A METHODOLOGY FOR EVALUATING AND SELECTING PROCESSES FOR NATURAL GAS CONDITIONING

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SUMMARY

The aim of this work was to develop a generic methodology for evaluating and selecting, at the conceptual design phase of a project, the best process technology for Natural Gas conditioning.

A generic approach would be simple and require less time and would give a better understanding of why one process is to be preferred over another. This will lead to a better understanding of the problem. Such a methodology would be useful in evaluating existing, novel and hybrid technologies. However, to date no information is available in the published literature on such a generic approach to gas processing.

It is believed that the generic methodology presented here is the first available for choosing the best or cheapest method of separation for natural gas dew-point control.

Process cost data are derived from evaluations carried out by the vendors. These evaluations are then modelled using a steady-state simulation package. From the results of the modelling the cost data received are correlated and defined with respect to the design or sizing parameters. This allows comparisons between different process systems to be made in terms of the overall process.

The generic methodology is based on the concept of a Comparative Separation Cost. This takes into account the efficiency of each process, the value of its products, and the associated costs. To illustrate the general applicability of the methodology, three different cases suggested by BP Exploration are evaluated.

This work has shown that it is possible to identify the most competitive process options at the conceptual design phase and illustrate why one process has an advantage over another. Furthermore, the same methodology has been used to identify and evaluate hybrid processes. It has been determined here that in some cases they offer substantial advantages over the separate process techniques.

Key words: Generic Selection Methodology, Simulation, Natural Gas Separation Processes, Hydrocarbon Dew-pointing, Hybrid Processes.

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CHAPTER 1

1.0 INTