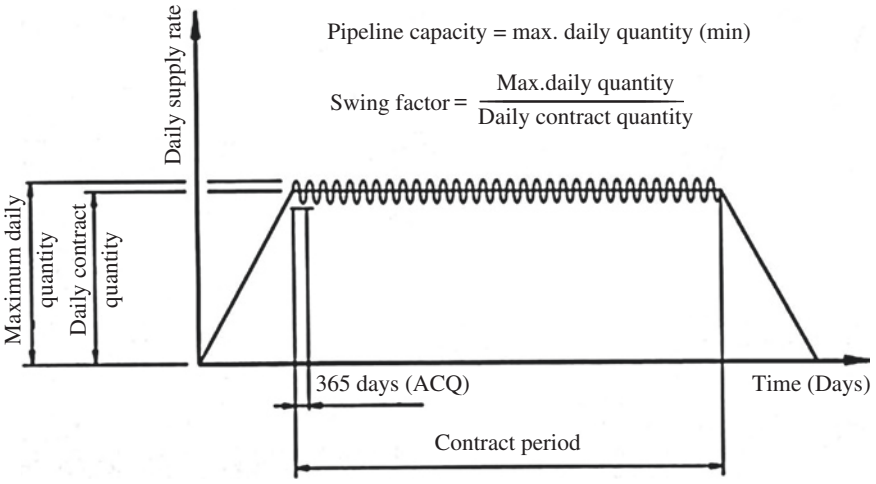


Fig. 8.3 Ideal total supply

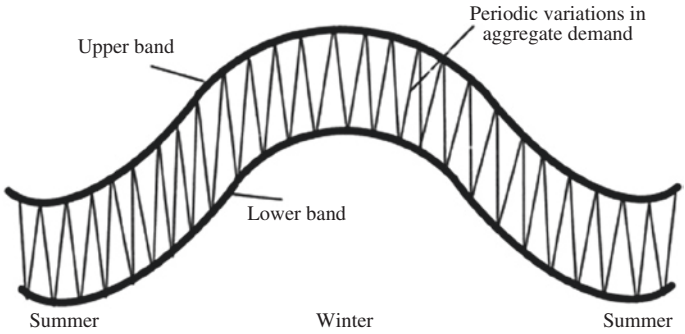


Fig. 8.4 Variation in demand profiles

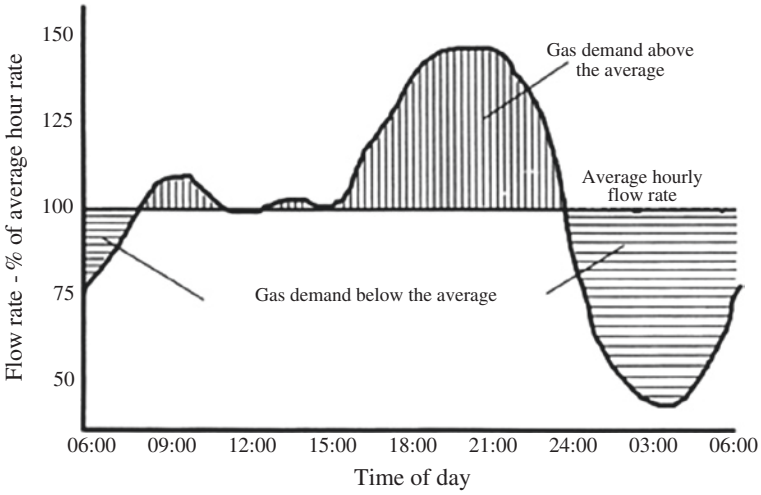


Fig. 8.5 Diurnal variations in natural gas

minimum production costs. The production of LNG can be utilised to some extent to modulate the profile as can the production of compressed natural gas. Both of these can be seen as storage.

Further complications can arise from climatic conditions and the work/usage requirements of customers who are willing to pay varying amounts for their gas depending on volume, and timing and duration of demand periods. All will require quality and security of supply at the lowest price. These areas of influence are depicted generally in Figs. 8.4 and 8.5 and can be further compounded by interruptible supply profiles, resulting from dual-fuel usage. Nevertheless, the market must be understood and estimated with each contract agreed to suit, as described in the next section.

8.1.1.2 Natural Gas and Market Sectors

The Power Generation Market
This market is characterised by the use of very large volumes of gas which may vary between winter and summer, and periods of transformation between those periods. This effect can be dramatic in countries where there is a variation in daylight hours and climatic conditions as, for example, in areas of Northern Europe and the UK. The resulting Diurnal variation (as shown previously in Fig. 8.5) can further affect the daily demand profile. The implications of this are as follows: while they demand high volumes of gas, they result in a poor daily load factor and can be further complicated by the influence of interruptible contracts. While all of this contributes to a low transportation cost and places the customer/buyer in a good position to bargain for a low price for the gas, it puts the supplier in a good position to agree "take or pay" contracts. This is good for the overall business in that the producer gets an agreed annual income which satisfies his production and the demands of repaying loan capital and satisfying shareholders.
A load factor measures the ratio of gas consumption at times of peak demand compared with average consumption. A declining load factor means that peak demand is growing more quickly than average consumption. A declining load factor presents issues such as increased spending on network infrastructure which is used for small periods of the year; increased requirements for peak load power stations that can be switched on and off quickly; and greater variation (or volatility) in demand [2–4].

The Chemical Feedstock Market
This market is characterised by the use of very large volumes of gas without the affects of seasonality and diurnal variations. Processes require a predictable chemical quality at a high-delivery pressure. This results in high-throughput rates and a high-load factor and a more efficient use of the pipeline and low-unit transport cost. There is also the possibility of dual use where natural gas may be used as a fuel as well as a chemical feedstock.

The Industrial Market
In this market, a large number of locations can be geographically spread and utilise large volumes of gas which are only partially affected by seasonality. Throughput is reasonable, but load factors can be variable. Conversions can be affected by high cost of installing a pipeline feed from the gas main supply.

The Commercial Industry Market
This market can utilise a wide range of volumes and be affected by diurnal variations and seasonality. They can be even more geographically spread. The implication to the gas company is the supply profile can be the result of the aggregation of volumes supplied by a number distribution companies and provide poor daily and annual load factors and high-investment costs.

The Domestic Market
This market is geographically spread and utilises comparably very small volumes of gas which are subject to diurnal and seasonal variations.

The implication to the gas company is the supply profile can be the result of the aggregation of volumes supplied by a number of distribution companies and provide very poor daily and annual load factors and very high-investment costs.

Other Market Issues

Future Load Growth

While the existing real market should be the basis for the case for investment in the project, any perceived growth should be recognised, quantified and factored into the overall case as sensitivity in the financial analysis. This provides an indication of the overall business potential and financial security, and of course will satisfy and encourage investment agencies to invest in the project. Future load growth should be real and not a fanciful look into the future and should be supported by facts and logic. The finance industry does not like to gamble and will seek a substantial basis for the predicted growth.

Inter-fuel Competition

This is when other fuels seek to be used as alternatives. Natural gas of course, unlike oil (possibly its main competitor), does not have an international price base and is free to compete in all markets. It is fast becoming the fuel of choice for all industries due to it being environmentally friendly, controllable and with security of supply due to large reserves. However, it would be a mistake to take inter-fuel competition for granted as suppliers of other fuels seek to solve the environmental problems associated with their use and compete on the basis of comparing the net cost of providing heat at the burner tip. One area yet to be developed, certainly in the UK, is the use of natural gas to power transport. While the use of gases to power vehicles goes back to the wartime years, it has never been developed due to the dominance of oil and the lack of an infrastructure. Petrol stations are owned by the oil companies and are now selling LPG mostly in parts of Europe. However, less developed countries, where the infrastructure for the sales of fuels to vehicles are being developed, are building service stations for the delivery of natural gas at the pump and the population seeing the resulting economies. Natural gas is now being offered in the USA but has yet to be exploited in the UK and Europe and a number of car and heavy vehicle manufacturers are offering dual fuel and gas powering as an option. This will provide much needed competition of pricing and affect the world's dominance and reliance on petrol and oil. The use of natural gas in transport could be the next great influence on economies, resulting in cheaper transport costs.

The Development of New Fuels

The world has now become increasingly environmentally aware due to the increased domination of oil and its perceived dwindling of recoverable reserves.

While the use of natural gas is increasing, and backed by huge reserves, it remains a fact that the reserves of coal significantly exceed those of gas, and of course gas can be produced from coal. Other fuels such as nuclear and hydrogen are in differing stages of use, and development, and provide a constant threat to the business of providing power and heat. Large amount of capital is in the process

of being invested in wind power in the USA, Europe and UK, and projects exist to examine the ways of producing power from the large tidal flows which exist, certainly around Europe and the UK.

However, these methods of utilising natural resources are not without their obvious problems. At this stage, natural gas seems to be the most available source of energy.

Governments

Governments are under constant pressure to reduce the cost of living and in particular in the areas of heating and transport. They prioritise these in their manifestos along with refining the country's economies, and the fact that they may be up for election/re-election may produce changes in government pressure. This can manifest itself in many other ways which are in their control such as import and export duty, royalties, purchase tax, duties of fuels, etc. It is possible that a government of a country can change a number of times in the life of a gas project, and therefore, political awareness is important, and working closely with governments is paramount. The fact that gas pipelines now cross many borders, means that the supply of gas can become internationally political and it must be recognised, and be a major factor in trying to solve conflict. There have already been instances where the gas supplier has switched off its supply to other countries for political reasons. The supply of energy and ownership and control of reserves has in the past been a major factor in declaration of war between countries, and it is logical to conclude that this could still happen as fuels become less abundant. This of course can result in higher risk of investment capital, and investors will seek a higher financial return commensurate with that risk.

While supply by pipeline presents the lowest delivered price for natural gas along with security of supply, it does in itself present its own problems. The fact that pipelines cross-borders make the delivery politically sensitive. This is further exacerbated by ownership issues and pipeline charges, which may affect the cost of delivery. Indeed, there have been instances where gas supplies have arguably been more effective than military action in resolving disagreements between governments. This of course does not mean that selling and control of gas supplies are a deciding factor but one cannot rule out that they may in turn be the cause of military action as have the issues associated with the supply of oil. All need to be assessed and taken into account at the exploration stage and prior to seeking and procuring investment capital.

8.1.1.3 Gas Pricing

When seeking to negotiate a price for the sale of gas, it is important to recognise and distinguish between mature and immature markets. In immature markets, monopolies may exist where gas is purchased from the producer by a single organisation and sold directly to the end-user, effectively dictating prices and contractual terms. State-regulated price structures may exist for producer and consumer, which may be politically directed to the extent that some industries may

be excluded from supply. For example, a government of a country rich in coal reserves may exclude the use of gas as a fuel for the supply of electrical power. This market of course is characterised by a take it or leave it attitude and an absence of inter-fuel competition.

The industry may be state owned, with field production sold under depletion contracts where the whole field capacity is sold by joint venture. Pricing may be a composition of base price, further affected by price indexation formulae and provisions for price revision. Any short-term contracts will be characterised by a published tariff system of pricing and contracts which are standard. The market is divided into tariff sectors by the seller of the gas to market sectors. The whole market being characterised by a lack of competition on price and volume and a take it or leave it principle. On the other hand, mature markets may have long-term contracts agreed but these are decreasing in number and volume with time as competition develops within the industry. Medium-term contracts may exist for sales by producer direct to end-user similar to a large number of short-term contracts where the duration may vary to satisfy the end-user.

Contract provision and complexity may depend on contract duration, and consideration of alternative suppliers may take place as gas is offered at varying prices. Contracts may take a standard form of agreement and offer an agreed price for the contract period. This has been developed to such an extent in the UK that gas supply companies have now turned into energy supply companies offering financial inducements for home owners to buy electricity and gas from them as a package. However, the base price of the gas must be competitive with other fuels and based on a comparison of the *NET COST OF USEFUL HEAT AT THE BURNER TIP.*

The determination of base price should take into account cost (economic) savings arising from the use of natural gas, i.e. higher efficiencies, reduced pollution, no storage requirement, improved product quality, environment costs and reduced maintenance costs

The Buyers Objective

The agreed price from the buyer's point of view must cover the cost of conversion from an existing fuel and any capital and operational cost associated with the business of selling gas and then remain competitive with the alternative fuel price. This should include any business taxes which of course can be variable depending on government policy. A cost/benefit margin, for example, profit which should be included and that will at least permit the price to remain competitive in its markets.

The Sellers Objective

In real terms, the seller's price must permit recovery of investment, operating and finance costs. Care should be taken to include all associated costs and the relative cost of inflation over the life of the project. Money has a value which varies in relation to time which is difficult to calculate.

In business terms, 5 years can be almost regarded as geological time, as world economies, finance and governments change. However, the affects have to be estimated and sensitivities analysis carried out on the business case. In addition,

a profit element is required commensurate with risk, shareholders dividend, and finance company expectations.
The aim of pricing mechanisms must maintain the division of economic benefit between buyer and seller and is vital to the success of the business, hence the importance of price revision and review clauses. They are not a gamble on future trends. Gas price revision formulae can be simply multiplicative or additive or more complex and has restraints incorporated within them (e.g. top and bottom limits)

Contracts agree a base period, a review period and frequency of any agreed price revisions. Frequent price revisions will result in the price staying close to the market values whereas infrequent revision, if indices rise will be good for buyer and bad for seller.

The Base Period and Review Period
The base period is the starting point and is the period prior to the period to which the base price applies and published data must be available during the base period. The review period is the period for which price is calculated, and base period/review period relationships must be consistent. Base period values may affect price.

Price Break Clauses
The advantage of including a price break clause is to provide protection against unforeseen risks. The disadvantages, of course, are they are very difficult to negotiate and may lead to prolonged disputes if used.

The pricing clause should be reviewed at regular intervals. If market conditions change, the price changes. The price break clause may be included by the buyer or the seller, and arbitration should be sought if no agreement is reached. Clauses may be simple where the buyer takes a fixed margin and the balance goes to the gas producers. Prices could be revised if competing fuels indexation components change. Complex revisions may reflect major changes in circumstances and relative values to parties, and the base price may also be changed.

Negotiation
The negotiation of a contract can, if not managed properly, result in confrontation where each side is trying to maximise its own benefits. The need to win, or to be seen to win, at the others cost can override the main objective which should be that each side should have a financially viable business at the end of the negotiation. It is essential to appreciate that a contract is an agreement between the parties and all should be comfortable with the outcome. If this is not so, the management of the contract may become confrontational and difficult to manage. The following are some areas where negotiation can become difficult and result in a situation where both parties are unable or unwilling to negotiate and take an immovable stand on their conditions, .i.e. *Deadlock*

- The determination of the base price: this is the initial price which will be subject to review.
- The buyer will want to ensure the price covers the cost of conversion from existing fuel and then remain competitive with alternative fuel price. One requires also a cost/benefit margin that will at least permit the buyer to remain competitive in its markets.

- The seller would like to get a price in real terms that will permit recovery of investment, operating and financial costs of the gas project and a profit element commensurate with risk, shareholders dividend and finance company expectations. These conflicting objectives of buyer and seller may lead to a situation where discussions cannot continue and may eventually end up in a deadlock.
- The price revision gas formula would be used to review the gas price time after time during the life of the contract. It will contain all the parameters considered relevant to the value of the gas with respect to time, by the buyer, and the seller. The aim of pricing mechanisms is to maintain the division of economic benefit between buyer and seller and not to gamble on future trends.
- Gas price revision indices and review periods are potential areas of misunderstanding and disagreement, which may lead to *Deadlock* if not properly managed. Terms such as "frequency of price revision, base period and review period" should be well understood by both parties and the base period/review period relationship must be consistent.

The "take or pay" provision is the quantity of gas, when supplied by the seller to the buyer in accordance with the contract terms, must be paid, whether taken by the buyer or not. This quantity can be taken in the future as "make up" gas, at no cost within the period specified in the contract. It is usually expressed as percentage of Annual Contract Quantity (ACQ). This is an essential component of most long-term gas contracts. The seller would like to have guaranteed revenue to justify investment and obtain finance.

The "swing factor" is the measure of degree of fluctuation of quantity of gas, required by the buyer, which the seller, would be obliged to provide when the buyer nominates. The buyer would like to have flexible off-take to match the seasonal, daily or hourly fluctuations, while the seller, would like have constant rate of production because off-take flexibility costs money in terms of investment and operational costs.

Quantity provisions need effective and understanding communications during negotiations and if they are not properly managed can lead to deadlock.

A Strategy for Breaking

First reframe and represent your position on the issue. Let it be clear, concise and adequate to avoid wrong interpretation. It is possible that the other party misunderstood your position, which caused the deadlock. One may then soften your position and offer an exchange of concessions. For example, lower take or pay quantity for lower *swing factor*. Make sure the costs and benefits of these concessions are well understood. This might make a lot of difference to the opposing party and resolve the deadlock. Suggest that both negotiating teams should consult their management with a common proposal on the issue at stake to resolve the *deadlock*. Call off the meeting during which the negotiation came to a deadlock to get time to analyse and evaluate issues already discussed and agreed upon. Make necessary consultations and prepare for a new proposal for the next meeting.

Suggest the formation of a small committee comprising people from both parties to work out an acceptable common position on the issue, which both parties would present to their management for consideration.

Recommend possible concessions that can be offered during the next meeting to break the deadlock to your management for consideration highlighting the costs and benefits of these concessions. Identify the other party's concern on the issue at stake and appreciate their concern and give assurance that their concern will be respected as much as it is feasible during the next meeting. Call for the next meeting to resolve the deadlock and move forward.

8.1.1.4 The Legal Framework of Gas Contracts

The upsurge of interest in natural gas as an environmentally friendly fuel makes it more necessary than ever for gas contract negotiators to have an understanding of the important role of the law in relation to gas contracts. The main reason for the need to structure gas contracts on a sound legal basis and to draft them with great care is the long duration of the contracts, often 15–25 years [6]. During such a long period, there will almost certainly be significant economic, commercial and technological changes, which have a bearing on the contracts. Contracts must be structured and drafted in such a way that they will be able to survive such changes.

The difference between oil and gas sale contracts

Oil sale contracts are generally of short duration, rarely more than 1 year. Gas contracts have a totally different time scale, normally 15–25 years. The reason for the difference of duration is physical. It stems from the relative ease of transportation of oil as compared with gas. Oil may be relatively easily transported by sea or on land, and this fact has been an important element in the creation of an international market in oil.

Speaking in general terms, it is not a matter of concern to an oil producer on the UK Shelf, or in the Gulf, or Nigeria, where in the world its production is marketed. There is a worldwide market, and, in normal market conditions, there is no reason for the producer to fear that its production will not be sold at a profit. In contrast, gas is not so easily transported. Transportation of LNG is of course practicable but liquefaction and reconversion of gas makes this gas expensive in comparison with pipeline gas.

In consequence, the majority of gas supplies are transported in the gaseous state in pipelines to the specific market where they are to be sold. Because of the very large capital cost of developing major gas fields, it is very important to a gas producer to ensure that, before developing a new gas field, it has entered into long-term contracts for the sale of a large part of the production of the field. The buyer, because of its normal concern for security of supply over a long period of time, will also in most cases seek a long-term contract.

The Contents of a Typical Gas Contract

The contract will contain provisions dealing with the following matters:

i. The obligation of the seller to sell, and of the buyer to buy over the contract period, a quantity of gas and also provisions dealing with the daily quantities to be delivered

ii. The contract duration and start-up date
iii. Warranties by the seller relating to title of the gas and to taxes
iv. In the case of depletion contracts, reservations by the seller for its own use of a part of the gas from the field
v. Take or pay obligation, and rights of the buyer associated with that obligation, such as rights relating to make up and carry forward amounts of gas.
vi. Base price, price review provisions and, sometimes, price reopener provisions
vii. Billing and payment
viii. Force majeure
ix. Default (and damages)
x. Expert determinations
xi. Delivery point, quality and pressure
xii. Assignment

It is the contractual provisions dealing with these matters which constitute the essential framework of a gas contract.

The Main Areas of Law Which May Be Applicable to Gas Contracts

Some are more relevant to the interests of one party rather than the other and not all of them will be applicable to every gas contract. However, all of them should be considered before entering into a gas contract. As we go through each of these areas, we will look at their relevance to gas contracts in general, and then describe their relevance under UK law and in some instances comment on their relevance under other systems of law. There are nine main areas of relevant law. They are as follows:

(a) *The Law Dealing with the Granting of Rights to Produce Oil and Gas*

Relevant matters will include the duration of production licences and the terms relating to royalties, including royalties in kind and production consents.

The UK law is contained in the Petroleum Production Act, 1934. The regime set-up under this legislation provides for grants of exploration licenses and production licenses for a specified area. The production licences grant a right for licensees to search for, bore, and get hydrocarbons from, the specified areas for a specified period.

It is important for the buyer of gas to know who the licence holders are and that the parties who intend to sell the gas do in fact have title to the gas which they propose to sell. It does not always follow that the licence holder will have title to the gas, for example, for tax reasons, the licensee may have arranged that the gas should be sold by one of its subsidiary or associated companies. The duration of the licence is clearly a fundamental element to bear in mind in considering the duration of the contract. It is worth noting that under the UK law there is no automatic right to an extension of a licence if at the end of the licence period there are unrecovered reserves still in the ground.

(b) *The Law Relating Specifically to Sale of Goods* (*as Distinct from Land*)

Relevant matters dealt with in this area of law will include the time at which the ownership and risk in the goods passes to the buyer. In the UK, the relevant

statute is the Sale of Goods Act, 1979. This contains general rules about sale and the rules relating to the warranties which are deemed to be given by the seller (e.g. title to goods and quality). The law relating to contracts in general are as follows:

Relevant matters under this heading are the law relating to mistake and misrepresentation.

(i) In the UK, this law is mainly to be found in the common law which is a body of uncodified customary law. However, the Misrepresentation Act, 1967, and the Unfair Contract Term Act, 1977, are also relevant.

(ii) In contrast to the position in the UK, the general contract law is, in most countries, codified. For example, France has the Commercial Code and the USA has the Uniform Commercial Code which applies in almost all states.

(c) *The Law Relating to Force Majeure*

Force majeure provisions are common in all types of commercial contracts. They have a particular importance in gas contracts largely because the contracts are of such long duration. They are particularly important for the seller because of the hazards of production and transportation of gas. Under English law, and the law of most countries whose legal system is based upon English law, there is no statutory definition of force majeure. In the case of contracts which are subject to English law, or the law of such other countries, a specific force majeure clause should be drafted which meets the needs of the individual contract. In contrast, in France, and in most other countries whose system of law is based upon French law, there is a statutory definition of force majeure.

The definition in French law is limited and not satisfactory for gas contracts.

The definition of force majeure requires that, in order to qualify for force majeure relief, the event which caused the failure must satisfy two tests:

(i) It must be one which could not have been prevented, and
(ii) It must be one which was not foreseeable.

Many events which may not be capable of being prevented are foreseeable. For example, the possibility that a submarine might collide with a platform may not be capable of being prevented but it is foreseeable. A producer would properly require force majeure protection for such an event. The legal solution is to extend the statutory definition of force majeure in the contract so that the event is covered.

(d) *Special Legislation Relating to the Buyer*

Where the buyer is granted by the state a licence to carry on an activity as a gas distributor, which is the case for most utilities, it is important for both buyer and seller to consider the terms of that licence. In the UK, there is special legislation relating to British Gas PLC. British Gas is a public gas supplier under the Gas Act 1986 and under that Act it has an authorisation to supply gas for an unspecified period. There are no special statutes governing the activities of the principal UK electricity power producers (National Power and PowerGen).

(e) *Tax Law Relating to Production and Sale of Hydrocarbons*

Production taxes may be a factor to take into account in evaluating the profitability of a project and may have contractual implications particularly in relation to the termination of depletion contracts. In the UK, there are special laws limiting the right of producers to set off expenditure on one hydrocarbon project against the profits made on another such project, the ring fence concept. Also petroleum revenue tax ("PRT"), a tax on hydrocarbon revenues and not on profits, is a relevant factor in determining profitability. The impact of sales taxes may be relevant and should be considered because they may give rise to disputes about whether they are to be borne by the seller or passed through to the buyer.

(f) *Competition Law*

The competition law relating to joint activities may have an impact on a group of sellers or buyers in relation to how they should contract, as production licences are often held by more than one party this area of law can often apply to sellers. The competition law relating to monopolies may be applicable to the buyer.

For example, in the UK, the statute which deals with joint activities is the Restrictive Trade Practices Act, 1976, which may apply to the activities of producers in joint developments and also to joint buying consortia. The Fair Trading Act, 1973, and the Competition Act, 1981, are relevant to monopolistic behaviour. Also the competition provisions of the Treaty of Rome can be relevant to gas contracts.

As a single European gas market develops the importance of the competition provisions of the Treaty of Rome (Articles 85 and 86) will increase not only in relation to gas contracts but also in relation to gas transportation.

In relation to US-based companies and their UK subsidiaries, the provisions of US anti-trust legislation are also relevant.

(g) *Arbitration law*

This is normally only relevant in international contracts. Those were the seller and buyer carrying on business in different jurisdictions. In the case of contracts for the supply of gas from the UK shelf to the UK, the parties normally accept for settlement of disputes the jurisdiction of the ordinary English courts in preference to arbitration. The reason for this is that arbitration is perceived by most companies to be more expensive than proceedings before the ordinary courts. That factor is considered to outweigh the benefit of the secrecy which accompanies arbitration proceedings. The relevant provisions of English law are contained in the Arbitration Acts 1975–1979. In the case of international contracts, an agreement to settle disputes by arbitration is often desirable as neither party will agree to submit to the jurisdiction of the courts of the other country. It is important in these cases to agree at the time of entering into the contract procedures for appointing the arbitrator(s) and for conducting the arbitration. Frequently, parties agree to be bound by the arbitration rules of international arbitration bodies such as the International Chamber of Commerce (based in Paris) or the Stockholm Chamber of Commerce. There are other similar bodies in the Middle East and Asia.

(h) *The Governing Law of the Contract*

If the parties are incorporated in the same state, then it is not normally necessary to specify that the law of the state of incorporation applies. However, it is necessary to deal with the matter where the parties are incorporated in different states. This choice of governing law is important as it establishes which state's law is to apply to the contract. In practice, the choice is between the law of the state of the buyer and that of the seller. There are legal dangers in selecting the law of a state which has no connection with the contract.

It is important to remember that the choice of law determines the rules which will apply to interpretation of the contract. These rules differ considerably from one legal system to another. This is well illustrated by the differences between English and Norwegian law. Under English law, it is not permitted, when interpreting the contract, to refer to any document other than the contract itself. Also under English law, the intentions of a party are established by a strict interpretation of the language used in the contract. Under Norwegian Law, the position is quite different, it is possible when interpreting the contract under that law to refer not only to the contract but also to working drafts of the contract and Minutes of Meetings. Also the interpretation of the language used is not so strict. This difference must be borne in mind by anyone proposing to purchase gas from Norway as it is likely that Norwegian sellers will insist that Norwegian Law is the governing law of the contract.

(i) *The Language of the Contract*

It is customary in the gas industry for contracts to be in the English language. However, one party may insist upon a text in another language. In such a case, it is important to agree on provisions to cover the case where there is a conflict between the two texts.

(j) *The Differences Between the Depletion Contract and the Supply Contract*

There are essentially two types of gas contract:

(i) *The Depletion Contract*

This contract, is one in which, the producer agrees to sell the whole of the economically recoverable reserve ("ERR") of a field. All the gas in the field is dedicated to the buyer except for such quantities reserved for the seller's own use. The economically recoverable reserves are those reserves which a reasonable and prudent operator might reasonably be expected to recover from the field under the terms of the contract. The ERR which is actually recovered from a field will depend on such unpredictable factors as reservoir performance and the fluctuations of the contract price. It follows that the quantity to be sold under a depletion contract is uncertain and that the contract duration is uncertain. It should be noted that it is the buyer who bears the risk of this uncertainty; the seller bears the risk that the development which has been put in place may not be profitable. It should also be noted that it is possible for a producer to terminate a depletion contract before any gas has been produced if it can show that there are in fact no economically recoverable gas reserves in the field.

(ii) *The Supply Contract*

This contract is one in which the producer agrees to supply a specific quantity of gas over a specific period of years. In contrast to the depletion contract, it is the producer who takes the responsibility, and therefore the risk, of ensuring that the supply of gas contracted to be sold is made available to the buyer.

Under the provisions of a supply contract, the producer cannot claim force majeure relief on grounds of unexpectedly poor reservoir behaviour. If the gas is not made available, the producer will be in breach of contract and liable to damages unless force majeure relief is available. The distinctions drawn between a depletion contract and a supply contract have been put very starkly.

Depletion contracts are negotiated which contain some elements of a supply contract, and supply contracts are negotiated which contain some elements of a depletion contract. For example, some depletion contracts contain what is known as a minimum plateau period, which is a period in which the producer guarantees a specific level of deliveries thereby providing the buyer, during this limited period, with the level of security of supply obtained under a supply contract. In the case of supply contracts, there are instances where the contract provides for force majeure relief for reservoir failure and thereby gives the producer a level of risk which is comparable with that of a depletion contract.

In the UK, supply contracts have come into use as a result of the privatisation of British Gas, the Monopolies and Mergers Commission report already mentioned, and the emergence of the new market of gas supply to power stations. Prior to that time, gas bought by British Gas was bought under depletion contracts. However, supply contracts have been for some years the normal method by which Norwegian and Russian gas has been sold into continental Europe.

It is interesting to note that although the essence of a supply contract is an obligation to supply a specific quantity of gas over a period of years, the contract must accommodate the wishes of the buyer to take different volumes from day to day.

The buyer's need for variation in quantities will depend upon a number of factors including weather variations. The duration of the contract should therefore be sufficiently flexible to permit early termination, if the total contract quantity agreed to be bought has been bought by the buyer before the end of the originally envisaged termination date, and to permit an extension of the contract if the quantity has not been taken by that date.

(k) *Contractual Terms*

The following is an explanation of some of the contractual terms which are used in connection with take or pay.

The Take or Pay Obligation

This is an obligation on the buyer to take delivery of a specific quantity of gas over a specific period (normally a year but it may be quarterly or half yearly), or if it does not take delivery of that quantity pay the seller for a quantity of gas equivalent to that which it fails to take.

Minimum Bill Quantity
This is the quantity of gas which the buyer undertakes to take or pay under its take or pay obligation. This quantity is derived from the Annual Contract Quantity. In order to compute the Minimum Bill Quantity, it is necessary to deduct from the annual contract quantity the quantities which the buyer failed to take in the contract year for reasons of force majeure, and the quantities which the seller failed to deliver for any reason in that contract year.

This figure is known as the Adjusted Annual Contract Quantity. From that adjusted annual contract quantity, a deduction is negotiated and the figure resulting from such deduction is the Minimum Bill Quantity.

Daily Contract Quantity
This is an agreed daily quantity by reference to which the buyer makes its nominations for the deliveries of gas on each day.

The*Annual Contract Quantity*
This is the quantity of gas which the buyer expects to take in a contract year. It is equal to the product of multiplying the Daily Contract Quantity, applicable for the relevant contract year, by the number of days in the contract year.

Make Up Right
This is a right of the buyer in any contract year, after it has made a take or pay payment and has taken delivery of the Annual Contract Quantity for that contract year, to take delivery free of charge of a quantity of gas up to an amount equal to the quantity in respect of which it has previously made a take or pay payment.

The*Carry Forward Right*
This is a right of the buyer to reduce the Minimum Bill Quantity in a contract year, where in previous contract years the buyer has taken a quantity in excess of the adjusted Annual Contract Quantity.

The Take or Pay Provision
The important legal point to bear in mind in connection with the take or pay provision is that the payment made by the buyer if it fails in the relevant period to take delivery of the minimum bill quantity required under the take or pay provision is not a damages payment. It is a payment made in order to meet an obligation to maintain a minimum cash flow. By contrast, a payment by the seller under the default clause is a damage payment. The damages will be determined in accordance with the law if not agreed in advance by the parties.

Price Review and Price Reopener Clauses
Generally, it is important to appreciate the different functions of the price review clause and the price reopener clause. The price review clause is automatically applied at stated intervals say 3, 6, or 12 months.

It is intended to cause the contract price to move upwards or downwards against certain defined indices and thereby ensure that both from the buyer's point of view and the seller's point of view, the gas price in normal circumstances will over the life of the contract remain in line with the commercial aspirations of the

parties when they entered into the contract. By contrast, the price reopener clause does not apply automatically; it cannot operate unless invoked by either party. It may only be invoked at stated intervals, normally not more frequently than 3 years. If the parties do not agree that the circumstances in which it is expressed to be capable of being invoked have arisen then, unless the person called upon to adjudicate upon the dispute, decide that they have arisen the clause cannot be invoked.

The clause is designed to deal with those instances where, as a result of the occurrence of events which were outside the control of the parties, the operation of the price review mechanism has not proved capable of maintaining the original commercial balance of the contract.

The commercial balance will have been upset if, in the case of the buyer, the buyer is no longer able to make an adequate level of profit on the resale of the gas, and in the case of the seller, if the contract price which is payable is not in line with the then current prices payable in respect of gas supplied under similar contracts. If the parties cannot agree on the revised price, the dispute is referred to an agreed third party whose discussion as to the revised price is binding on the parties.

The Price Review Clause

The commercial considerations involved in choice of indices and percentages to be attributed to competing fuels will no doubt take most of your time in the coming week. From a legal point of view, there are two important elements in a price review clause.

The first is that the indices required for the operation of the clause should be derived from reliable and readily available published statistics. In the case of the UK contracts, it is usual to take statistics from the government publication "energy trends" but in some cases industry publications such as Platt's are used.

The second relates to the continuity of the contract. In this connection, in order to ensure that at all times a price can be established, it is important to include in the contract provisions to cover non-availability of, and change of indices.

The Price Reopener Clause

From a strictly legal point of view, such clauses have their dangers. The reasons for this are that, inevitably, they must be drafted in general terms and that general drafting always brings with it a danger of dispute as to the meaning of the clause. I say "inevitably" because the parties cannot foresee the precise situations for which they are trying to legislate (e.g. the reconstitution of the OPEC Cartel), if they could do so the lawyers could draft precise provisions to cover them.

The experience of the operation of price reopener clauses in the Norwegian and Russian supply contracts seemed to be "satisfactory". The reason for this seems to be mutual self-interest. The parties tend to agree on the revised price, as they are both fearful of putting the determination of the contract price into the hands of a third party, in addition, as parties to long-term sale contracts are in practice almost inextricably linked to each other, it is not in the interest of either party to allow the balance of the contract to be permanently upset by insisting on unreasonable price terms.

It is, of course, possible to draft price reopener clauses which do not put the determination of the price in the hands of a third party. For example, the parties may agree to meet and discuss if they consider that the gas prices which they are receiving or paying have fallen out of line with the underlying understanding between the parties about profitability of the contract at the time when it was entered into. Such clauses have no legal "teeth" but they can be useful if the parties genuinely wish to make their long-term contract work in changing circumstances.

The Importance of Clear Drafting

The clear drafting of gas contracts is of paramount importance. The contracts are normally of such long duration, and involve such important commercial interest, that it is unrealistic to draft them in the belief that the parties will be able to "put them in a drawer and never look at them again". Points will arise and because of the nature of the contracts, the sums will be almost always significant. It is rare to come across a dispute with a value of less than £5 million and some claims are far in excess of that amount. If the drafting is ambiguous, much management time will be taken up in dispute and the relationship between seller and buyer may well be damaged. In most cases, the parties will reach a compromise, but the balance of that compromise will be determined by the words employed in the contract, with reference to the different methods of contractual interpretation under different systems of law e.g. under English and Norwegian law. This difference will have an effect on the drafting style. A "broader" brush approach will be applied in a contract which is subject to Norwegian law than a contract which is subject to English law. However, it would be highly dangerous to adopt the broader brush Norwegian approach to a contract which is subject to English law because of the strict rules of interpretation of the contract under English law and the fact that references to working drafts and minutes is not permitted.

8.1.1.5 Principles of Project Evaluation

Although the careful evaluation of any project both ahead of its realisation, and also after, to assess the degree of accuracy of the methods used. Assessment discussion is mainly concerned with the technical and economic evaluation carried out by a financier before he agrees to underwrite a project. There are broadly speaking three reasons for this:

(i) Most projects depend at some stage on external finance, and it is, therefore, important to know which aspects will be considered critical by a financier.
(ii) Economic success is, clearly, of equal importance to both financier and owner/operator of a project.
(iii) Any form of feasibility study ahead of realisation is invariably much cheaper than continuing with a mistaken project or, even if the basic concept is correct, moving in the wrong direction.

It may also be worth mentioning that not all financiers and investors seem to insist on carrying out project assessments of their own. Clearly, they have the alternative

of using external consultants, often paid for by the promoter of the project. They may also be content with feasibility studies carried out by, or for the borrower. But in either of these cases, they will generally insist on replies to certain questions, on a number of predictions of certain "business ratios", and on direct contact with operational staff and, if applicable, outside contractors. It is proposed to discuss now in somewhat greater detail, the type of information required to make a meaningful financing decision.

Project Background

Projects are not developed in a vacuum and, obviously, any financial organisation will want to know the political and economic circumstances which have given life to and will regulate the operability and economics of the project. Such information will include, particularly in the case of gas and oil projects, the political facts of life both as far as the location of the project itself is concerned and also in its marketing and raw material supplies area. Is it likely that there will be political changes in the foreseeable future? Are there any ethnic or religious problems? Are there territorial claims which could affect the project, its source of supplies, its markets and its staff?

Little need be said at this stage about the contractual, legal and regulatory background to the project, beyond saying that detailed information about the status of the project in all these respects must be available. It will, clearly, be critical for any assessment that all legal and regulatory requirements have been met before embarking on the project and that the promoter's proposals are within the legal framework of the country in question.

An important aspect of any project is taxation of profits, of turnover, of raw materials and of products. Personal taxes on local and expatriate staff income can also affect project operability. Arrangements must be made regarding royalties payable on raw materials and duties on products. While it may not always be possible to guarantee that duties, royalties and taxes will remain the same throughout the project life, some information can be gained from past experience; there should be information on all major changes in the past.

In addition, an important aspect of any project will be the economic climate of the project country and, if they are different, its markets and its suppliers. Information should be provided on inflation, unemployment, trade balances in the project location and, if these are different, in the countries supplying raw materials and in the prospective markets. Both present situation as well as past experience and forecasts of the future will be required.

Other areas in which background information ought to be provided are population, population growth, level of schooling, linguistic abilities, availability of education labour, skilled labour, trained staff, graduates, restrictions on expatriate staff, labour laws, employment cost and social charges, quality of social services and need for company contributions in kind or in cash are similarly significant and religious and cultural taboos, dietary rules, tribal and national customs which might have a bearing on employment or staff training should also be reported.

Although all the above-mentioned aspects of project background are important in regard to all industrial and mining and some infrastructural developments, there

are certain project-related points which are of particular importance where natural gas is to be produced, liquefied or exported. Essentially, it is the openness of the authorities to accept outside investment and the absence of blocking powers by individuals, organisations, government departments, etc. Strict adherence to contracts, even if conditions change, is very important, and an honest administration with little or no corruption is desirable.

Markets

Again intensive studies of available outlets for the output of a project are an essential prerequisite for valid decisions about finance for all projects. However, the low degree of flexibility of natural gas projects and the close links, both physical and commercial, which exist between producer and consumer make it even more important for gas project financiers to be informed in some detail about intended sales, year by year, and if there are substantial variations, season by season for both the initial or build-up, phase, the plateau phase and sometimes even the run-down of the project.

Since natural gas markets do not stand on their own but depend on climatic conditions, industrial progress, general energy use, living standards and political decisions such as acceptance or otherwise of nuclear energy and energy import policies, any useful market study must deal with these aspects. It will further address itself, in the case of industrial sales, to the prospects for consumers' products such as electricity, fertilisers, chemicals, non-ferrous metals, etc. Admittedly, a detailed follow-up of all downstream activities would be quite impracticable; but there should be at least qualitative indications whether non-contract customers are likely to succeed and will continue purchasing natural gas. Especially, where the entire production of a plant or production train is earmarked for export, or destined for a limited number of customers, it will be desirable to have detailed information on customers' markets and their likely development.

Since it is hardly ever possible to tie-up all production by long-term contracts and most contracts will have to have at least some, sometimes a great deal of flexibility, to allow for changes in the economic climate, etc., a good market study should also provide information on alternative outlets to those envisaged and indications on how these would affect profitability.

Pricing

Any assessment of the success or otherwise of a project will always be based on price forecasts. These may be underpinned by contracts, provisional or definitive, by studies of energy availability and future energy prices, and by assessments of our own and competitive production costs. Nevertheless, this is the area in which most project evaluations by producers, consultants and financiers have gone wrong, the main reason being the unsatisfactory nature of energy and especially of natural gas supply and demand balances both locally and worldwide. The familiar scenario is the OPEC story and the inability of this group of producers to match world energy demand with production by acting as swing producers, and the consequent fluctuations in energy prices. Inevitably, these have affected gas prices in the past and will no doubt do so in future. A basic assumption which, I believe,

must be fed into any price forecast is that there will be no shortage of natural gas in any of the major consumption areas of the world in the near- and medium-term future.

Minimum price forecasts must therefore be based on the cost of supply of the cheapest competitor: not an easy thing to do since much of the international gas trade is politically inspired and the FSU, for example, until quite recently, had no equivalent of a cost of supply as perceived by the capitalist world. Fortunately, this situation is now changing and it may be more meaningful in future to talk of the cost of piping FSU (Former Soviet Union) gas to Europe or moving FSU LNG to Japan.

There are, however, other problems associated with gas pricing: most international gas prices are denominated in US$, and the US currency has been fluctuating widely in the recent past. If a gas producer has large outgoings in currencies other than the US$, e.g. to repay Yen denominated loans, he may find this difficult if his customers settle in US$. Information on the currencies involved in a project is therefore essential.

Technology

All financiers, promoters and consultants connected with any project will wish to be reassured concerning the technical feasibility and future performance of the equipment installed. While they generally do not have in-house expertise to assess submissions by consultants or promoters they will often resort to cross-checking using more than one expert. In the case of natural gas and oil projects, there will be need for further information on field geology, geophysics and oil and gas producibility.

The most common controversy appears to arise over fundamental technology. Clearly, it is important that the new project should be at least as cost-effective as its nearest competitor. Equally clearly the technology used should be reliable and, therefore, tested in other projects. How one can reconcile these contradictions appear to be the great problem.

Normally, all major projects are tendered for competitively by a number of contractors; one way of dealing with this situation is to give the contractors submitting tenders, who should know more about project technology than anyone else involved, a free hand to propose whatever technology they feel they can guarantee as to reliability and lowest operating cost. Comparative feasibility studies by consultants can help and should be referred to, or submitted together, with requests for finance.

Since construction has to be paid for all along the line, time spent building is important and must be reduced to a minimum. One can offer premiums for early completion and/or penalise a contractor; but since the promoter almost invariably has to provide certain services, or else co-ordinate the work of several contractors, it is not always easy to allocate blame for delays. The alternative of employing a managing contractor with full responsibility can be expensive.

In any case, information about technology, tendering procedures, type of contract chosen, reasons, premiums and penalties, anticipated time of construction

and first production, etc., should be submitted. Financiers for obvious reasons will also want to know about the site and its ownership, size of buildings and type and capacity of machinery, utilities and of sites, storage tanks, control equipment, road and rail links, jetties, harbours and other transport infrastructure, office and staff facilities, and unused land surrounding the plant.

Staff and Management

Although it is difficult to assess the quality of staff and management for a prospective project, especially since recruiting may only proceed apace with construction, it is one of the essential tasks of consultants and project evaluation staff to report on training and ability of future operators and management. This may involve forming a judgement on company culture and employee relations, in overseas locations on the need for expatriate experts or management, pre-completion and in-house training, and regarding the involvement of future managers with the construction phase of the plant. All this is much easier if the project in question only involves expansion of existing facilities, and most of the operators and managers of the new plant will be selected from existing staff. Speed of expansion and experience gained in the old plant by staff to be transferred, as well as their involvement in the planning and construction of the new facilities will have to be assessed, in addition to their training, capability, character, etc.

Operating Costs

Ultimate economic performance of the plant will depend on input cost, output realisation and operating costs of the new facility. Only the latter is under the direct control of the operator and information must, therefore, be provided on how it can be contained.

It is generally possible to reduce manning by increasing investment in control and automation; however, there is a limit beyond which it might be unsafe to operate the plant without human supervision. Proposals for plant manning should originate with the contractor, and information should be provided on numbers, responsibility and quality of operators. Together with salary levels and adding in the necessary overheads, this should result in total personnel costs for the plant.

Other operating costs, such as purchased utilities—cost of utilities produced in the plant must be treated the same as production costs—are more accurately predictable in the short term, but they can change over the lifetime of the plant. Nonetheless, the best possible forecasts of the consumption and cost of power, fresh water, cooling water, high and low pressure steam, chemicals catalysts, lubricants and process fuel must be made.

If plant depreciation is considered part of the operating cost, this should be specifically mentioned and it must be subtracted for in the calculation of some of the performance ratios which shall be discussed later.

The Finance Plan

Any assessment of a project would be incomplete if it did not provide detailed information on how the entire plant expenditure is to be provided. A complete finance plan will take into account both prospective dates of payments to

contractors and equipment suppliers, salaries of staff connected with the project, and initial supplies of raw materials, utilities and other project requirements, i.e. the working capital. Financial charges, such as interest and capital repayments, must also be provided for. Any production during the initial phase should be credited.

Financial coverage must be assured at any moment in the form of promoter's equality, government subsidies, suppliers' credits, loans from financial institutions including international bodies such as the World Bank, the Regional Investment Banks, Government Agencies, etc., and finally grants from governments and others. Since it is generally uneconomic to pay interest on loans before the money is required, it will be ideal, though not always practicable, to accept loan disbursements precisely on the date when the funds are required.

While financiers of a project are of course familiar with the cash flow (disbursements of loan instalments, repayment of capital and interest) of their own contribution, they will wish to know about other loans and other means of financing the project. They will want to check that cash availability at any moment during construction and initial running of the project matches cash calls; in other words that the project realisation will not run into liquidity problems.

Project Profitability

It could be argued that once all the qualitative and numerical information mentioned above has been collected, it should be left to outside assessors of the project to form their own opinion. In actual fact, it is normal practice to evaluate the data further in the form, first of all, of an overall cash balance over the life of the project.

As mentioned before, cash flow analysis is required during construction and initial operation; it is quite easy to extend our cash flow forecast further, given sufficient accurate information on raw material costs, product prices, sales volumes and changes in operating cost. The remaining items of information such as interest and loan repayments are known; they are part and parcel of loan conditions, and while some of the grants received may be conditional, it is not too difficult to make more than one forecast, one with the other without the grant.

Obviously, there will always be some degree of uncertainty about forecasts, particularly in the long term, and it is advisable to allow for such variations in the evaluation. Almost inevitably, there will be more than one "case", usually three, i.e. a best, a worst and a most probable case. To take away from the subjectivity of such assessments, it is common practice to introduce variations at a lower level, e.g. regarding future gas prices, royalties, salary levels and to rely on computer programmes to integrate these into the forecast. Further refinement calls for a probability analysis of the variants which have been introduced, with the result that one obtains not one or three but a whole range of profitability. Little can be gained by such sophistication.

A further item of information which must be fed into the cash flow forecast is taxation. This may be a simple income or corporation tax percentage of net revenue; it may be a graded tax or it may start at some given level of income, throughput or return on the promoter's investment. Again this should be clearly visible at the time of the evaluation.

Having established inward and outward cash flows for each year, or shorter period, it remains to estimate the active life of the project; this can be a function of

available reserves of raw materials, particularly for oil and gas projects, or it may be the result of obsolescence. Obsolescence may be due to wear and tear and the increasingly uneconomic nature of maintenance and repairs; it may also be due to technical changes and the arrival on the scene of more advanced and more economic processes. The residual value of the project will vary accordingly. But it should be evaluated and inserted as a cash inward flow item in the last year of the project.

One of the most common criteria of profitability is the internal return on investment (IRI), and it is customary to calculate it as the interest rate at which the sum of the discounted cash flows for all the periods equals zero. Initial cash flows, during construction and before plateau operation of the project, will inevitably be negative and only at a later stage will there be positive net cash flows; it follows that there must always be an IRR value which equals zero.

A mathematically simpler correlation is the average net income of the project expressed as a percentage of the total investment on completion of the project. In this instance, but not in the case of the other ratios, it is common practice to allow for annual depreciation and/or depletion. This is often referred to as the classical rate of return (ROR).

Other correlations such as break-even time can be based on either straight or discounted cash flows and are the number of periods—years, half years, months—which have to elapse before the sum of the cash flows, initially negative, later positive, equals zero.

One can assess a project on the basis of its overall net cash flow, sometimes over and above a basic return. In other words, all cash flows are first discounted to a standard interest rate, and only the excess over and above this income is considered a net profit. The sum of these residual values or the total residual cash flows can be discounted forward so as to obtain a net final value, or backward to produce a net present value.

While the above ratios are the most common correlations used to compare projects—they can be refined to distinguish between financial and economic profitability, returns before and after tax—there are numerous other measures of the success or otherwise of a project. While it is merely advisable to compare alternative proposals using several different ratios—they do not always tell the same story, it is absolutely essential to compare like with like, and to be quite sure that published data or results of somebody else's calculations are based not only on the same basic information but that the method of calculation is also the same.

The principles discussed are illustrated by means of the example below. It will be noted that the calculations of cash flows and returns were carried out by means of a PC; this has been done to avoid arithmetical errors but, for a short-lived project as the one treated here, it is not at all essential. In fact, the calculation will only take very slightly longer if performed on a pocket calculator with facilities for evaluating financial data.

Investment for the project is spread over years 1 and 2; subsequently, it has been assumed production rate will be uniform over years 3–9, and at the end of that year, the project will be wound up, the site recovered, and the equipment assigned a scrap value.

Finance for the project was to be in the shape of equity, supplier credits, bank loans and grants. Interest rates were assumed to be 10 % throughout, and repayments were spread evenly over 5 years. A substantial grant or subsidy towards the investment was to be provided, on the one hand, while normal tax and royalty payments were to be made to the host government, on the other.

The promoter's cash flow consists of: Sales Proceeds + Loans + Credits + Grants + Tax Credits—Loan Repayments—Credit Repayments—Interest—Overheads—Operating Cost—Raw Materials—Royalties—Taxes—Investment—Equity

Project cash flow, on the other hand, is made up of: Sales Proceeds + Grants + Tax Credits—Overheads—Operating Cost—Royalties—Taxes—Investment. In other words, the promoter's cash flow is less than all financial terms.

A further difference between "project" and "promoter" cash flows is that the promoter contributes the equity, i.e. for him it is an outflow. As far as the project is concerned, equity is just one of the sources of finance, the same as any other.

The difference between economic and financial terms is that the former disregards taxes, tax repayments and grants.

Finally, to calculate a classical average return over the life of the project, we consider only total investment, and set against it total net income (not deducting interest and loan repayments), deducting however capital depreciation.

It will be noted that quite different returns result underlining the need for a precise definition of the parameter applied. The second set of returns is based on economic rather than financial evaluation. As discussed in the text previously, payments such as grants, royalties, taxes and other direct government contributions by a project have little bearing on its inherent profitability; an evaluation which does not take them into account is, therefore, more informative in regard to its desirability for the community as a whole.

The evaluation of project proposals, in the realm of natural gas and petroleum, is basically no different from that of other projects, although project background is generally of much greater importance. This is mainly due to the essentially political nature of most projects, to their location in sensitive areas, frequently in less developed countries, and to the large investments involved, which often require action by non-nationals both as promoters and as financiers.

It is also essential in the case of many gas and oil projects to take geological data, and especially field information, into account since in many instances these will be the critical factors which determine ultimate profitability. It has frequently been argued that this additional degree of uncertainty of oil and gas investments justifies a higher return and establishes a higher cut-off point for project feasibility.

8.1.1.6 Investment Appraisal

The objective of Investment Appraisal

Companies do not stand still but continually make changes. Most changes involve new capital expenditure on fixed assets. Such capital expenditure may be for replacement assets or for more efficient assets or in connection with new products,

production methods, factories, branches, etc. The acquisition of an entire business or company takeover situations can also be regarded as capital expenditure. Clearly, decisions on such capital expenditure should not be made without thought and the processes by which possible capital expenditure is considered is called capital budgeting or capital investment appraisal. There are various techniques for considering such capital expenditure.

Capital Expenditure
Capital expenditure is expenditure on fixed assets. Fixed assets are long-lived items such as land, buildings, plant and machinery and vehicles. These are called tangible assets. In this, the acquisition of intangible assets such as goodwill, patents and trademarks will be included. Fixed assets can be acquired singly as in the purchase of a replacement lorry or lathe or the construction of a factory extension. They can also be acquired as a set as in the setting up of a new branch, a new production line or a whole business. Capital expenditure often involves very large sums and is non-routine, so it is very important to have a system to deal with projected capital expenditure. Such systems are called investment appraisal or capital budgeting. The important thing is to distinguish between capital expenditure and revenue expenditure.

Revenue Expenditure
Revenue expenditure is spending on the supply and manufacture of goods and the provision of services charged in the profit and loss account in the accounting period in which they are consumed. In most companies, this is controlled by systems of budgetary control perhaps with standard costing and variance analysis. Methods of appraising capital expenditure

Payback Period
A simple way of appraising projected capital investment is to use payback. This is simply considering the length of time, which is required for a stream of cash receipts flowing from the investment to recover the initial investment. Suppose a replacement machine was being considered. The cost (after deducting the proceeds of the sale of the old machine) will be say £5,000. The new machine will cost less to run, and the annual savings are estimated at £1,500 a year. Thus:

$$\text{The payback period is } 3\frac{1}{3} \text{ years } (3 \times \pounds 1,500 + 1/3 \times \pounds 1,500).$$

For management, knowing the payback period is useful in deciding whether or not to go ahead with an investment and deciding between possible alternatives. Some companies have a hurdle such that investments are not accepted unless the payback period is say less than 3 or 4 years. Points about payback period include:

- Cash flows are considered, not accounting flows.
- Future cash flows are not equivalent to current cash flows as current cash can be invested to earn interest. £1,000 today can be invested at say 10 % so that in one year's time, it is equivalent to £1,100. This important idea is not recognised by payback.

- Payback period prefers short-term investments rather than investments which take a long time to pay off. Short termism is a criticism that has often been levelled at British management
- In a time of rapid technological change, payback period has some obvious benefit
- Projects that payback quickly may present less risk, and payback is sometimes used as a proxy for the riskiness of the project.

Discounted cash flow and Profitability Index

This is a more sophisticated approach to capital investment appraisal but is not so simple to understand. The important concept is that money has a time value, for example if you have £1,000 today (the present time), then it can be invested to earn say 10 % interest. This means that its value grows as shown in Table 8.1.

Table 8.1 is easy to understand. However, we can see the process in a different way. Table 8.1 can say that the £1,210 after 2 years is equivalent in value to the £1,000 now. Also say that the present (=now) value of £1,210 receivable in 2 years time is £1,000. If the interest rate was say 15 % the figures could be illustrated differently as typified in Table 8.2.

So, the present value of £1,322.50 2 years in the future at 15 % rate of interest is £1,000. It is possible to calculate the present value of any future sum, given its date and a rate of interest. Using the following relationship:

The formula can also be applied to estimate the net present value (NPV)

$$\begin{aligned} \text{NPV} &= \text{future sum} \\ &= (1+r)^n \end{aligned}$$

where r is the rate of interest.

This has to be expressed as a percentage, i.e. if the rate is 16 %, then $r = 0.16$, and n = the number of years. If the figures already calculated and taken as £1,322.50 at 15 % 2 years in the future. Therefore, the NPV can be:

$$\text{NPY} = \frac{13,225}{(1.15)^2} = £10,000$$

Table 8.1 Example of cash flow

	£	
Now	1,000	
Interest year 1	100	1,100
Value after 1 year interest year 2	(10 % of £1,100)	1,200
Value after 2 years		1,210

Table 8.2 Example of interest rate

	£
Interest year-1	1,000
Value after 1 year	1,150
Interest year-2	172.50
Value after 2 years	1,322.50

Table 8.3 Tabulated form of calculating discount factor

Year	Interest rate			
1	10 %	12 %	14 %	16 %
2	0.91	0.89	0.88	0.86
3	0.83	0.80	0.77	0.74
4	0.68	0.64	0.59	0.55

Table 8.4 Example of internal rate of interest

Year	1	2	3
	£'000	£'000	£'000
Cash flows	20	30	40
Discount factor at 14 %	0.88	0.40	0.27
Present values	18	23	27

Alternatively, tabulated form which is shown in Table 8.3.

From Table 8.3, the discount factor for a sum 3 years away at 14 % is 0.67.

So if we need the present value of £4,800 3 years away at 14 % needed, then the answer is £4,800 × 0.67 = £3,216 or as 0.67 is a rounded figure which should be, perhaps, specified as £3,200. Using the related formula for present value gives £3,239.86.

Internal Rate of Return

This is a similar but more difficult alternative to NPV. It actually presents a number of conceptual and practical problems not addressed here. As a result, it is not very commonly used in business. The objective of this approach is to see what rate of interest a proposed project will return. Suppose that a company is considering a joint venture project abroad, which will involve an outlay now (year 0) of £65,000 and the expected returns over the 3-year life of the project are: Year 1 = £20,000, Year 2 = £30,000 and year 3 = £40,000. Thus, what rate of return is given by this project? An alternative statement of the problem is:

What rate of discount will discount the cash flows in years 1, 2 and 3 back to the outlay in year zero?

This can only be found by trial and error. Starting with 14 % interest rate, therefore the internal rate of return will be shown in Table 8.4.

Present values in total are £68,000. This is more than the outlay, so we can say the project pays more than 14 %. Lets try 15 %, this gives figures of £17,000 + £ 23,000 + £26,000 = £66,000 (= £17,000 + £23,000 +£26,000). This is approximately correct and we can say the project will pay about 15 %. The company may have a rule that all projects, which pay above the cost of capital of 12 %, are acceptable and then this project would be considered to be acceptable. Some points about internal rate of return are:

- Most textbooks calculate the rate exactly (e.g. 14.63 %). However, the forecast cash flows are estimates with a large margin for error, and it is not really possible to specify the rate of return accurately. In fact, it may give management a spurious confidence in the accuracy of the forecasts.

Table 8.5 Example of expected return

Year	1	2	3	4	5
(in £0.000)	5	6	8	6	5

- The calculations can be tedious and in practice are carried out on a computer.
- The higher the internal rate of return in relation to the cost of capital, the less risky the project appears to be.

Accounting Rate of Return

The net present value and internal rate of return approaches use cash flows. A disadvantage of this approach is that measurements of cash flows are not made or are reported so there is often no feedback on the success of the forecasts made. As measurement and reporting are done of accounting flows, some advocate appraising projects by using the accounting rate of return. Suppose the company is proposing to open a new branch and expected initial outlays are £30,000. Thus, expected returns are as shown in Table 8.5.

The returns shown in Table 8.5 are after depreciation, thus the average return is £30,000/5 = £6,000. Also, the accounting rate of return is $\frac{£6,000}{£3,000} \times 100 = 20\,\%$

An alternative is to take the average capital employed which, assuming that the investment will be steadily depreciated to nil over the 5 years is

$$\frac{£30,000}{2} = £15,000$$

In this case, the accounting rate of return will be:

$$\frac{£6,000}{£15,000} \times 100 = 40\,\%$$

Points about this method include:

- It is in effect a rate of return on capital employed.
- The accounting flows are subject to the vagaries of accounting measurement including different depreciation methods.
- The method is not intellectually rigorous, and there are no universally accepted methods of applying it.
- Its use is not recommended.

Practical Investment Appraisal

In practise, there is agreement that some approach to formal capital investment appraisal is desirable if only to clarify thinking. For example, management may decide that they fancy a company aeroplane but to do so without careful estimates of costs, revenues, savings, alternatives, etc., may well lead to a bad decision. However, the intellectually rigorous methods of appraisal (net present value against internal rate of return) do have drawbacks including the fact that management do not always understand them.

The simple methods (payback and accounting rate of return) are easier to understand but lack rigour. Many companies adopt a combination of methods.

This may mean net present value as a first hurdle and payback as a second hurdle. Also in practice proposals may pass the tests and be rejected by management and many fail the tests but are adopted anyway.

8.1.2 The Project Team

Roles and Responsibilities and the need for teamwork

The project manager is responsible for the organisation of the management team and ensuring that all aspects of the project are considered in the business and operational sense. A project team should of course be lead by the most experienced person and he/she should organise the team and its reporting regime in achieving the overall objective. The team should include heads of functions responsible for exploration, design, procurement, construction, commissioning, legal and finance, Sales and marketing, personnel and safety all working in a coordinated manner to their own plans which should be coordinated into a macro-project plan all to an agreed timescale. Specifications should be written which include all details of exploration including estimated expenditure associated with the exploration for gas along with a detailed plan.

Engineering functions to provide design, details and estimates of the process and plant design, construction, commissioning and training of personnel, process and maintenance, and procurement of all plant required to supply gas to a sales outlet. Value analysis and hazard and operability analysis should be carried out along with the provision of a reinstatement plan.

The legal and financial sections should bring together all aspects of finance associated with the project and presented within the total financial plan. Estimates and records of cash flow are required for the whole project up to, and including, the stage when the project runs into profit. This may also be responsible for establishing and securing the supply of capital funding for the project. It will provide advice on, and draw up, all legal documents and contracts associated with the project and agree the legal basis of the project.

Marketing

The Sales and marketing team should be formed and be active as the soon as the source of gas is discovered, and quantity and quality of the source gas are known. This is essential for the analysis of the business case for sourcing capital finance for the project. It should provide all market analysis, strategy and planning including analysis of risk associated with the sale of the gas

Human resources should research and advise on all aspects of government, social and educational influences including requirements for training, staffing and personnel management including the labour laws of the countries associated with the project.

The preceding paragraph covers a huge amount of ground and is not aimed at being definitive but merely to provide a structured example and to argue the case for planning of all functions as a basis for the business of natural gas.

Project Planning

Gas projects have to be successful because we live in a highly competitive, minimum capital environment where there is no world price for natural gas. Therefore, we have to be totally capital productive and natural gas and other associated capital projects have to be completed within time and budget. To achieve this, the project must be well managed, organised and controlled. Planning is an essential component of project control and offers to the business and project manager the perfect mechanism for coordination and communication within the project. For example: a natural gas production plant, whether large or small needs a variety of parts, vessels, pumps, pipework, valves, instruments, steelwork, etc., to be in the right place at the right time. This means that the various pieces of equipment must be designed, ordered, manufactured and delivered in the correct sequence to enable the plant to be constructed in an orderly way and within budget. Planning, therefore, is about the work and the ways in which it is done. We need to know when various jobs and activities can or must be performed.

The start and completion dates of each activity in the project should be accurately forecasted and managed to achieve them. Materials and equipment should be delivered to site just in time, and the money made available to pay for them. These and many other such activities involve an analysis of both the durations of individual activities and manpower and equipment involved. It follows then that project planning is essential to the provision of capital money so that a successful project can be completed to time and specification.

The investor will want to know the amount of risk involved and consequently will expect a return on invested capital commensurate with that level of risk. This of course coupled with the fact that gas projects are often in third world countries, which may have volatile governments, and borders, and not a penny of the returns on capital invested will be realised, until the project is complete, and gas is sold to a complete market, at its optimum planned rate and price. Therefore, planning of the market demand rates and pricing of the gas forms a fundamental of the financial plan and projected production date become crucial to the case for investment.

One of today's problems is that there are a number of huge sources of gas which are in various stages of development to supply a world market which is dominated by oil. Therefore, whereas governments know that there is an increasing demand for gas as an alternative supply of energy, there is an oversupply potential in the world and due to the absence of a world price for gas, each project has to compete with other fuels in its own market.

Planning Philosophy Commitment

If planning is to be successful, it must have the basic ingredient, commitment. All parties that contribute to the plan must have complete commitment to it at all times. Commitment has to come from a senior level, and in the case of capital projects, from the project manager. He must set the scene and ensure that at progress review meetings, all members of the project team can demonstrate, they are working to the plan.

All members of the project team must be encouraged to identify precise activities of work, nothing airy fairy can be entertained and realistic estimates of durations of jobs and manpower must be made.

Everyone involved must resist at all costs the optimistic and equally the pessimistic. Although both are bad practice, the pessimistic estimate is possibly the lesser of the two evils, inasmuch as the completion date for the project will probably be nearer the mark when you consider that slippages will always occur during the lifetime of the project. All information must always be issued to the planner, be it good or bad, so that the effect on the project can be assessed and corrective actions taken if necessary.

Finally, commitment to plans can never be realised if the plans themselves are too complex. Plans can be made to be as large and as comprehensive as the project to be undertaken. Logic networks can cope with as many or as few activities as the degree of control requires. Plans can be drawn to cover even nuts, bolts, joint rings and every small item you think are essential to the building of the plant. However, be aware that production of vast amounts of activities requires a large amount of updating and there is also the risk of introducing spurious logic. This will lead to the plan having to be modified regularly in an attempt to reflect apparent change. People can quickly lose confidence in plans that change too regularly. Once priorities are set, it is unproductive to change them often. It is also unrealistic, for effort cannot be transferred backwards and forwards from jobs. The continuity of the project will inevitably suffer. Also people will not understand the plans and as a consequence will not use them. Gone is the commitment. Only the necessary amount of detail in the plan consistent with control is needed. Concentrate on fundamental activities which experience has taught are usually the one's which will cause problems.

The Elements of the Natural Gas Project Plan

The overall plan may be made up of a number of planned activities: the exploration and analysis, the financial, the sales and marketing, the engineering and design, the construction, the commissioning, the future development, the decommissioning, the production and the safety and operational and any other plans which may contribute to the success of the project. These in turn will have many sub-plans which may make up the overall master plan. Therefore, in order for planning to be effective, the environment must be correct. The project team must be set up as early as possible. And the project procedures must exist for all disciplines and must be clearly understood. Planning will be of little help in an ill-disciplined environment. All parties must understand their commitments to the project plan. Key components which effect project organisation, and hence, progress must also be established as soon as is practicable. Key dates for things, such as information flow, are obvious early contenders, but perhaps the most important element of project organisation is the definition of the physical components of the project. Once a detailed plan is established, and individual jobs specified and timed, job methods for each event can be written and associated risk assessed. These can be timed to suite progress, and there is learning from successive jobs

which can be applied to later events. This highlights where events can impinge or affect access and equipment and eliminate any risks resulting from jobs being carried out in parallel. One such example is where large pressure vessels have to be installed within a structure and structural members may have to be left out to permit access for the installation of the vessel. Frequently, this is overlooked and will result in having to remove structural items from a building which could put the structural integrity of the building at risk. Not only does this compromise the safety of the building but it can add cost to the project. Hence, it is critical to the success of a project to employ planning engineers.

The Involvement of the Planning Engineer

Usually, the need for planning on a capital project is identified by the project manager. However, sometimes, if the project is of high-strategic importance, the need for planning may be initiated by the Board of Directors. To obtain maximum benefit from planning, the planning engineer must be appointed at the earliest moment in the projects life. The degree of influence which can be brought to bear upon the direction in which effort should be exerted is highest when the project is at its birth. The point at which the planning engineer can be meaningfully involved varies, but in general, the earlier the better.

An example of early involvement of planning was where the planning engineer was involved during the research stage of the project when R&D requested help in determining which avenues of research should and could be pursued within the constraints of the project timetable. This involvement may start anything from around 2/3 years before capital sanction and could well involve laboratory or pilot plant effort to decide the final route for the process.

Planning Opinion

Unfortunately, planning can be viewed as a waste of time and effort. Especially when employing smaller contractors on smaller jobs. But large jobs and projects are made up of a larger number of smaller jobs and can utilise thousands of smaller contractors. Some will argue that plans can go wrong, and job duration, and manpower times, are almost impossible to estimate. They are costly to produce and result in a plan that may be inaccurate and raise expectation.

The counter argument is, of course, that planning is a fundamental discipline, and an organisational tool which can aid decision-making. It is a system of monitoring and, yes, will raise expectation that the project will be completed on time. Planning is the key to successful *Business and Project Management.*

A project should have an overall macro-plan which is fed by a number of detailed plans. It will provide a base for costing, scheduling, estimation of personnel requirements, and forecasting cash flow. Target setting and management are key to its achievement and the reporting of progress fundamental to the business case.

The critical path method (CPM) is a project modelling technique developed in the late 1950s by Morgan R. Walker of DuPont and James E. Kelley, Jr. of Remington Rand [5].

CPM is commonly used with all forms of projects, including construction, aerospace and defence, software development, research projects, product

development, engineering, and plant maintenance, among others. Any project with interdependent activities can apply this method of mathematical analysis. Although the original CPM programme and approach is no longer used, the term is generally applied to any approach used to analyse a project network logic diagram. The essential technique for using CPM is to construct a model of the project that includes the following:

(i) A list of all activities required to complete the project (typically categorised within a work breakdown structure),
(ii) The time (duration) that each activity will take to completion, and
(iii) The dependencies between the activities.

Using these values, CPM calculates the longest path of planned activities to the end of the project, and the earliest and latest that each activity can start and finish without making the project longer. This process determines which activities are "critical" (i.e. on the longest path) and which have "total float" (i.e. can be delayed without making the project longer). In Project Management, a critical path is the sequence of project network activities which add up to the longest overall duration. This determines the shortest time possible to complete the project. Any delay of an activity on the critical path directly impacts the planned project completion date (i.e. there is no float on the critical path). A project can have several, parallel, near critical paths. An additional parallel path through the network with the total duration shorter than the critical path is called a sub-critical or non-critical path.

Originally, the CPM considered only logical dependencies between terminal elements. Since then, it has been expanded to allow for the inclusion of resources related to each activity, through processes called activity-based resource assignments and resource levelling. A resource-levelled schedule may include delays due to resource bottlenecks (i.e. unavailability of a resource at the required time) and may cause a previously shorter path to become the longest or most "resource critical" path.

A related concept is called the critical chain, which attempts to protect activity and project durations from unforeseen delays due to resource constraints. Since project schedules change on a regular basis, CPM allows continuous monitoring of the schedule, allows the project manager to track the critical activities and alerts the project manager to the possibility that non-critical activities may be delayed beyond their total float, thus creating a new critical path and delaying project completion. In addition, the method can easily incorporate the concepts of stochastic predictions, using the programme evaluation and review technique (PERT) and event chain methodology. Currently, there are several software solutions available in industry that uses the CPM method of scheduling. A schedule generated using critical path techniques often is not realised precisely, as estimations are used to calculate times. If one mistake is made, the results of the analysis may change. This could cause an upset in the implementation of a project if the estimates are blindly believed, and if changes are not addressed promptly. However, the structure of critical path analysis is such that the variance from the original schedule caused by any change can be measured, and its impact either ameliorated or adjusted for.

Indeed, an important element of project post-mortem analysis is the as built critical path, which analyses the specific causes and impacts of changes between the planned schedule and eventual schedule as actually implemented. Whereas the original critical path diagrams were produced manually and may have taken days to rework, computerised systems are easy to change and the affects of changes observed immediately. The resulting scheduling systems can be invaluable in computing cost advantages, etc.

Sales* and *Market Planning

It is essential that the market for a new gas field be established so that the business case can be made. This is fundamental to estimating the return on capital employed and involving business partners in funding the project. Therefore, a sales and marketing plan should be produced as soon as the volume of recoverable reserves and quality of gas are known. The plan must define the potential markets and magnitude of possible sales of gas and the opportunity to compete with and replace alternative fuels. This might be an existing gas market with potential growth in sales represented by conversion from existing fuels. The location of potential gas sales must be defined along with the cost of utilising alternative fuels. Unlike oil, gas has no world price structure and so any new business will require to compete in each individual market. The cost of delivery and marketing of natural gas must be estimated along with any marketing and distribution costs. SWOT analysis should be carried out for all market sectors and competing fuels and supply companies. Any other factors arising should be included in producing the marketing plan.

Definition of*Potential Markets*

The market must be defined in terms of load factor and volume characteristics for gas use. The affects of the geographical location and climate should be assessed along with types of customer, e.g. domestic, power, industrial, etc.

The magnitude of sales of existing and alternative fuels should be quantified. Total energy demand should be assessed identifying which fuel is used and why for each market and subsequently the potential for market penetration/substitution by gas. Potential growth in sales should be identified by:

- An analysis of potential customer's business and markets
- Price/volume relationships
- Government policy (e.g. environmental)
- Development/construction projects
- Location of potential "gas sales"

Existing and future sales must identify

- location of ownership transfer (who is responsible?).
- Location of metres
- Identification of "the customer"
- Location of potential sale relative to the existing supply system.
- Compatibility of supply and demand characteristics.
- Cost of alternative fuels.
- The delivered cost to customer's site.

- Take into account all associated costs.
- A comparison of energy cost at burner tip.
- Premium uses or added value for natural gas.

Cost of delivery and marketing

- Cost of field, transmission and distribution facilities.
- Sales, billing, customer service costs (can be very high).
- General marketing and market entry costs.

SWOT analysis carried out identifying

- Strengths
- Weaknesses
- Opportunities
- Threats

For the product, own company and competitors, other factors to be addressed are as follows:

- Effect of government/international action/policy.
- Privatisation/introduction of competition
- Effect of changes of ownership of any transmission and distribution structure.

The Market Planning Process

- Define start point (Where are we now?)
- Where do we want to be and when? (define objectives).
- How do we get there? (strategy and tactics)
- Compare actual with target. (Are we where we expected to be?)
- Measure, reformulate objectives, and incorporate experience into new plan.

The market plan should taken into account (see Fig. 8.6)

- Sales forecasts
- Budgets (investment, Sales and promotion, customer service, operations, etc.).
- Cash flow forecasts.
- Finance

8.1.2.1 Economic Efficiency

There are with 3 types or areas where efficiency can be measured from a business aspect. They are referred to as *productive,allocative and dynamic efficiency.*

Productive efficiency requires that production of an output be undertaken using the least cost technology. It is the method or technique of producing goods or services, which provides the least cost of production. The implication of productive efficiency to pipeline is that the productive system has to be working at full capacity, and consequently, the energy per unit cost transported will be low. Therefore,

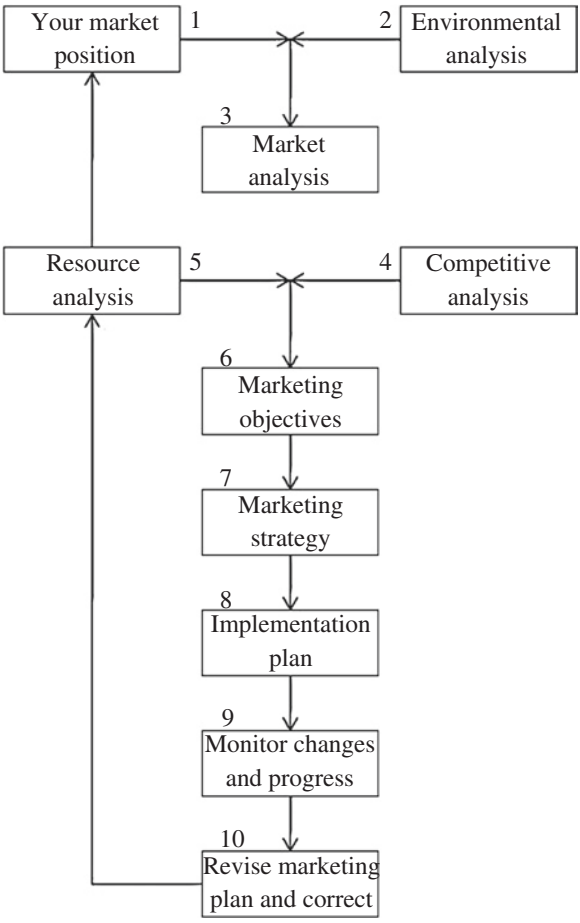


Fig. 8.6 Steps in marketing planning process

productive efficiency can easily be achieved in open access or common carriage pipeline access systems because both systems permit and encourage competition.

Allocative efficiency requires that the market price for a product is equal to the cost of marginal unit of production. It is to ensure that the prices match the cost of supplying a product or service. The implication of allocative efficiency to a pipeline is that the price of gas should be able to match the landed cost of gas at any location. This can only be achieved if the gas market is liberalised and competition is allowed in trading of gas and the pipeline operates a free access or common carriage system.

Dynamic efficiency requires that new investment should be made when the cost of producing from the original marginal source of production exceeds the cost of making the new investment. This means that lower priced producers/service providers replacing higher priced producers/service providers over time. There is no dynamic efficiency in a pipeline business because if you can make investment in

pipeline infrastructure, the pipeline will remain in good condition underground operating without requiring any huge operating/maintenance cost that will justify any further investment. There will be no time within the useful life of the pipeline where the maintenance/operational costs of the pipeline would be higher than the costs of replacing the entire pipeline with a new one. An energy regulator will optimise overall economic efficiency of a gas pipeline company operating in a mature gas market by assessing the competitive capacity utilisation of the pipeline that will result in a low tariff, which will be reflected in the end-user price of the gas.

8.1.2.2 LNG and the Gas Chain

The gas chain in the context of natural gas delivered to user by pipeline was discussed earlier in this chapter. However, let us consider whether this model is appropriate for the LNG gas chain.

The key factors of the gas chain model described in Sect. 8.1.1.1 can be applied here to LNG. Figure 8.7 shows the gas chain modified to produce LNG as an end product.

From Fig. 8.7, this can be seen that the pipeline is replaced by the liquefaction process with transport by ship to a receiving terminal where it goes through the process of regasification and then into the distribution system.

If we now compare the LNG model shown in Fig. 8.8 with the requirements/definition of the gas chain as we delivered by pipeline, this complies with the key factors and elements except that it is necessary to have a complete process from wellhead to the customer. Both the liquefaction and regasification plants can be designed and operated for continuous operation which leaves the business implication of transporting the LNG by ship.

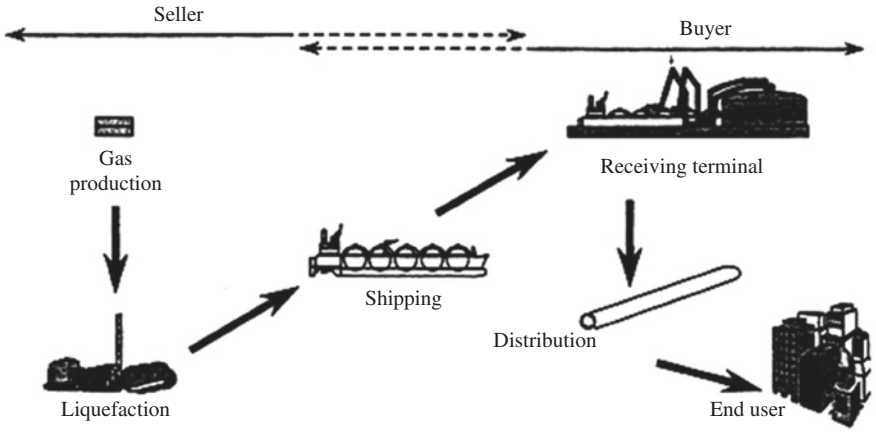


Fig. 8.7 Typical mode of "gas chain" for LNG production

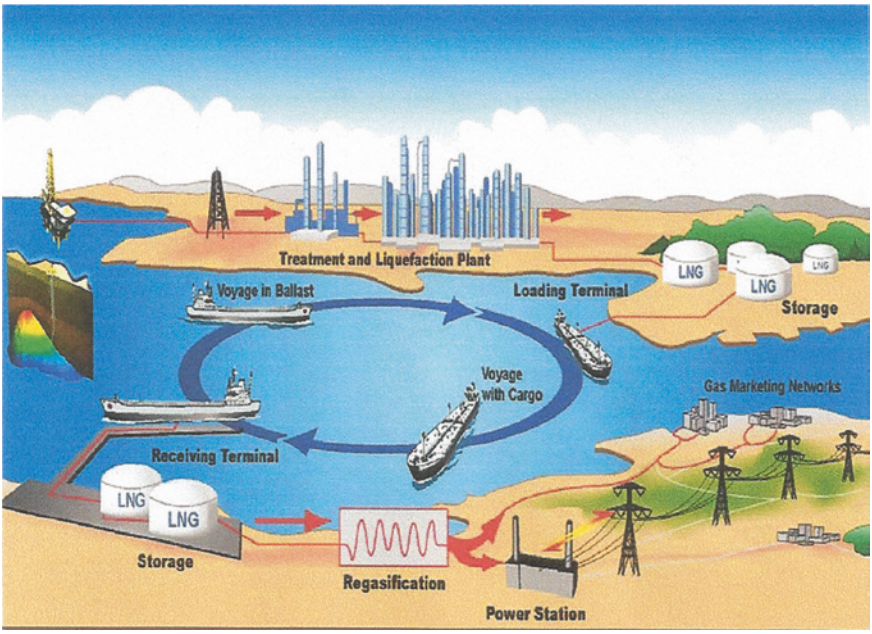


Fig. 8.8 LNG and the gas chain [6]

Therefore to determine the number of carriers needed, the following list should be considered:

The capacity of each ship
The average service speed
The distance between LNG plant and receiving terminals.
LNG plant production rates.
Number of berths
Priorities at the receiving terminals
Risk of delays to carriers.
Numbers of ships available/required

Also added to the following key parameters defining the market are as follows:

Plateau volumes
Price
What build-up is required
Take or pay levels
Annual and or seasonal flexibility
Duration of contracts
Timing of start-up
Quality of gas required

From this data, one can ensure parity with pipeline operation. Furthermore, one must also ensure that operation and maintenance regimes are in place for continuous operation so that the chain becomes complete. This in practise is achievable, and therefore, it could be argued that the gas chain model is suitable for the LNG chain. The strongest argument is of course its use as a business model in making the case for securing capital investment.

8.1.2.3 Interfuel Competition and Pricing to Industrial Consumers

Although most industrial consumers in Europe, the Far East and North America can now expect to pay a price for gas which is more or less related to the value of crude oil, heavy fuel oil (particularly for interruptible supplies), gasoil, LPG or coal, the extent of such "linkage" varies widely, as the following examples will show:

- In Australia, prices are determined by competing fuels (off-peak electricity in some cases) and varying widely by state.
- Austria: built up from import prices (Russia, Germany and Norway); close to gasoil parity for small industry and to low sulphur fuel oil (LSFO) for large industry, with few interruptible contracts.
- Belgium: government-controlled firm prices escalated monthly against the average frontier price of gas and a proportion of changes in indices of wages and prices; large interruptible linked to LSFO. Some coal links in power station contracts.
- Canada: because of more production than export capacity, prices in Alberta were historically below US levels. Now, the situation has been reversed, and Canadian prices have risen to American levels.
- Denmark: gas prices changed monthly and linked to local and Rotterdam oil quotations (gasoil and/or LSFO).
- Finland: no interruptible and related to price of gas imported from Russia, which tracks Dutch export price; generally below LSFO parity, partly because of high-oil taxes.
- France: firm prices changed quarterly and reflecting oil price movements as well as Gaz de France's purchase prices and distribution costs; interruptible is linked either to fuel oil (mostly 2 % sulphur) or to a discount from firm price.
- Germany: standard contracts for smaller users and freely negotiated prices for larger consumers reflecting competing fuels, a convenience premium for gas, the investment needed to switch from gas and, for smaller consumers, the budget needs of municipally owned distributors; increasing proportion of gasoil prices in indexation formulae, allegedly for environmental reasons; less interruptible than in other European countries.
- Italy: monthly indexation (for firm) of the commodity charge against a varying basket (depending on volume) of high-sulphur fuel oil (HSFO), LSFO and gasoil in Rotterdam and Italy over the previous 12 months, with six-monthly

indexation of the demand charge against a proportion of changes in indices of wages and prices; interruptible linked to LSFO in Mediterranean and changed monthly.

- Japan: because of the link between import prices of LNG and crude oil, gas sales to industrial consumers other than power stations are mainly for direct heat rather than steam-raising, with prices more or less reflecting market levels for gasoil, kerosene or liquefied petroleum gases, plus a premium for convenience, etc. Now increasing use of coal in power stations in addition to LNG, fuel oil and nuclear.
- Netherlands: quarterly link with LSFO at Rotterdam for all but the first 170,000 m^3/year (gasoil-linked), with decreasing supplements by volume tranche. Transportation tariffs above 10 million m^3/year.
- Spain: monthly links with LSFO and HSFO (firm and interruptible).
- Sweden: related to prices of competing fuels (usually 0.4 % sulphur fuel oil or gasoil and sometimes coal).
- Switzerland: largely interruptible and usually related to gasoil for environmental reasons.
- U.K.: complete gas-to-gas competition: British gas trading (now part of Centrica) has lost over 90 % of firm market over 2,500 MMBTU/year, over 80 % of interruptible sector (including power stations) and about 60 % of market between 250 and 2,500 MMBTU per year. Prices are averages, while those for new contracts are 40–50 % higher, i.e. now similar to levels on the continent. Prices are now ultimately determined by spot market levels, which are however now influenced by trade through the interconnector, thus re-establishing a link with oil.
- U.S.: the most direct link is with spot gas prices offered by producers rather than competing fuels, although some power station contracts are tied to coal and heavy fuel oil, if dual- or triple-firing alternatives are in place. Gas prices to industrial consumers are normally changed less frequently than those of oil, with the result that the former can sometimes be out of line for several weeks if oil markets are in a volatile state. Gas shortages have, in the past caused spot and city-gate prices to rise well above fuel oil, particularly in Los Angeles. Coal prices are still well below those of fuel oil and interruptible gas.
- Interruptible gas prices (where available) are once again broadly in line with those of heavy fuel oil; in the UK., this is true for new gas contracts, mainly because of the interconnector's link with pricing on the continent.
- Consumers of firm gas normally pay prices which are above those of heavy fuel oil but which are now well below gasoil parity, partly through time-lags and partly because of high-excise duties on oil.
- Coal remains cheaper than either gas or fuel oil.

It should however be remembered that market comparisons reflect only the gross calorific values of the fuel and do not take account of:

- The generally greater convenience of gas
- Its lower sulphur content

- Preheating costs for heavy fuel oil
- Storage costs for fuel oil and gasoil
- Investment required to switch from one fuel to another

As liberalisation on the continent of Europe approaches, some oil–gas links may break, as was the case until 1999 in Great Britain, but this will take time. Governments and government policies change in a world which is now more environmentally aware and against a background of reducing availability of oil at the right price. This will affect the business of natural gas throughout the world and introduce much needed competition in the business of energy supply. Those involved in the gas business should use the information in this section as a base for exploring the effect of changes in government policies and involvement in gas projects.

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Chapter 9
Innovation and Research

9.1 Introduction

The subject of innovation was firstly realised by the economist Joseph Schumpeter [1] in the 1930s. He classified the concept of innovation into five distinct categories which were also summarised by Goffin and Mitchell [2] as follows:

- ***Introduction of a good to consumer or new consumer with better quality***
- ***New methods of production in industry, maybe used in other sector too***
- ***The opening of new market***
- ***The use of new source of supply***
- ***New forms of competition, leading to the restructuring of the sector***

9.1.1 Definition

Today, nearly 1,000 hardcover listed books by Amazon alone cover the topic of management. There are so many conferences and technology-based platforms that they have regularly become the conduit of the promises to capture the innovations. The subject also eternally captivates the business leaders in inspiring innovation through improving and changing products, process, services, technologies and/or business models into the corresponding organisations.

In spite of numerous theories, tools and approaches in innovation management, there is *no clear and available methodology* that can assist the business leaders in improving innovation performance. It should also be noted that *R/D is the source of the innovation* and driver of many generated ideas. Organisations must NOT fall into the trap that produces sophisticated products without understanding the

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requirement of the market. The effectiveness of the innovation can ONLY be viable when every functional area of the organisation contributes to innovation. These areas are as follows:

- Marketing: *Identifying customer needs*
- R/D: *source of innovation leading to new products and services*
- Operations: *Key roles in innovation*
- Finance and accounting: *Essential support for RoI*
- HR Management: *Hiring, developing and motivating people*
- Outside Resources: *Open innovation—Industry or universities or consultants.*

In the context of what was explained above, organisations should consider the definition of innovation [3] in its broadest sense as follows:

> ***"Collation of knowledge, experience and ideas converted into new or improved products processes, business models and services in gaining competitive advantage or strategic leverage in the market".***

Hence, *innovation* generally refers to renewing, changing or creating processes that are more effective products or ways of doing things. An innovative process should answer the three basic questions as typified in Fig. 9.1.

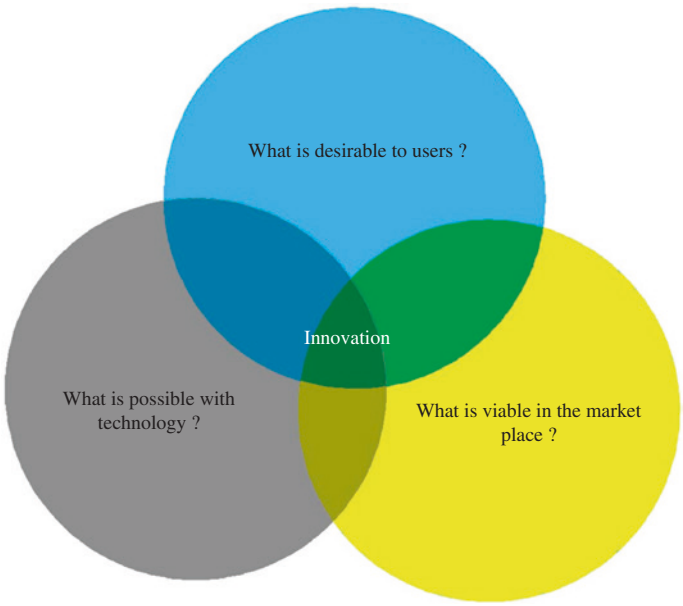


Fig. 9.1 Typical innovation matrix

9.1.2 Phases of Innovation for a Gas Company

The authors throughout have endeavoured to reclaim a suitable innovation model which can suit the organisational structure and the activities of a given gas company. The authors believe that the *5-Steps Innovation Framework* model that was originated by Goffin and Mitchel [2] can fit well within a given gas company ecosystem, bearing in mind that Goffin and Mitchell named this model as "*The Innovation Pentathlon Framework"*.

9.1.2.1 Gas Company 5-Steps Innovation Framework

Most projects either start from an initial idea from a potential customer or are generated inside the organisation. Such ideas usually exceed the available resources or funding for them to turn into active projects. An initial screening process is essential to avoid overcommitting resources and failure to deliver the desired results. Moreover, the business case is a key controlling document as the fundamental charter for the project and must be subject to regular revision and updating. The question should also be asked, i.e.

How often do organisations put effort and money into a project to find at the end that no one needs the outcome? Selection is the decision of the senior management who need to be given enough data to make the decision.

Two approaches are used:

(i) qualitative method
(ii) quantitative method

For the *qualitative* method, any project selection process requires the manager to ask some basic questions:

- Will the proposed project maximise profits
- Will the proposed project open new markets, consolidate market position
- Will the project maximise utilisation of existing resources (i.e. people)
- Will the project maximise the utilisation of existing manufacturing capacity
- Will it boost company image
- Will it increase risk faced by the company
- Is it within the company's current skill and experience

The *quantitative* method focuses on financial data to support the case. The information needed includes the following:

- Payback period
- Return on investment
- Rate of return

Innovation should progress through a number of phases before it becomes commercially viable. In 1988, "the creative gap" by Majaro [4] was proposed as an

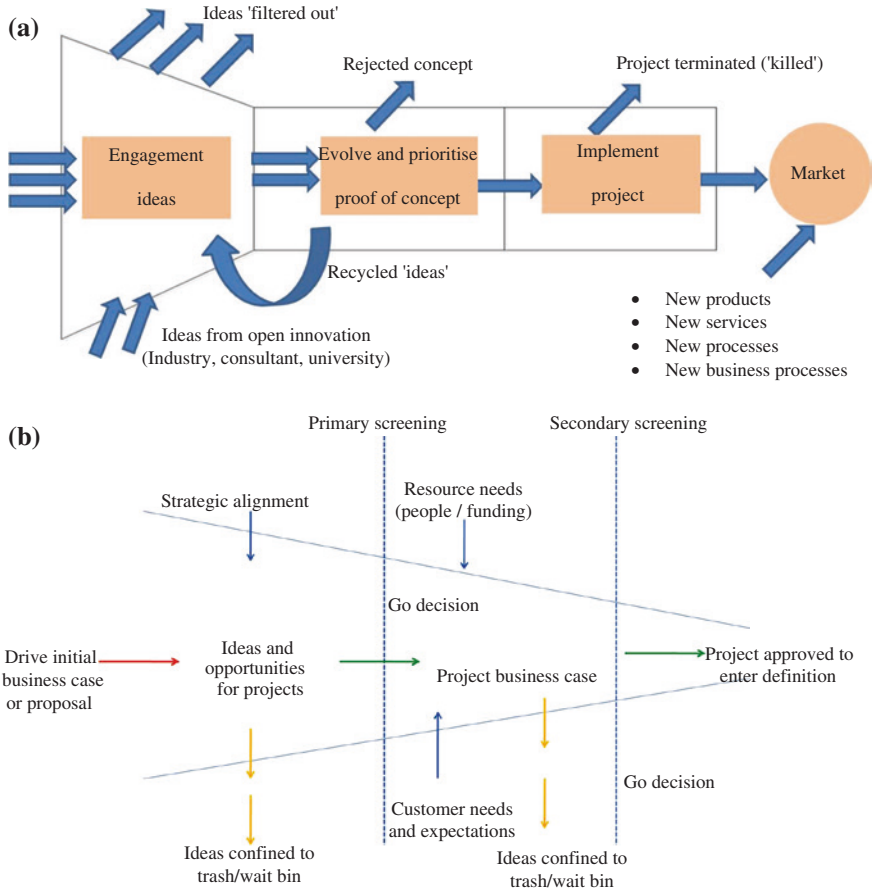


Fig. 9.2 **a** Typical phases of an innovation using the development of funnel model and **b** alternative illustration of the funnel model for project selection

acceptable process for an innovation. It was initially developed for the pharmaceutical industry, and afterwards used by many other organisations. Figure 9.2 shows the typical phases of innovation with a funnel of ideas generated with some ideas filtered out, if not viable, whereas others progress further and are developed into concepts.

As can be seen in Fig. 9.2a, certain concepts can be rejected or recycled. Furthermore, the project may be cancelled or modified during implementation. It has been experienced that the funnel model could lead to the overlapping of the different phases and with common problems which necessitates an iteration process. The 5-Steps (or Pentathlon) innovation framework, shown in Fig. 9.3, however, provides the essential linkage between the innovation strategy, people and the organisation. This is fundamentally the two additional elements which should

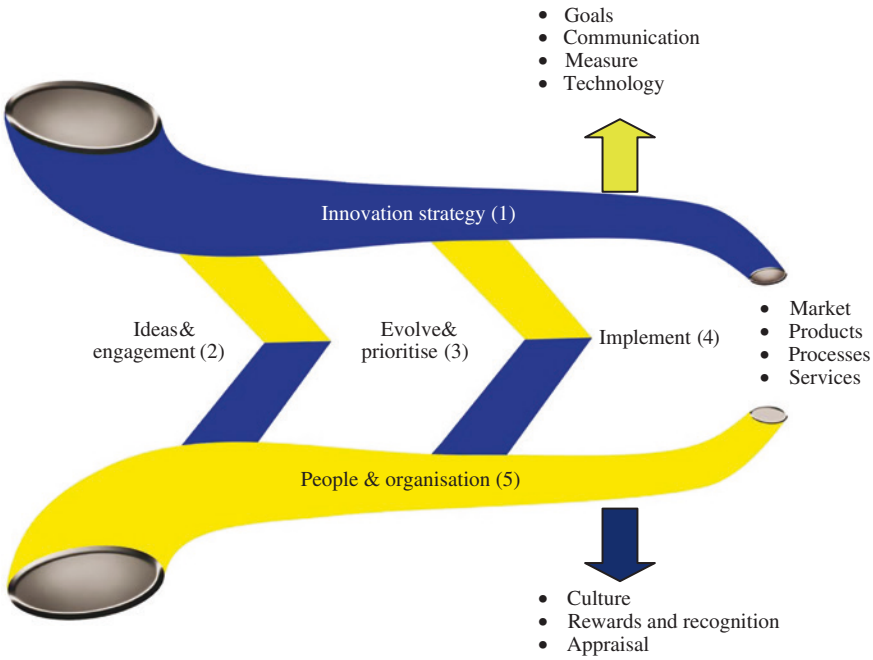


Fig. 9.3 Proposed gas company 5-steps innovation framework (option-I)

be built onto the funnel model. All the related elements of the 5-Steps Innovation have been echoed onto it in order to provide a better insight into the proposed model.

A further proposed innovation model in which gas company can adopt is typified in Fig. 9.4.

9.1.2.2 Proposed Model for Gas Company Innovation Strategy

However, an alternative model shown in Fig. 9.4 can also be adopted while maintaining the philosophy of the 5-steps innovation framework. Each step is numbered from 1 to 5 to signify the 5-step. Human resources play an active part in a gas industry's innovation process. Human beings perform the technology and implementation processes. They ask the following questions:

- Are ideas quickly developed into new products/service concepts?
- Are creativity techniques and workshops used?
- Do human resource policies support a "culture of innovation" through stimulating a creative, problem-solving working environment?

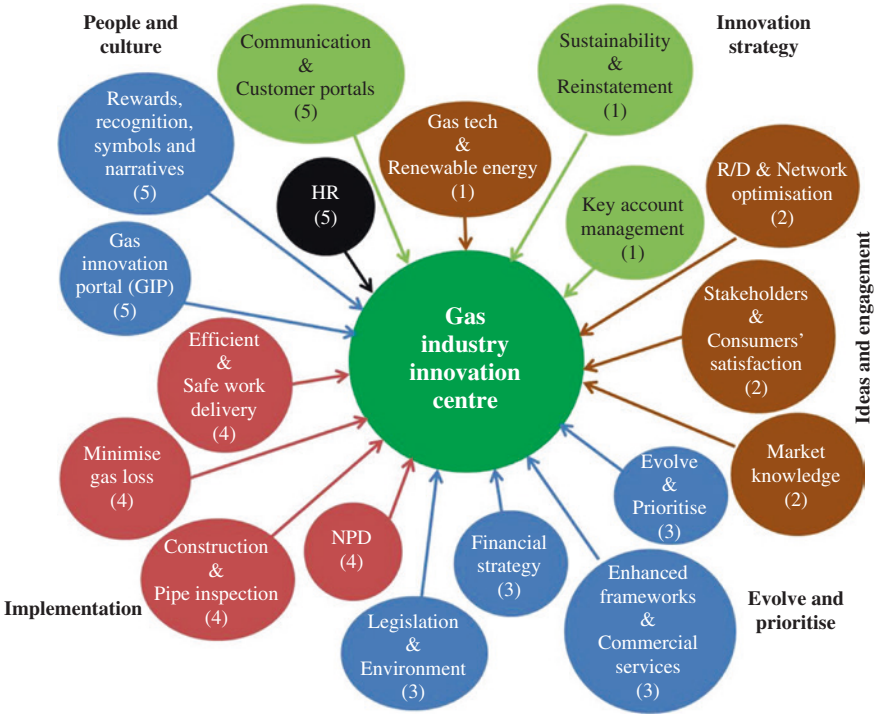


Fig. 9.4 Proposed gas industry model (option-II)

Humans provide the answers to these questions. Hence, this model introduced the human factor into the innovative process.

In the following Sections, the *5-Steps Innovation Framework* shown in Fig. 9.3 will be utilised and described in formulating an innovation management ecosystem within a given gas company.

It is important to realise that each area of "5-Steps" SHOULD NOT independently be chosen to improve the performance of the company. Each area is complex in its own right with complex interaction between them and sometimes difficult to predict. Gas companies must continue analysing the benefits that the companies can offer to their customers/stakeholders and maintain them. Companies should also ask how their value is to be maintained and enhanced into the future. The typical example of this analysis could be using the current TQM tools in linking the product features to customer satisfaction. Kano [5] suggested three categories in linking to customer satisfaction (i) basic feature (must be there), (ii) performance features (creating concept, using market advantage) and (iii) excitement feature (provides unexpected value to customer). It is also the responsibility of the management to ensure that the contribution of *Continuous*

Improvement (or as expressed by Japanese "Kaizen") is fully communicated to the entire workforce of the gas company.

9.2 Gas Company Innovation Strategy (Step-1)

On the passage of building strategy within the gas company, some major steps are worthy of note:

1. Identification of where innovation is needed;
2. Determination of the strand of the business that requires the innovation (i.e. products, processes, systems, services and company and culture);
3. The extent to which the change is necessary;
4. Uncovering the best way to protect the innovation from competition by management.

The critical success factors are a leadership-defined strategy for growing and sustaining the innovation ecosystem. Moreover, there should be a commitment to making innovation a core strategy for the gas company growth and renewal, a pledge of long-term funding and a time horizon that accommodates learning and adjustment to processes.

> ***"Innovation is a multidisciplinary group process which must be supported top-down and driven bottom-up"*** [3].

In developing an innovation strategy, a gas company must do the following:

- Treat the innovation strategy as a continuous process event **NOT** as a single-point improvement.
- Have a full understanding of the limits of their existing products, services, systems, processes and company and culture and to fully realise her competences.
- Fully realise the gap between what is desired for now is a requirement of the future projections.
- Have a clear understanding that the limitations do not put the organisation at a competitive disadvantage.
- Protect your innovation using intellectual property rights (IPR) tools.
- Choose and time the innovations with care.
- Use proper road mapping and prepare participation of all involved.
- Recognise as competencies in innovation within the gas company mature, the company is ready to move from products, processes, services to business process.
- Provide efficient training and recruitment together with HR planning.

9.3 Ideas and Engagement (Step-2)

In the engagement state, it is important that the gas company ***senior leadership team*** inspire the innovation vision and set an expectation on the structure and mission of the innovation intended. It is also essential that in this process gas company engages employers, and partners as well as customer requirements. Furthermore, innovation requires creativity, the application of knowledge and protection of the resulting ideas and engagement. It should be borne in mind that creativity is NOT simply for *totally new ideas*, it is also *solving known problems* as well as the resulting creativity from substantial research, either individual or group works. This is a second element of the *5-steps innovation framework* (see Fig. 9.3). This element should not simply be based on critical market research which relies mainly on direct questioning rather than more market intelligence techniques which involves observation and indirect questioning that enable the company to identify breakthrough in order to revitalise existing markets or create new ones. According to Goffin and Mitchell, the "hidden needs" of the customer are difficult to capture through traditional market research.

The gas company must continue in creating a culture of creativity and innovation in which innovation project teams can progress, which in fact is people management. Employees of a company must feel empowered and recognised by their company for doing great things and motivate others to do great work.

> **The workforce must be encouraged to *participate in their innovation efforts, develop innovation portal, make the progress transparent, demonstrate that their input is valued, review ideas regularly and reward people for successful efforts*.**

There are various techniques available for innovation, such as the classic ***brainstorming*** in which a group of people are asked to describe any ideas that come to mind as solutions to a problem or as opportunities for new products, services of business and then by recording the ideas and reflecting these back to the group to simulate further discussion. For example, why certain customers are happy with a product and some are satisfied.

Another technique could be ***"left–right brain alterations"*** [2], as the left side of the brain is capturing analytical aspects and the right side controls our abilities for intuition, art and visualisation. For example, the improvement of a service product which can be from left side of the brain asking a question such as what is the core product? How quickly is delivered, what are the key performance indicators. By contrast, the right side lead us to ask, how does the customer perceive our service, how do they feel about the service, etc.

Gas companies should give close consideration in developing ***an ideation portal***, the site invites ideas from everyone with some set parameters, depending

on the product, and each time a new challenge is issued, the site is refreshed. Otherwise, it becomes like the old suggestion box in a corporate lobby and the usage tends to decline. Moreover, every idea has to undergo top-management scrutiny and if approved, seed funding is available, resources are quickly identified, and high-level priorities are set.

It is also worthwhile to note that a high percentage of new product developments do not fully reach their business objectives. It is thus pertinent that company products, services, systems and process innovation are aligned to the customer input through effective market research.

There are a number approaches to identify the customer problems and their requirements. Goffin and Mitchell [2] proposed a suggested approach which is summarised below. Although it seems to be generic in places, a gas company could align this to their needs.

- **Survey research**: direct and open questions may have low response rate, and also questionnaires should not be designed poorly
- **Focus groups**: two-way mirror research with market researches; hidden needs could surface. Particular individual can dominate the discussion and prevent inexperience moderators.
- **Structured interview**: users and customers undergo a structured interview; powerful, hidden needs could emerge. Experience interviewees needed with 5-6 different products and services
- **Empathic design**: putting product designers in shoes of users with systematic observation. Need an expert to conduct this. Vast amount of qualitative data may be generated which needs effective analysis.
- **Lead users**: extreme user needs in the current market. Workshop could be set up for creative discussions. Difficult to identify the lead users, time-consuming and outside the normal working hours.
- **Tester session**: customers presented with the prototype of the product. Observe the reaction of the customers. May require expensive virtual prototyping equipment.
- **Co-joint analysis**: customer chooses their preferences of the products or service. Enable pricing model, product identification. Limitation on accuracy of the findings due to having artificial nature.

Protection of ideas also enables companies to protect and exploit their knowledge. There are many kinds of intellectual property rights (IPR) such as patents, copyrights, trademarks and design right. Gas companies should continue to establish who is responsible for filing costs and maintenance fees if licensing patent or starting a joint venture.

In summary, for *Ideas and Engagement* step, gas company should do the following:

- Must create environment that stimulates creativity and generation of ideas.

- Develop ideation portal, encourage and recognise and actively collect ideas such as gas innovative portal (GIP).
- Devise an effective customer input through market research
- Protect ideas from competitors.

9.4 Portfolio Management Office: *Evolve and Priorities* (Step-3)

9.4.1 Introduction

For a gas company to create a viable solution for effectively prioritising and selecting projects, it must first define a framework within which it will operate. This framework should be created within the concept of ***PMOPortfolio Management Office*** [3]. The office is responsible for dealing with all of the company's projects as a single portfolio. Figure 9.5 typifies when the PMO could be placed within the gas industry management structure. While the projects within the portfolio are still managed as individual initiatives, from the senior management perspective, the office must be viewed as a collective means of establishing the company's goal. It may be for many projects that they never see the light of day, but choosing the right project allows the company to succeed and thrive, while the wrong project can take the company into a wrong path without recovery. Figure 9.6 shows a typical gas company management and PMO.

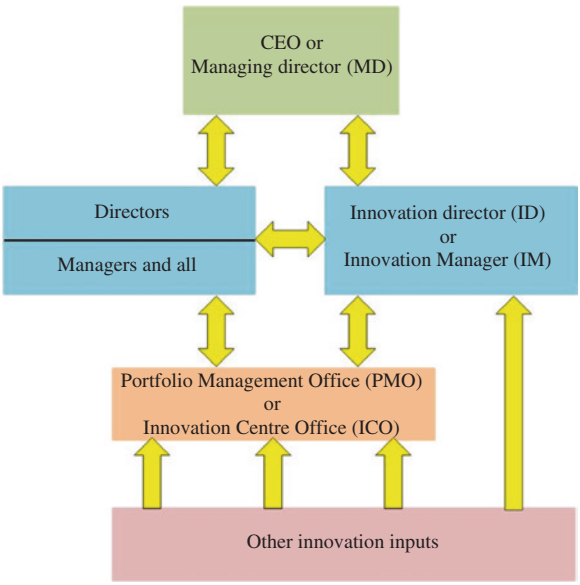


Fig. 9.5 Gas company PMO

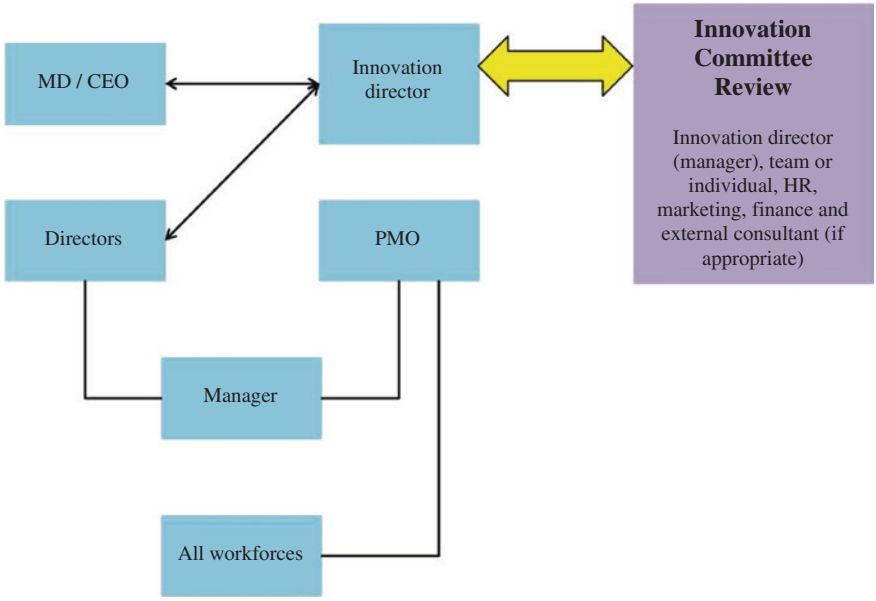


Fig. 9.6 Proposed innovation management organogram with innovation director (or manager) and PMO

PMO enables gas companies to address some fundamental questions like the following:

(i) Are we doing the right projects
(ii) Are we confident about the chosen projects and will they be delivered on time and meet the company goals and
(iii) Last but not least, for the projects that we have completed, are we consciously changing how the organisation operates in order to realise the value the company expected.

Established and operational PMO will also provide the gas company with the following:

(i) Maximum value strategically and on ROI;
(ii) Improved visibility in terms of strategic goals of the company, less political decision-making and a shared understanding of what the company is doing and why;
(iii) Improved effectiveness with regard to an ongoing project performance, benefits and achievement of the company change;
(iv) Reduced execution risk in terms of investment risk, project with high success, allocation of resources and prioritisation of initiatives, control and monitoring at portfolio level and suitability of environment for successful delivery of the project;

(v) Improved governance and control;
(vi) Improved integration between portfolio, programmes and project management;
(vii) Enhanced stakeholder and/or customer engagement;

PMO main responsibilities are categorised as follows:

i. Assess current state and requirements of gas company in achieving their innovation objectives.
ii. Develop vision and outline to support, guide and enable gas company to achieve their innovation objectives.
iii. Develop roadmap and business case for implementation.

9.4.2 Project Portfolio Management

Project portfolio management (PPM) is not only making the right decision, it is also about reviewing those decisions regularly, changing them if appropriate and assisting staff and colleagues to understand and accept the changes. There are ***financial methods and non-financial methods*** by which projects can be valued. Gas companies should continue to improve their process by reviewing the adopted methodology that is currently in place which can be aligned to PPM. In the following, selecting the right project and ranking the project for selection and managing a portfolio is briefly described.

In general, ***financial method*** can include the net present value (NPV), sensitivity analysis, decision tree analysis for multistage projects or Monte Carlo simulation. The financial method, however, could sometimes be misleading and takes a lot of effort, and since the projects are at the conceptual stage, it could result in a number of uncertainties which could make it cumbersome to develop a detailed business plan.

Non-financial method is based on a project *scoring system* and the method is well documented [6] and at the moment is the best technique that is available. It is better to use so-called *anchoring statements* in ensuring that all participants use the scales in a similar manner. Scoring system enables the team to think about all aspects of the project and not just the cost and technical feasibility. If the information is shared in an open way, the group will reach a consensus on each factor. Although one must avoid the dominance of an individual or senior members influence on the process. An example of anchoring statements is proposed by Goffin and Mitchell [2] which is interesting to consider. This is based on a scoring method (rating 1–10) based on criteria such as sales increase over 3 years, cost reduction, pricing, efficiency, customer impact and technology platform for growth. Although some are generic statements, it is, however, possible for a gas company to consider and revise the contents to suit their related project scoring. Similarly, scoring systems can be used in design processes and R/D projects but in a different fashion, relating more to the specifications and under the "wish" and "demand" category. It is important that the information is circulated to all not

"only a few" managers and when a meeting commences, everyone starts knowing all the facts.

Furthermore, in choosing and managing a portfolio, this should also be considered with respect to the following:

(a) ***Strategic alignment***: if a gas company's long-term aim, for example, is to move into a new technology or enter into a market or percentage of innovation investment directed towards the project or include strategic priorities as factors in the project scoring system.
(b) ***Time and resources***: completion time and resources allocated to projects should be realistically balanced against the number of projects to be delivered at any one time since the company has a limit to manage change at one time. On the other hand, there may be a good reason to launch some innovations together to maximise their impact.
(c) ***The risk/reward profile***: There are varieties of methods to assess risk. Data may be taken from NPV analysis or from a project scoring system. In assessing project risk, for example, by scoring (1–10), a gas company could consider dividing the project as follows:
    (i) ***Technical***: consider the novelty of the project, uncertainties of the technical issues, has feasibility demonstrated and knowledge of market for the product available.
    (ii) ***Commercial***: is the market ready to enter and do we have capability/experience and expertise in our structure.

In summary, it is important, as mentioned previously, that the selection process is as open and as objective as possible with no official timetable, just as required and leave no one with any disappointment when the project is cancelled; otherwise, it can create disillusionment. *It is after this process that the project is presented to senior management team.*
Innovation is risky, but the risk must be borne by the company not an individual. Furthermore, unsuccessful innovators should be rewarded for their efforts. The final stages of this assessment process, scoring, NPV and risk and reward is a fully evaluated business plan.

9.5 Implementation of Innovation Project (Step-4)

Gas company should create a viable solution for effectively prioritising and selecting projects which can define a framework within which it will operate. This framework should be created within the concept of PMO. The portfolio management office could be structured in a number of ways depending on its type and the organisation's requirement.

This is a final stage of the innovation funnel on the 5-steps innovation framework (see also Fig. 9.3) and should not be treated as a separate activity. After this,

ideas are collected and prioritisation decision made. Implementation of an idea into innovation must be treated as a project and building the project team, and executing the project is something of a unique experience while taking into account design, market potential and legal evaluation. After review is finalised, design documentation is prepared and project team collaboration workspace provided; given that final funding is approved, the project then commences. Good project management is an essential ingredient as well as people skills. The project manager assigned to an innovation project must also manage the reactions of customers and whoever has been affected. NPD is an obvious innovation project which demands learning and experimentation on the route to a novel result. Common sense prevails in which any innovation project, even at implementation stage, faces high levels of uncertainty and risks. It is thus important that *risk management techniques* are used as well as constant project reviews. What seems to be needed is project *risk analysis* in a more coherent manner:

(i) Incorporating failure mode and effect analysis [7], which are used for products and processes, could be used for projects. This can be done by scoring method from 1 to 10. This will provide better insight into the risk management processes which can lead to a reduction in costs and lead times in R and D projects for example.
(ii) Similarly, the project links to customer needs in which all elements of the design contribute satisfactorily to them. Again, use the *approaches to identify the customer problems and their requirements* and score them and link them to customer needs from 1 to 10. Other method is quality function deployment (QFD) as described by Don Clausing [8].
(iii) In terms of *services,* [2] is to get full understanding in terms of the following:

- The value of the services the company provides (customers pay and gain, needs, time spent by consumers to receive the value)
- The form and function of services (delivery, added value)
- The nature of the service experience (how competitor's service is perceived, what is negatively perceived with current services or otherwise)
- The outcomes from services (tangible and intangible, clarity, additional outcomes appreciated).

9.5.1 Implementation and PMO

Implementation of the innovation project should be managed through the PMO as previously described. Innovation must not be treated as a "one-off" event; it is rather a continuing need for PMO to manage the processes and project individually or a stream of projects which could occur simultaneously. Managers should also install a specific process for managing innovation projects in terms of, say, progress, prioritisation, project review with improve a communication and as a conduit for learning and improvement. There are various adopted methods, for example for NPD or managing R/D function, such as *stagegate* [9]. Basically, the

stagegate is based on giving final approval to move the project to the next phase and if not, the project is stopped or delayed until deficiencies are corrected; this is called a *phase review of stage gate*.

It should be noted that if during the phase-gate process project experienced delay and moving forward, then one should consider peer review or design review stage. This can be done by the design team through PMO by inviting a group of colleagues from outside of the project and asking them to carry out an in-depth review with the aim of picking up any problems that the team may have missed. The reviewing team do not take over the project or even share responsibility for the work. They only offer advice to the team which also creates a better teamworking environment.

Innovation could also require *collaboration* with other companies through joint projects or strategic partnership (i.e. university) and could provide a gateway for major investments in implementation which is the part of *5-steps innovation framework*.

9.5.2 Post-Project Review

Post-Project Review (PPR) is also another important element in the implementation process after the innovation project via PMO. The core purposes of PPR detailed by Rajegopal [3] are summarised in the following:

- ***Timing***: PPR immediately after the completion of the innovation project
- ***Scope and moderation***: no blame culture must be adhered to here and also led by moderator with experience who could be from PMO.
- ***Learning from success and failure***. This provides mechanisms in "lesson learnt".
- ***Review all aspects of the projects***, technical, problem-solving, process adopted and running project.
- ***Dissemination of the results*** on intranet under "lesson learnt" is meant to bring the problems faced at the start of any new project to remind people of the problems to avoid.
- ***Linking the learning into action through PMO*** to avoid repeating mistakes between each project.
- ***Efficiency and effectiveness of people*** working on a number of projects with graphical representation (efficiency against number of projects).

In summary, gas company must ensure that robust *innovation management* processes are in place through Innovation Manager or Director and PMO. Appraise and manage risk as well as managing the process through customer-focused and install control processes such as stagegate or improve and

align their current system to innovation projects. Avoid delays and queues in managing innovation projects and develop a competence in managing collaboration and closely conduct PPR while not jeopardising the gas company strategic position.

9.6 Creating a Culture of Innovation Within Gas Company (Step-5)

People and organisation is the fifth element of the "*5-steps innovation framework"*. Great innovations come from people at gas company, and to create an effective innovation culture, the company should generally consider the following points as also noted by Rajegopal [3]:

1. ***Growth and strategy is aligned***: Ensure growth platform is pursued and innovation strategy is fully determined in all areas of gas company activities (ideas and engagement, evolve and prioritise and implementation).
2. ***Creating a close cross-functional process through PMO*** for transparent decision criteria for assessment, decision-making, governance and balancing a mixture of projects
3. ***Letting market help gas company innovate***: reduce market research, release more products and let the market dictate which ones are worth major investment (although it is important to have full understanding of market and R/D *leading* to technology which subsequently *leads* to consumers and customers).
4. Inspire innovation as an organisational discipline.

There are a numerous collated articles and publications currently available that relate to building an innovation culture in organisations. Readers should consult these articles or similar for further details for example [10–12].

9.7 Proposed Innovation Process for Gas Industry

The gas industry should consider adopting a simple innovation management process. This could include the input and output innovation process in which the corresponding inputs comprised of resources, time and investment with allocation of measured revenue to innovation (i.e. 5 %) and with output (process, product services, etc.) measured against commercial impact of innovations as indicators.

In recent years, the present authors applied the 5-steps (or Pentathlon) to gas industry suppliers inspired by Goffin and Mitchell model with some additions in order to suit the business operation of the company. Figure 9.7 shows the model that was followed in embedding the innovation management with supplier gas industry (SGI).

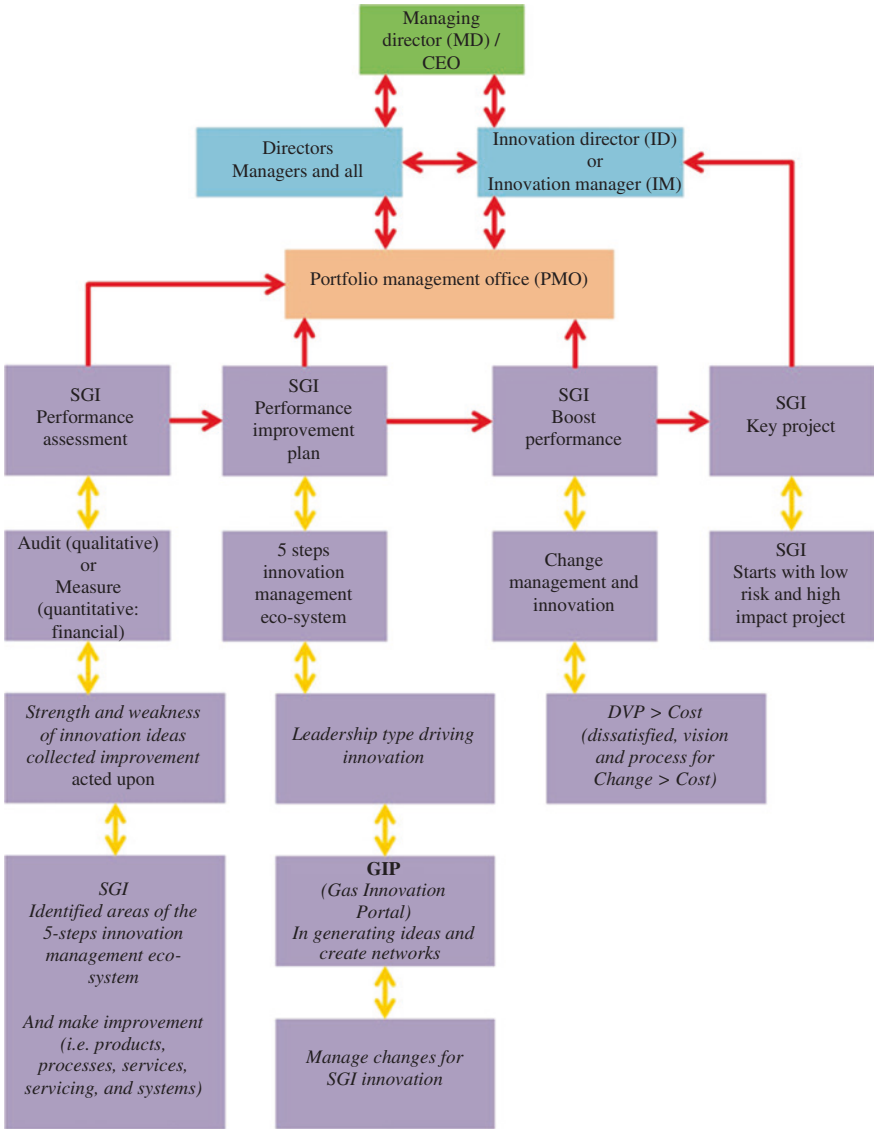


Fig. 9.7 Process for improving innovation performance for the SGI

9.7.1 Performance Assessment

To assess the performance of a SGI, an intangible audit is conducted with respect to the 5-steps innovation framework. Note that tangential (or financial) auditing could also be carried out using the past performance. In assessing the performance, qualitative methods were used rather than quantitative (financial) measure

Table 9.1 Target group activities

	Job specification	Number	Questionnaire
Tier 1 ***(MD)***	Managing director/CEO	1	Generic, performance, process and technical
Tier 2 ***(Senior Managers)***	Operations, sales finance	3	Generic, performance, process and technical
Tier 3 ***(Managers)***	Production/works, logistics/job expediting purchasing/MRP quality/ human resources marketing UK sales export sales gas sales service financial IT	11	Generic, process and technical
Tier 4 ***(Team Leaders)***	Assembly dept. stores/goods in premium dept. design/research gas and utilities dept. service engineer fabrication etc.	11	Generic, process and technical
Tier 5 ***(Operators)***	Gas operators	26	Generic

since it would be sometimes difficult to provide accurate financial forecasts in the rapidly changing energy environment.

In summary, the company is divided into various tiers in terms of their job description and the corresponding working activities as typified in Table 9.1.

The performance assessment intended to measure and improve assessment is grouped into the following subheadings:

- *Products/Service and Servicing:* final end-user products and services.
- *Processes/Systems:* the tools, methodologies and ways of working to deliver the products.
- *Company:* corporate development and company strategy.
- *Culture:* the company's culture and the workforce's ability to implement innovative ideas, people and investors in people.

9.7.1.1 Methodology, Results and Analysis of Innovation Audit

Three questionnaires outlined in this document are to be used to establish a benchmark of the level of innovation within a supplier gas industry. The questionnaires have been designed to encompass the five areas of the 5-steps innovation framework (ideas and engagement, evolve and prioritise, implementation, people and organisation and innovations strategy). The subsequent findings of the audit will be used to aid in the development of an innovation strategy throughout the organisation. The audit is to provide a comprehensive analysis of the level (and perceived level) of innovation throughout the company. As mentioned before, the target groups that will take part in the innovation audit are divided into five tiers (as shown in Table 9.1) and will cover all departments within the company. The

depth of questioning will be determined from the individual's position within the company. This strategy of questioning will provide both a hierarchical and role perspective of innovation throughout the organisation. Some of the questions highlighted in Table 9.1 are proposed by Rajegopal [3], but modified here to suit activities of the company.

Questionnaire Overview
The three questionnaires consist of the following:

(i) *Generic questionnaire* that will be applicable to all employees of the company and will be completed by the entire workforce *(all tiers).* The rationale of this questionnaire is to obtain the entire workforce's perspective of innovation within the company and to assess the feeling of innovation across departments (see Table 9.2).
(ii) *Performance audit questionnaire* this is a more searching questionnaire that quantifies financially innovation within the company. This will be completed by only *Tier 1* and *Tier 2* Management (see Company Organogram) (see Table 9.3).
(iii) *Process and Technical audit questionnaire* this will establish how the process of innovation is undertaken within the company, both at a hierarchical level and across the departments. The questionnaire will be completed by *Tiers 1, 2, 3 and 4*. The interviews for the process and technical audit will take the form of a structured interview following a set of clearly defined questions (see Table 9.4).

It should be noted that innovation covers the whole spectrum of operations within the company. For the purpose of this audit, these are defined as follows:

- *Products/Service and Servicing:* final end-user products and services as shown in Table 9.2.
- *Processes/Systems:* the tools, methodologies and ways of working to deliver the products.
- *Company:* corporate development and strategy of the company.
- *Culture:* the company's culture and the workforce's ability to implement innovative ideas, people and investors in people.

9.7.1.2 Analysis of the Company Audit: Performance Assessment

This section provides the SGI with an analysis of the data that were obtained through the innovation audit. The findings are to be used to aid in the formulation of an innovation strategy and to devise an action plan to improve the performance of the company (through "senior and middle managers").

The Generic Questionnaire was completed by 120 employees and covered all operations within the company (sales and marketing, engineering and R&D, operations, servicing, finance and IT) and across all tiers.

Table 9.5 provides an overview of the total respondents' answers to the survey to the five areas of innovation. The feedback is further broken down into the individual departments.

Table 9.2 Generic question (all tiers)

	NO	*Where appropriate answer as follows: (Yes or No) and/or either (Poor, Satisfactory or Good)*	Yes	No	Poor	Sat.	Good
(1) Ideas and Engagement.	**1**	Are ideas collected on a regular basis?					
	2	Are ideas quickly developed into new: products/processes/systems/services (ways of working)?					
	3	Are creativity techniques and workshops used to generate ideas?					
(2) Evolve and Prioritise	**4**	Are meetings regularly held (formal and informal) to discuss Company progress with regard to new ideas that will affect you?					
	5	Are you aware of how the performance of the product/processes/systems/culture/service (ways of working) affects Company performance?					
	6	Can you see a more efficient way of streamlining decision-making?					
	7	Does the responsibility of screening decisions lie too high within the organisation hierarchy?					
(3) Implementation	**8**	Are new products/processes/systems implemented relatively quickly and efficiently?					
	9	Do you see in your area of work a bottleneck?					
	10	Is this a bottleneck because too many innovations are attempted?					
	11	Are team meetings effectively used?					
(4) People and Innovation	**12**	Are you aware of all the aspects of innovation (not just product design), i.e. (products, processes, systems, corporate development, culture, services, servicing?)					
	13	Do you feel that the Company is innovative? (i.e., development of new products/services/processes/systems/culture)					
	14	Do you feel that you understand the importance of innovation?					
	15	Do you feel that you can become more involved in innovation?					
	16	Is there a blame culture if things go wrong?					
	17	Is innovation covered in your annual personal appraisal with your Line Manager?					
(5) Innovation Strategy	**18**	Are you aware of an innovation philosophy within the Company?					
	19	Are you aware of your role within the innovation strategy of the Company?					
	20	Does top Management spend sufficient time supporting all stages of innovation?					
	21	Do performance measures reflect the innovation strategy?					
	22	Are performance measures: simple, appropriate and valid?					
	23	Is there a Company framework to reward and recognise individual achievement in developing innovative ideas?					

Table 9.3 Performance audit questionnaire (Tiers 1 and 2)

(1) What percentage of your revenue was spent on R&D?
(2) How many new innovative products and services did your Company launch last year?
(i.e.: Product/Processes/Systems/Company Development/Culture)
(3) How many innovative products have you got in the pipeline?
(i.e.: Product/Processes/Systems/Company Development/Culture)
(4) How many innovative ideas were submitted within the last 12 months and how many of these were approved and completed?
(5) What proportion of your revenue comes from innovative products or services launched in the last two years?
(6) What savings were made from innovative ways of working (i.e. from changes to processes, systems, etc.) in the last two years?
(7) Do departments openly collaborate on innovative projects (please provide any evidence)?

Table 9.4 Process and technical audit questionnaire (Tiers 1,2,3 and 4)

(1) Describe how ideas are collected within the Company?
For example: (Where do the ideas originate?/Are any techniques used to create ideas?)
(2) Describe how these ideas are prioritised?
For example: (Are regular meetings held on the ideas going forward? What methods do you have in place to check progress and viability?)
(3) Describe the implementation process for new ideas?
(i.e.: Product/Processes/Systems/Company Development/Culture (dependent upon the department))
(4) How does the workforce regard innovation within the company?
For example: ***(Is the innovation process understood by all employees and are there initiatives to promote innovation within the Company?)***
(5) Explain the innovation structure within the company
For example: ***(Is there an innovation philosophy, goals etc.? Does senior management spend time supporting it? Are performance measures in place to reflect that strategy?)***

Table 9.5 Generic questionnaire overview

Positive rati rating (%)	(1–3) Ideas	(4–7) Prioritisation	(8–11) Implementation	(12–17) People and innovation	(18–23) Innovation strategy
Total average	**Weak (22.7)**	**Average (47.8)**	**Average (40.1)**	**Average (48.2)**	**Weak (25.7)**
Engineering and R&D	**Weak (33.3)**	**Average (45.8)**	**Average (41.7)**	**Average (55.5)**	**Average (45.7)**
Operations	**Weak (17.5)**	**Average (45.6)**	**Weak (38.7)**	**Average (45.0)**	**Weak (22.3)**
Servicing	**Weak (33.3)**	**Average (58.3)**	**Average (45.8)**	**Average (50.8)**	**Weak (22.3)**
Finance and IT	**Average (55.8)**	**Average (58.7)**	**Average (53.6)**	**Average (56.9)**	**Average (45.7)**

Key (Positive Rating)
$39 \geq$ *Weak* $40 \leq$ *Average* $\leq 59\ 60 \leq$ *Strong*

The table provides a *"Positive Rating"* for the individual departments and a "*Total average"* for all the departments. The rating was obtained by calculating the percentage of positive responses to the generic questionnaire. It should be noted that the figure is based upon the assumption that each question has an equal rating

From the table, it is apparent that the workforce does recognise that the company is innovative and that they see the value of innovation and do want to get involved. This is significant, as the workforce across all departments is open to developing innovation within the company.

Further investigation is required with regard to the response of a blame culture, as a significant number of respondents highlight that there was a blame culture within the company. This can be a potential obstacle to innovation as employees will be less likely to suggest and try new ideas in fear of failure and getting the blame for it.

The survey highlights that even though innovation is clearly occurring within the company, the process has been organic and lacking in clear systems across all five areas. It is normal that out of five steps, a company could be weak on 2 or 3 areas. This was not clearly the case here. The questionnaire indicates that a clear and coherent innovation policy and strategy for the company has to be developed and processes put in place to formalise and embed innovation. This initiative has to be clearly communicated to the entire workforce to ensure a culture of open innovation throughout the company.

9.7.2 *Performance Improvement Plan for Supplier Gas Industry*

In order for the senior management to take the lead and transform all the weakness to strength through the SGI 5-steps innovation framework, a workshop was arranged as an "away day" to improve and agree innovation strategy, as typified in Fig. 9.8. This included both senior and middle managers to ensure a united front is exposed in identifying and building an effective improvement plan within the organisation (Fig. 9.7). Subsequently, all tiers within SGI should undergo "in-house" training prior to full rolling out of the SGI *Innovation Management Ecosystem.*

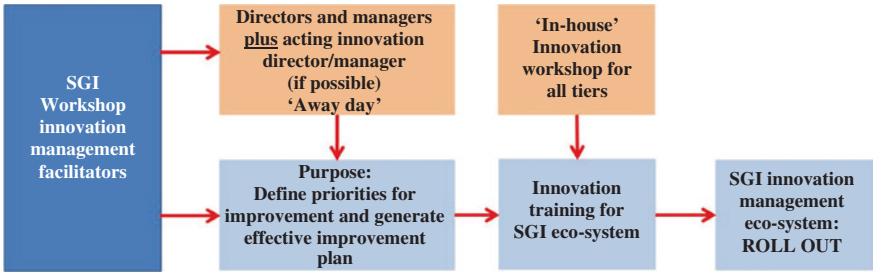


Fig. 9.8 SGI workshop for setting innovation improvement plan strategy

9.7.3 Boost Performance and Key Projects

As shown previously in Fig. 9.7, 'boost performance' is the next stage of improvement in innovation management ecosystem of SGI. This methodology is far from reach and is subject to more fundamental research. However, SGI has been attempting to implement this through the suggested ingredients as shown in Fig. 9.7 with various "key projects" with low-risk and high-impact projects. The duration of the entire process could take 18 months or more before any real outcome is to be evident. Patience and perseverance are virtues in innovation management.

9.8 Innovations in Gas Industries: *Case Studies*

The following four exemplified case studies have been provided which were driven through research and innovation process, leading to commercialisation. These examples could be related to both upstream and downstream areas of the gas and oil industry.

9.8.1 Carbon Nanotube Production: Case Study-1

9.8.1.1 Introduction

Natural gas (NG) is composed of methane, but it is typically mixed with varying amounts of heavier hydrocarbons such as ethane, propane, butane and pentane and non-hydrocarbon gases such as water vapour, hydrogen sulphide, carbon dioxide, helium, nitrogen and other compounds. NG processing essentially depends on the gas composition [13, 14].

In this study, the utilisation of NG that was previously flared was investigated experimentally using spraying and atomisation techniques for the production of single-wall carbon nanotubes (SNCNTs). These materials (SWCNTs) are mainly made up of 96.3 % of carbon, 2.91 % of carbon monoxide, while other metal and non-metal additives form the remaining 0.79 %.

9.8.1.2 Design Procedures and the Set-up

This section is discussed in two phases. The first phase describes the design of the atomiser used for the production of SWCNTs which was designed by the Spray Research Group (SRG) at the University of Salford, while the second phase describes the experimental procedure for the production of SWCNTs using the atomiser that was designed in the first phase. The second phase was realised in collaboration with the University of Oxford.

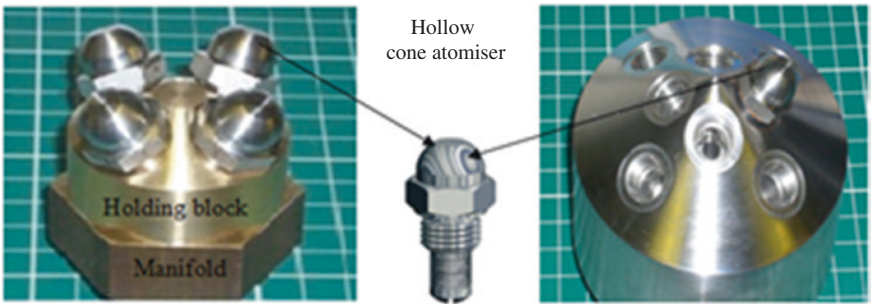


Fig. 9.9 Liquid spraying atomisers and holding block. **a** Manifold with 4 atomisers. **b** Manifold with 8 atomisers

The atomiser device was designed to operate at low pressure and consisted of a confinement tube with a cover for each open end, which were screwed onto the tube and a manifold of a hollow cone commercial swirl atomiser that interacted with a baffle plate (or *impactor*) as shown in Fig. 9.9. This baffle plate was used to separate the larger droplets and to produce a fine spray in the aerosol tube of the device Fig. 9.10. This is a novel method of producing fine spray droplets of $\leq$5 μm using standard pressure atomisers. The confinement tube was constructed from Perspex with dimensions: 250 mm length, 180 mm inside diameter and 3 mm wall thickness. On the centre of the top cover, an atomiser holding block of 50 mm diameter was fixed, for mounting the manifold of the four atomisers.

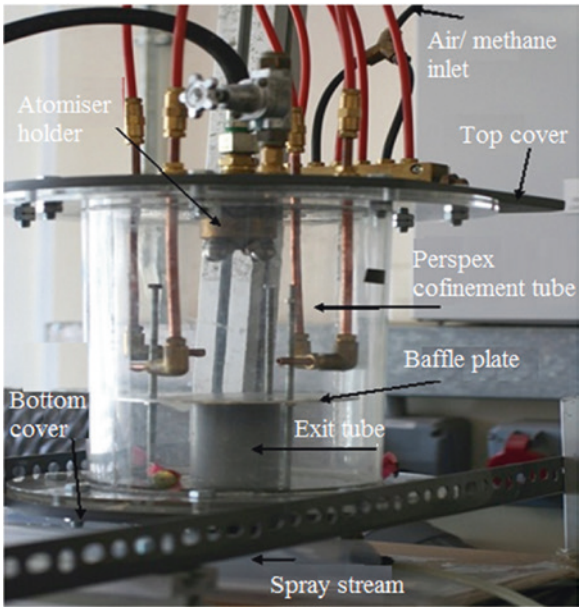


Fig. 9.10 Optimum baffle plate arrangement and atomiser assembly

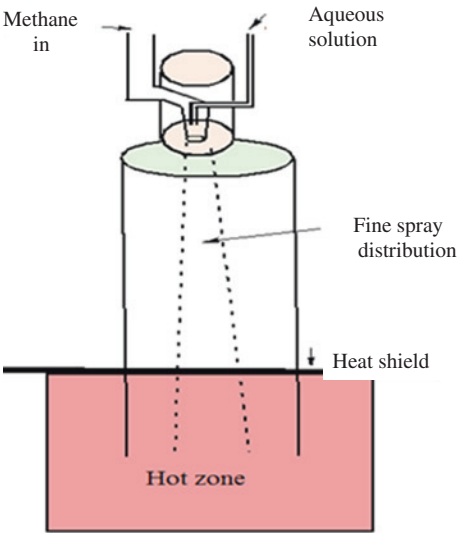


Fig. 9.11 Schematic set-up of the rig

The designed atomiser device assembly was mounted on a furnace (Carbolite type, STF 16/450 model). The furnace was heated to 1,200 °C, and methane was blown over aqueous solution of powdered metallic catalyst as shown in Fig. 9.11. The reaction products are as shown in Eqs. 9.1 and 9.2.

$$CH_4 + H_2O \rightarrow CO + 3H_2 \tag{9.1}$$

$$CH_4 \rightarrow 2H_2 + C\ (\text{SWCNTs}) \tag{9.2}$$

The carbon produced in Eq. 9.2 combines with the carbon monoxide in Eq. 9.1 to form carbon nanotubes.

Figure 9.12 shows the typical transmission electronic microscopy (TEM) images obtained during this preliminary test run. SWCNT particles were deposited as products and were studied using TEM. These limited results provide further

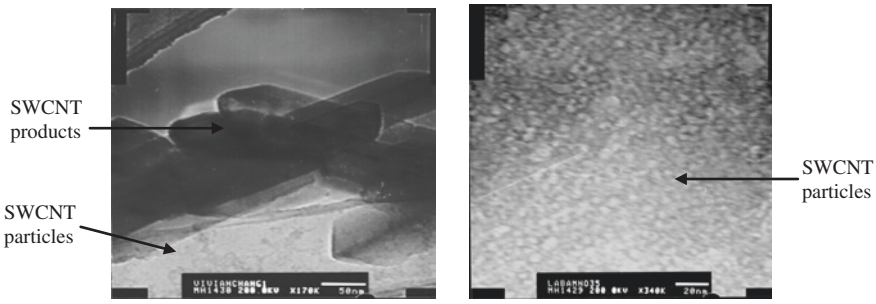


Fig. 9.12 TEM images of the produced SWCNT

assurance that flare gas can be utilised to produce SWCNT using the novel design for the atomiser device. A number of further trials have been conducted and processed at the University of Oxford which will be published in due course.

9.8.1.3 Final Remarks

The preliminary results showed that it is possible to produce SWCNTs from NG (mainly CH_4) that would have been flared by preliminary utilising the designed "atomiser device assembly". The utilisation of flared NG for SWCNTs production will reduce the CO_2 emissions into the atmosphere.

9.8.2 GTL Plant Effluent Treatment: Case Study-2

9.8.2.1 Overview of GTL Technology

The gas-to-liquid (GTL) technology consists of a chemical conversion of NG into a stable liquid by means of the Fischer–Tropsch (F–T) process. This conversion makes it possible to obtain products that can be consumed directly as fuel (diesel, kerosene and gasoline) or special products such as lubricants.

The conversion of NG to liquid is achieved by using two different routes: the direct and indirect routes [15]. The direct route involves an attempt to convert NG (methane) directly to liquid hydrocarbon. The difficulty of activating methane chemically makes this route complex and expensive. The indirect route involves the production of synthesis gas (or syngas) which is a mixture of H and carbon monoxide. It has a major advantage over the direct route. However, it requires a four-step process and elaborate separation and processing operations which are as follows:

- Desulphurisation of NG
- Conversion of NG to synthesis gas
- Conversion of synthesis gas to crude liquid products
- Upgrade of the crude products to finished products

Desulphurisation of NG 4 The NG feed should be free of water and sulphur to avoid catalyst poisoning. The desulphurisation step consists of passing sulphur (S) containing NG feed at about 300–400 °C over cobalt molybdenum (CoMo) catalyst in presence of H, to convert organic sulphur compounds to hydrogen sulphide (H_2S). This is then followed by adsorption of the hydrogen sulphide on zinc oxide (ZnO) in a guard chamber to form zinc sulphide (ZnS). This reaction reduces the sulphur level in the NG to less than 0.1 ppm wt which is the level that the oxidation and F–T catalysts can tolerate.

Conversion of NG to synthesis gas The conversion of desulphurised NG to synthesis gas (syngas) is accomplished by using any of the four commercially proven

reforming technologies [16]. Irrespective of the type of reformer used, the products of the reaction are the same; only the composition will change.

Steam Methane Reforming (SMR) Steam reforming is currently the widely practiced technology for syngas production. It is used for both large- and small-capacity GTL plants. Economic evaluation of GTL recommends it for smaller production capacities of under 5,000 barrel per day [17].

It involves reacting steam with NG over a bed of nickel oxide catalyst as shown in reaction 9.3. In complete combustion reaction, pure methane (CH4) molecule with two molecules of oxygen produces one CO_2 gas molecule, two water (H_2O) molecules in vapour form and heat. In practice, the combustion process is not always perfect, and when the air supply is inadequate, carbon monoxide (CO) and particulate matter (soot) are also produced.

$$CH_4 + H_2O \rightarrow CO + 3H_2 \tag{9.3}$$

A mixture of steam and desulphurised NG is converted to synthesis gas in one processing step in a steam reformer. The produced synthesis gas releases heat to boiler feed water to produce high-pressure steam. Typical operating temperature and pressure of a steam reformer are 870 °C and 400 psia at the outlet.

The disadvantages of conventional long tube steam methane reformers are their large size and high cost due to materials of construction [18].

Conversion of synthesis gas to crude liquid products The conversion of produced synthesis gas (CO and H2) to liquid hydrocarbon is a chain growth reaction of carbon monoxide and H on the surface of a heterogeneous catalyst. The catalyst is either iron based or cobalt based, and the reaction is highly exothermic. The basic reaction is expressed as follows:

$$nCO + 2nH_2 \rightarrow n\,(-CH_2-) + nH_2O \tag{9.4}$$

In the above equation, the $–CH_2–$ assembles to form hydrocarbon chains.

Upgrade the crude products* (*Hydro-cracking) In the upgrading step, the waxy paraffinic product is cracked in the presence of H to any required molecular weight product. After the hydrogenation, the product is fractionated in a conventional distillation section, to produce, diesel, kerosene, gasoline and waxes with a small proportions of gaseous products.

Figure 9.13 shows the emissions associated with various oil and gas processes including conventional oil, heavy oil and gas to liquids (GTL). As can be seen in Fig. 9.13, approximately 1,150 kg of CO_2 is produced by a tonne (7.52 barrels) of GTL product and based on over 20 GTL projects around the world, each producing about 34,000 barrels per day of GTL products; the volume of produced carbon dioxide calls for capture to avoid environmental emissions. These large volumes of carbon dioxide are available in GTL plant from the following sources.

(i) The NG supply
(ii) The reformer product and
(iii) The F–T reaction

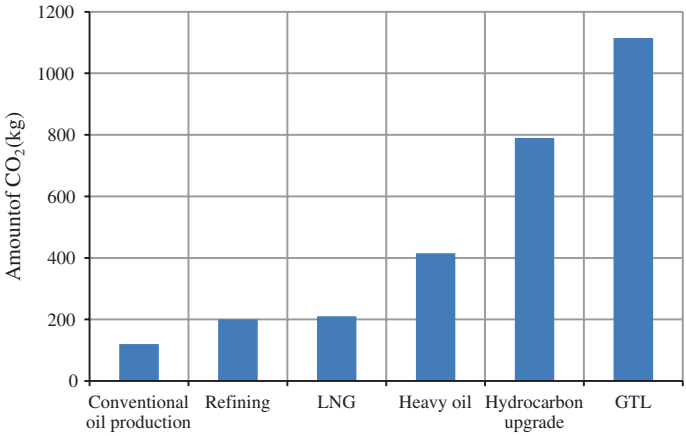


Fig. 9.13 Emissions associated with oil and gas operations

9.8.2.2 Effluent Quality and Treatment

All industrial operations produce wastewaters which must be returned to the environment. Effluent from the GTL plant is a source of environmental pollution. To exemplify, Table 9.6 shows typical effluent parameters from a NG synthesis plant in Nigeria.

Many NG conversion plants such as GTL use sulphuric acid to control the wastewater pH in the effluent treatment unit of the plant. The acid is purchased and stored in a storage tank within the utilities/offsite area. Late delivery of the acid will inevitably result in plant shutdown. Apart from this, the acid poses maintenance problems and it is very corrosive. Finally, accurate pH control using H_2SO_4 is difficult to achieve due to sudden pH increase at the end point (neutral

Table 9.6 Effluent quality before and after treatment

Parameter	FEPA* limit	Before treatment	After treatment
Temperature (°C)	40	35	27.57
pH	6–9	10.1	7.56
Conductivity (mhos/cm)	4,000	3,400	1,300
TDS (ppm)	2,000	3,100	515.15
TSS (ppm)	30	45	32
BOD at 20 °C (ppm)	20	25	22
DO (ppm)	5	7.3	5
Oil and grease (mg/l)	50	10	Traces
Heavy metals (ppm) (Fe, Ni, Cr, Zn)	3	1.73	1.49

*Federal Environmental Protection Agency (Nigeria)

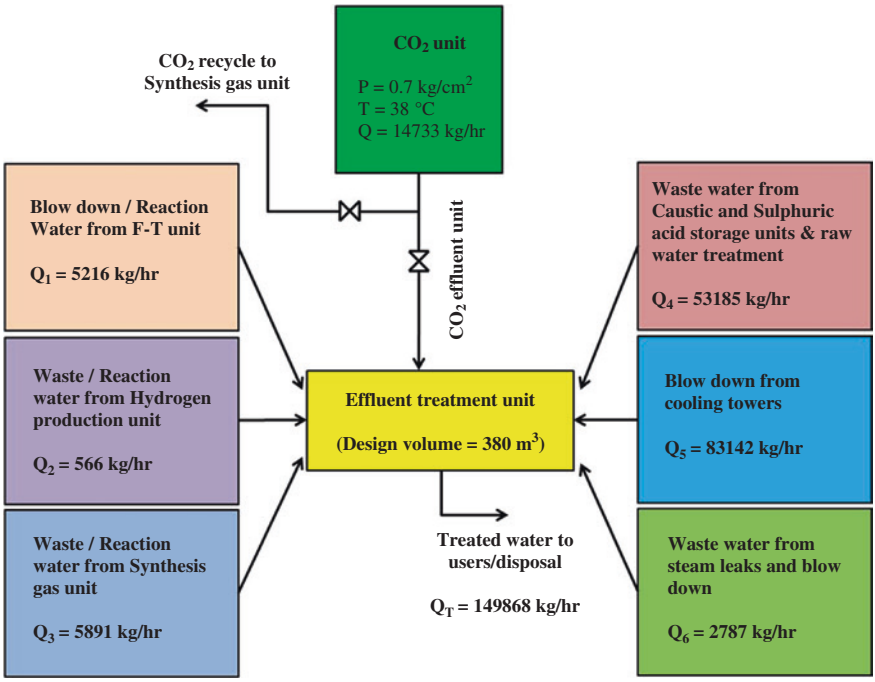


Fig. 9.14 Sources and quantities of effluent to the treatment unit

point) at a drop of small quantity of the acid. Hence, the replacement of H_2SO_4 by carbon dioxide which is produced within the plant complex as wastewater pH control can not only reduce the operating cost of the plant but also reduce carbon dioxide emissions from GTL plant to the atmosphere.

The advantages of carbon dioxide over mineral acids are as follows:

- The production of CO_2 could be within the plant which is readily available
- Due to its natural buffering action that cannot cause the pH to be reduced below 5, even if overdosing occurs

The effluent treatment unit receives the gas plant wastewater, GTL plant wastewater and reaction water from the F–T reaction. Other sources of wastewater to the treatment plant and their various flow rates are shown in Fig. 9.14.

9.8.2.3 Description of the Set-up

A laboratory-scale effluent system was designed and built to demonstrate the feasibility of utilising carbon dioxide produced as a result of reforming and F–T reactions for pH control in the effluent treatment unit of a GTL facility. The schematic of the system is shown in Fig. 9.15.

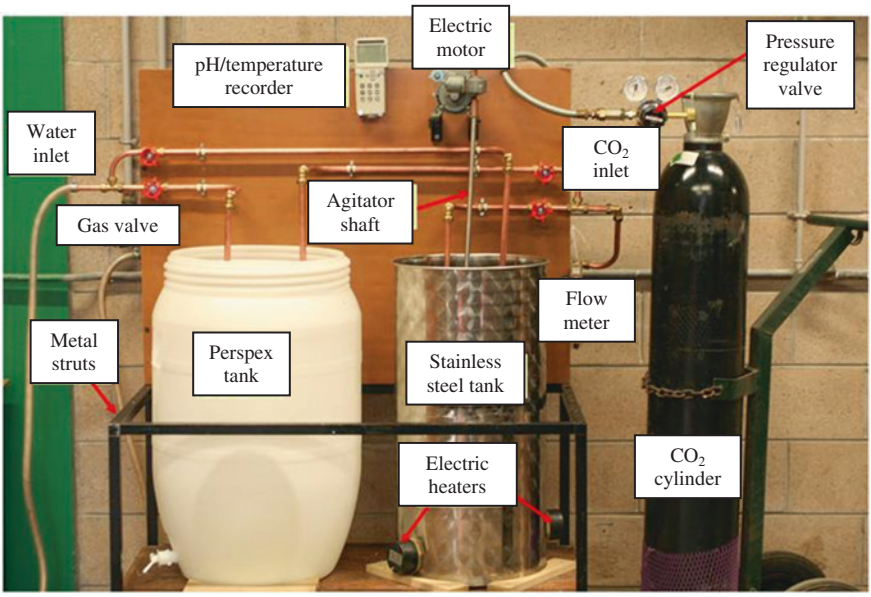


Fig. 9.15 Schematic laboratory set-up

Effects of effluent volume on CO_2 consumption Six sets of experiments were conducted for six effluent volumes between 30 and 100 l. The dispose carbon dioxide was fed into the tank containing the effluent water at a constant flow rate of 25 l/min. When the target pH of the solution dropped to ≤6.6, it was assumed the effluent was neutralised. The total volume of carbon dioxide and the change in pH with time for each volume was recorded.

9.8.2.4 Final Remarks

The experiments were carried out as described above and the values of the total volume of carbon dioxide required to neutralise a given volume of effluent water obtained. By plotting the graphs of carbon dioxide volumes versus pHs and volumes of the effluent, deductions could be made on total carbon dioxide requirements of any GTL plant, given the size and volume of the plant wastewater treatment unit. It will also help us to determine whether the carbon dioxide produced from the plant will be enough for the operation or, if too much, ways to handle the excess.

Figure 9.13 shows that to produce one tonne (7.52 barrels) of GTL product, an equivalent of 1,150 kg of CO_2 is produced in one day. Hence, a typical 34,000 barrels per day GTL plant will produce about 216,670 kg/h of CO_2. Figure 9.16 shows the consumption of carbon dioxide at different effluent volumes. As can be seen in Fig. 9.16, carbon dioxide consumption increases with increased effluent volume. From the experimental results, 100 litres (0.10 m^3) of the effluent volume

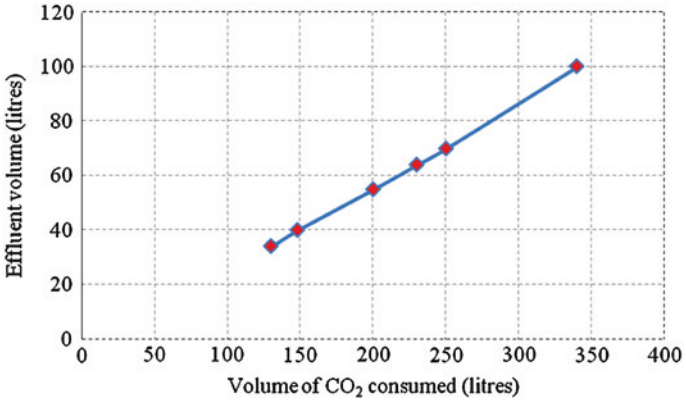


Fig. 9.16 Variation of effluent volume with volume of CO_2 utilised for neutralisation

consumed a total of 338 l/min (0.338 m^3/min) of carbon dioxide. Assuming that the effluent volume is operating at a capacity of 350 m^3, it will thus consume a total of 139,120 kg/hr of carbon dioxide. This shows that about 64 % of the total carbon dioxide produced in a typical GTL plant could be utilised within the plant.

About 64 % of the daily carbon dioxide production from the process reactions, which would have been emitted to the atmosphere, could be used in the treatment process. The consumption of carbon dioxide increased with increase in effluent volume; hence, more carbon dioxide will be captured and utilised as the effluent volume increases.

9.8.3 Scale Removal in Oil and Gas Production Tubing: Case Study-3

9.8.3.1 Introduction

One of the major problems facing the oil and gas industry is that of the mineral growth forming in hydrocarbon producing wellbore. Scale is a mineral deposit usually formed on surfaces in contact with water. Some common scales associated with oil field brines are calcium sulphate, barium sulphate and calcium carbonate. Scale deposition can be attributed to such factors as pressure drops, mixture of incompatible waters, changes in physical or chemical environment, or temperature changes that the well fluids encounter.

The primary effect of mineral growth in the tubing is to lower the production rate by increasing the surface roughness of the pipe and reducing the flowing area. The pressure therefore increases and production decreases. If the mineral growth increases, the access to lower sections of the well becomes impossible and ultimately the tubing itself will block completely. Treatment of scale down the

tubing-casing annulus protects down-hole equipment but may not be effective at the face of the formation.

9.8.3.2 Apparatus and Assembly

The schematic arrangement shown in Fig. 9.17 was designed to experimentally demonstrate oilfield scale removal through the use of flat fan atomisers to generate overlapping sprays from high-pressure water. It consists of a pump (1), unloading valve (2), safety valve (3), gauge pressure (4), three atomisers spray head (5), tubing sample (6), water reservoir (7) and water tank (8). The Speck Kolben pump (1) delivers high-pressure water at about 6 MPa and flow rate of 23 l/min to the "spray head". The purpose of this pump was to increase the pressure of the water to guarantee the desired high impact force which is necessary to remove the scale. The water pressure is regulated through the pressure gauge (4) to the spray head. A pressure relief valve (3) was provided within the pump circulation loop to protect the entire assembly from excessive pressure. As a further safety measure, an unloader valve (2) was installed within the pump as a regulator when the water pressure reaches the adjustment limit. Three pieces of Perspex tube were fixed together by means of flanges and aluminium bars. The top Perspex tube contains the scale sample tubular with

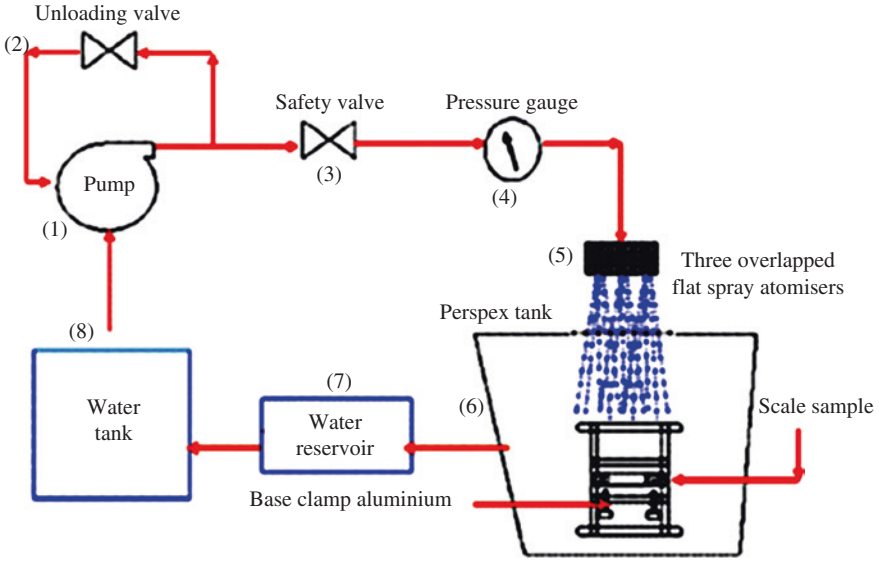


Fig. 9.17 Liquid spraying atomisers and holding block. *1* Water circulation pump. *2* Unloading valve. *3* Safety valve. *4* Pressure gauge. *5* Atomiser. *6* Perspex tank. *7* Water reservoir. *8* Water tank

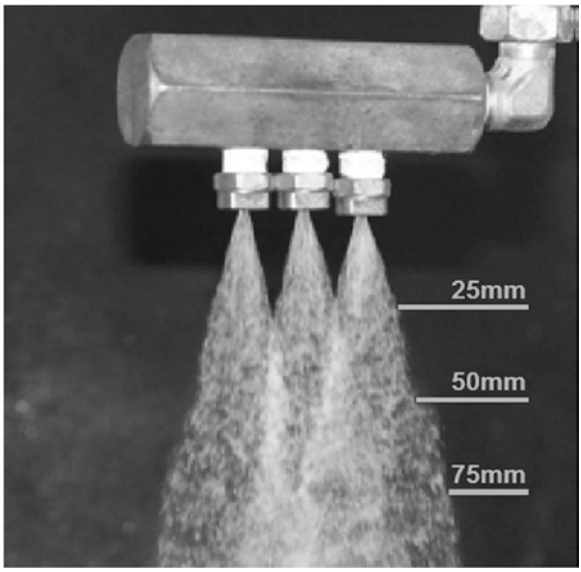


Fig. 9.18 Dense overlapping region

the middle one carrying the sieves. The bottom tube fixed below the bench returns the water to the reservoir tank. The water tank is made of high-density polyethylene (HDPE). This HDPE material is anti-rust, making it the best choice for the experimental work because of non-interference with the results. It was fitted with an outlet valve which was connected to the pump inlet line and an outlet valve for the return of water to the pump, thus maintaining a steady and continuous flow. A flat fan spray atomiser was used to provide the necessary water spraying stream to remove the corresponding scales easily (Fig. 9.18).

The aluminium clamp was fixed to the base of the Perspex tank and was pushed downwards to provide suction between the rubber seating and the bottom of the rectangular Perspex box. This arrangement allowed the scale sample to be rigidly clamped to the base of the tank, thus preventing it from any form of movement during the entire period of descaling.

The sieve was placed on top of the tank under the transparent rectangular Perspex box to collect scale removed from the sample during spraying operations. The mesh size employed for the collection of the removed scale has an aperture of (180 mm). It is made of stainless steel which conforms to BS410-2:200 British Standard Institution specifications.

9.8.3.3 Procedures for Scale Removal

The volume scale removal trials used the actual oilfield soft and hard scale samples, as well as a laboratory-made soft scale using candle wax as a means of

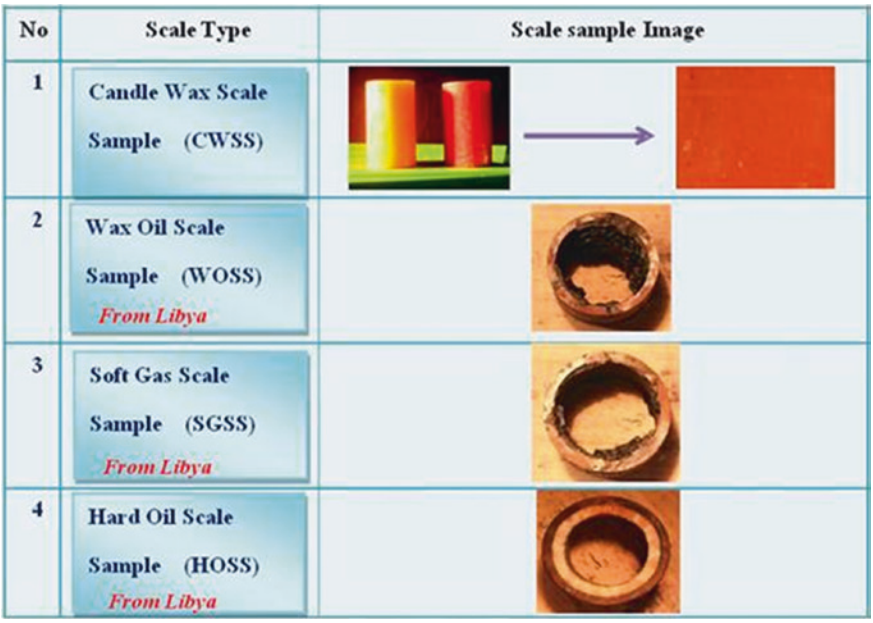


Fig. 9.19 Types of scale samples

comparison with the results obtained from the actual oilfield scale samples. The scale samples from Sirte Oil Company, in Libya shown in Fig. 9.19, and wax scales prepared from our research laboratory were used in this investigation.

The scale samples were placed on the aluminium base flange and secured in position. The spray head is lowered vertically downwards to carefully select a predetermined height to coincide with the radial axis of each sample, thus ensuring that the spray impacts the scale at the correct angle, at the right target, at different pressure, flow rate, height and precise timing, respectively.

The water pump was switched on to start the descaling operations at an initial water flow rate of 8 l/min. and a pressure of 3.7 MPa, and then adjusted to obtain the desired flow rate and pressure of 23 l/min and 6 MPa, respectively, for three overlapping flat fan atomisers. The overlapping spray jets from the atomisers were directed axially on to the scale sample to be removed as shown in Fig. 9.20. The experiment was performed on totally plugged pipe; hence, the jet was directed axially onto the scale. For partially plugged pipes, the jet will impinge tangentially on the scale.

After 15 min of spray operation, the water pump was switched off and the scale particles collected by the sieve were dried and weighed with a weighing scale. The procedure above was repeated for desired pressure of 4.8 MPa and at desired flow rate of 14 l/min using two overlapping flat fan spray atomisers for 10 and 15 min.

The snapshot of scale samples removed from soft and hard scales samples using three nozzles (3 N) placed at downstream position of 25 mm and

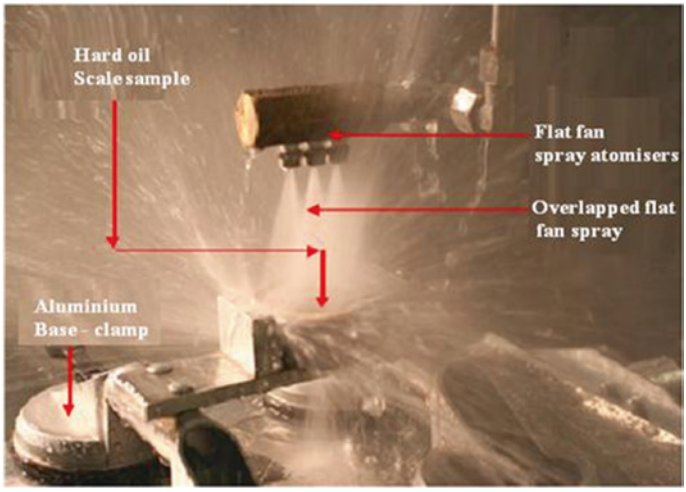


Fig. 9.20 Snapshot of descaling operation using three atomisers

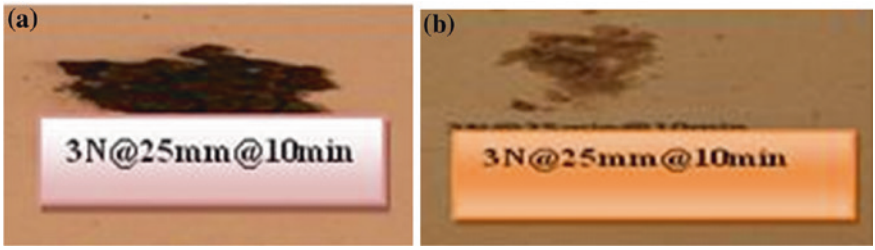


Fig. 9.21 Scale samples removed by high-pressure spray **a** soft scale samples and **b** hard scale sample

after 10 min (3 N@25 mm@10 min) of operation is shown in Fig. 9.21a and b, respectively.

9.8.3.4 Final Remarks

Experimental tests were performed for one, two and three atomisers at different downstream distances, flow rates and spraying time. Here, most of the results discussed centred on three overlapping flat fan atomiser arrangements at a pressure of 6 MPa and flow rate of 23 l/min though mention was made of one and two atomisers for comparisons.

Scale removal tests were carried out on oil scale samples using one, two and three atomisers at three different downstream distances of 25, 50 and 75 mm.

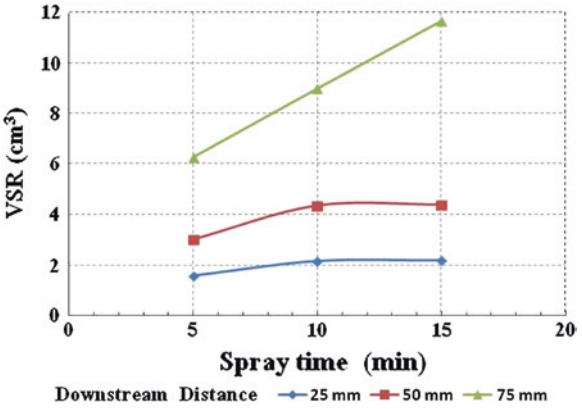


Fig. 9.22 Scale removals of soft gas scale sample (SGSS)

Figure 9.22 represents the volume of soft gas scale sample removed by using a combination of three flat fan spray atomisers at different downstream distances. The maximum volume of scales sheared off from the sample surface at 75 mm downstream, spray cone angle of 30° and at spraying time of 15 min was 11.688 cm^3. Applying the same experimental conditions on an oil wax scale, a total volume of 13.750 cm^3 was removed, while for hard scale samples as shown in Fig. 9.23, the volume of scale removed was approximately 0.989 cm^3.

It is interesting to note that with the application of one atomiser to three atomisers, there is a steady increase in the volume of scale removed, whereas for the three atomisers overlapping combinations, there is a significant increase in VSR compared with one- or two-atomiser applications. This confirms the dependency of VSR on spraying at high water supply pressure ($\leq$6 MPa) with high impact force ($\leq$0.657 MPa) and with overlapping configurations. As can be seen from Fig. 9.23, the most effective way of removing scale appeared to be with three-atomiser combinations particularly for downstream of the exit orifice. At 75 mm downstream, there are regions of highly dense overlapping sprays as shown in

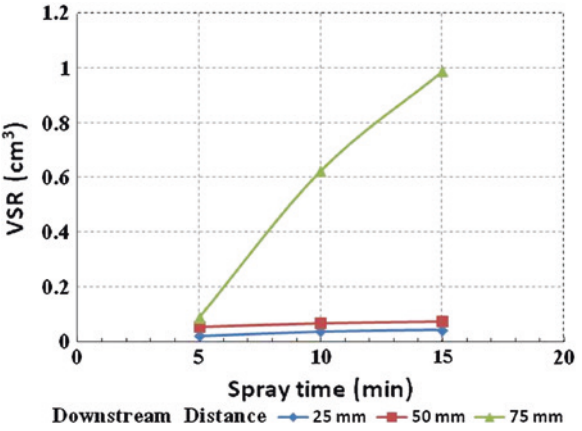


Fig. 9.23 Scale removals of hard oil scale sample (HOSS)

Fig. 9.18 which was due to high applied pressure which could provide substantial impact force in breaking up the hard oil scale.

From the results, it is evident that removing various volumes of scale depends on the selection of atomiser type (i.e. flat fan spray), number of atomisers, spray distance and time, supply water pressure and impact force of the spray.

It should be noted that all trials in this investigation were conducted under ambient conditions, at water pressures of 3.7, 4.8 and 6 MPa, flow rates of 8.0, 14.0 and 23 l/min and impact force of ≤0.657 MPa for one-, two- and three-atomiser configurations, respectively. Further investigations using a simulated pressure vessel (≤20 MPa) to remove the deposited scale under realistic condition as well as the gradual increase of supply water pressure are ongoing. The results and analysis presented here, from the current research, have provided an in-depth knowledge database which will greatly assist future investigators into this important and problematic area that currently exists in the petroleum industry.

9.8.4 Offshore LNG Unloading—Composite Hoses and BOG Analysis: Case Study-4

9.8.4.1 Introduction

Consistent growth of the liquefied natural gas (LNG) market due to increased world's energy demand has moved the industry towards exploration and production (E&P) of offshore gas reservoirs. The transfer system is widely regarded as the weakest component of the LNG chain [19]. While in mild conditions the system comprises rigid arms with swivels, in harsh environments this design no longer represents a reliable solution. Instead, flexible hoses have the ability to withstand tension loads and twist. Actually, these hoses are manufactured in diameters smaller than those required to obtain reasonable pressure drop through the system at typically 10,000 m^3/h. Thus, unloading a standard 150,000 m^3 LNG ship in approximately 12 h, two lines of 16″ are required. However, deployment of large-bore cryogenic hoses with diameters up to 20″ and 24″ reducing the system down to a single line is seen as a potential cost reduction.

These hoses are essentially a sequence of fabric sheets wrapped over an inner helical wire resulting in a periodically corrugated inner profile as shown in Fig. 9.24. This profile encourages turbulence, flow separation, and high shear stresses and induces the flow to swirl.

During unloading, LNG remains close to its boiling point, as a result, any energy or heat influx from the surroundings might cause a fraction of the LNG to vaporise. This boil-off gas (BOG) results from the energy added by the LNG pumps, friction losses, turbulence effects, heat transfer and flash vaporisation due to differential pressure between vessels [20]. The total BOG is estimated as follows:

$$\mathrm{BOG_{Total}} = \mathrm{BOG_{pipe}} + \mathrm{BOG_{hose}} + \mathrm{BOG_{falsh\ vap}} + \mathrm{BOG_{vap\ retrn}} + \mathrm{BOG_{LNG\ pumps}}$$

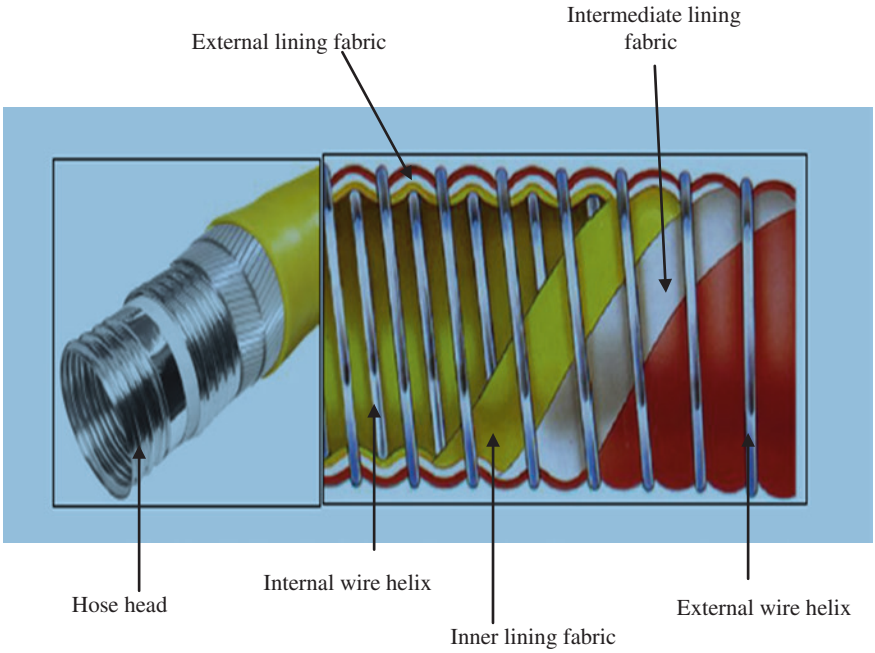


Fig. 9.24 Lightweight flexible composite cryogenic hose

If the amount BOG exceeds the capacity of the receiving terminal, flaring or venting methane into the atmosphere might occurs. In 2007, over 3.3 mtpa of BOG (nearly the annual capacity of a large LNG train) was lost during marine transportation. Its associated cost exceeded 1.275 billion USD [20]. Hence, research and development (R&D) of large-bore cryogenic hoses towards generation of BOG represents an important cost saving in regasification terminals.

A typical LNG unloading procedure is shown in Fig. 9.25. During unloading, some of the BOG is sent back to the ship tanks to keep a constant operating pressure within the vessels; the remaining fraction is diverted into the BOG compressor and the recondenser where the subcooled LNG condenses this extra BOG. Thereby, flaring or venting methane is avoided.

LNG unloading involves a large number of variables that are constantly changing throughout the entire process and making it difficult to obtain a unique solution. Thus, the novelty of this work relies not only on analysis of the interaction between components of a regasification terminal but also on estimation of BOG considering the entire range of values that each variable may reach during an unloading procedure.

To evaluate the consistency of our calculations, simulations in Aspen HYSYS were carried out and a comparison is presented in Table 9.7.

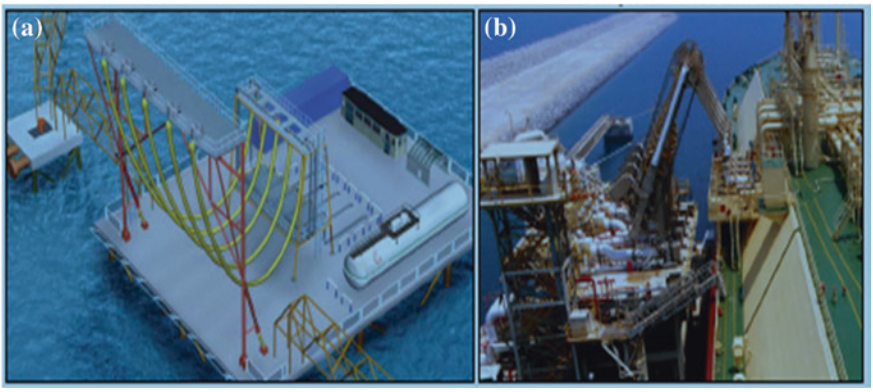


Fig. 9.25 A typical LNG transfer system **a** showing different transfer units and **b** hose/pipe connection to ship

Table 9.7 Accuracy of results predicted by Aspen Hysys and the new software

Energy consumption	HYSYS	Model	Units	Ref. err (%)
Compressor consumption	507.50	562.83	BHP	−10.90
Ship pump power	28.78	29.58	BHP	−2.80
Send-out pumps power	3,394.95	3,397.11	BHP	−0.06
Heat leak through unloading line	HYSYS	Model	Units	Ref. err (%)
Ship pipelines	114.86	109.34	kW	4.80
RT pipelines	142.30	135.67	kW	4.66
Hose transfer	262.91	251.52	kW	4.33
Boll-off through unloading line	HYSYS	Model	Units	Ref. err (%)
Ship pipelines	712.54	719.62	kg/hr	−0.99
RT pipelines	791.61	892.87	kg/hr	−12.79
Hose transfer	1,610.51	1,655.33	kg/hr	−2.78

9.8.4.2 CFD Analysis of Boil-off Gas

The Sparlat-Allmaras turbulent model was used to simulate the steady-state effects of pressure drop in isothermal and incompressible LNG flow. The fluid domain comprises a 16″ hose with helical corrugations and pitch of 30 mm, and the heights of the first and secondary corrugations are 7 and 1.5 mm, respectively.

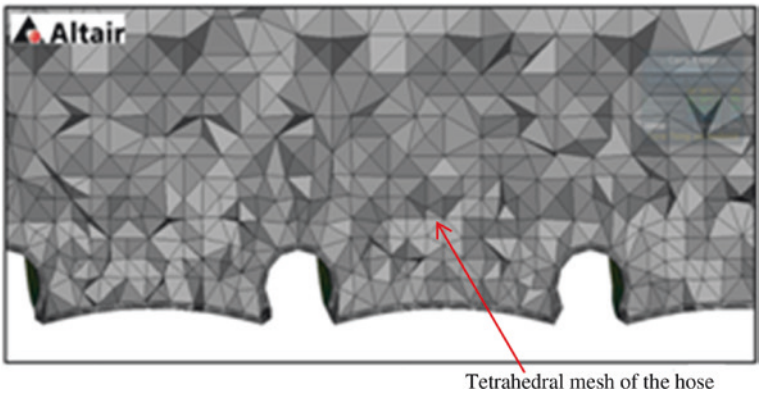


Fig. 9.26 Tetrahedral mesh of the cross section of the hose

A fluid domain length of 5D based on previous DNS works [21] ensures that the streamwise velocity is uncorrelated; thus, periodic boundary conditions can be imposed. In this approach, the velocity and eddy viscosity profiles at the exit are repeatedly reported at the entrance of the fluid domain; thereby, the entrance length is no longer needed to be computed. Due to our computational limitations, treatment near the wall consisted in a wall function. However, the height of the first layer was set to 200 wall units matching the results obtained in [22]. Figure 9.26 typifies the mesh of the cross section of the hose.

The average streamwise velocity solution detaches at the top of the primary corrugation, and recirculation vortices exist between consecutive corrugations. Moreover, the helical nature of these corrugations generates an angular component of the velocity-inducing swirl. A high-pressure zone occurs at 45º upstream of the corrugation, while low-pressure zones are expected at the top of the corrugation. This behaviour indicates that the flow periodically detaches at the top of each corrugation and reattaches at 45º upstream the next one.

The CFD result validates reasonably the Riley's correlation with a relative error of 9.5 %. The BOG reduces when a hose of 24″ is selected. Although the heat transfer is similar to the one obtained for 20 inches, the friction losses are approximately 60 % lower.

9.8.4.3 Effects of Hose Diameters on BOG

BOG as function of the LNG unloading rate at different hose diameters is presented in Fig. 9.27. It is evident that less heat transfer and pressure drop in a 24″ hose generates far less BOG than those obtained with a 16″ or 20″. Although the pressure losses in a 20″ ID hose are higher than those losses in a two lines of 16″, its associated BOG is smaller at unloading rates lower than 10,500 m^3/h. This tendency is expected as one single line reduces the area exposed to heat transfer.

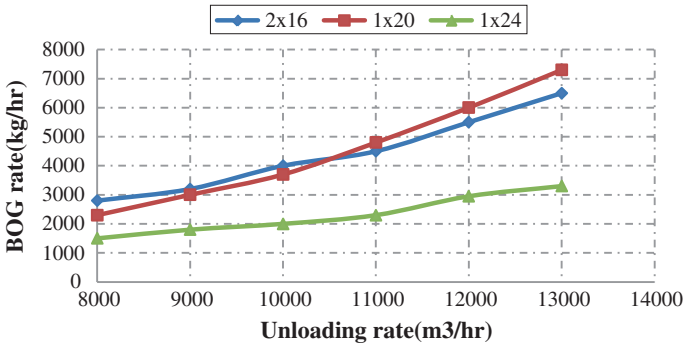


Fig. 9.27 Effects of hose diameter on BOG

However, at higher flow rates, the substantial increase in pressure losses in a 20″ hose becomes more significant. This result favours 16″ hoses at low unloading rates despite its higher pressure losses. In other words, the cost of running the compression station with lower BOG seems to be more important than the cost associated with the power consumption by the LNG pumps.

BOG can be controlled by the unloading rate. The higher the unloading rate, the higher the BOG sent to the ship tanks; as a result, the portion of BOG going to the recondenser is reduced. Nevertheless, as the unloading rate increases, the total BOG increases.

9.8.4.4 Annual Operating and Shipping Costs

Six target functions were forecast using Monte Carlo simulations at 95 % confidence level. These functions were as follows: energy consumption by LNG pumps and BOG compressors, BOG associated with the hose, total BOG, annual shipping costs and annual operating costs. Five thousand trials combining the upper and lower bounds of each decision variable allowed us to estimate their envelope and probability distribution. Energy consumption plus shipping costs were estimated using a triangular distribution of the energy prices and uncertainties of the friction factor obtained from CFD.

The benefits of large-bore hoses are seen by optimising annual costs. A constraint function which minimises both annual costs and BOG given by the hose was imposed. Results plotted in Fig. 9.28 have shown that low percentages of BOG can be achieved with a 24″ hose; however, a trade-off between BOG and annual costs needs to be considered due to the effects of the unloading rate. Thereby, low rates reduce the pressure losses and BOG at the expense of longer transfer time. The effect of diameter of the hose is quite interesting. Although a single 20″ line increases the energy cost, however, increasing the diameter further to 24″ leads to savings of the order of 35 %.

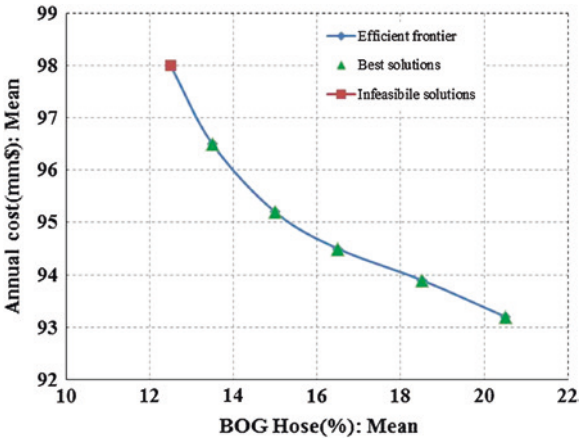


Fig. 9.28 Effect of BOG by the hose in annual costs

High vapour return temperatures generate large amounts of BOG and consequently higher power consumption by the compression station. However, these temperatures occur only during a few hours at the beginning of the unloading procedure.

9.8.4.5 Final Remarks

R&D of large-bore composite cryogenic hoses seems to be an optimistic solution looking towards BOG cost reduction in offshore LNG unloading. Although BOG is mostly affected by the hose diameter instead of the friction factor of the hose, 70 % of the total pressure drop comes from cryogenic hoses regardless of the hose diameter. Thus, reduction of friction factor by using liners is still an important area of research.

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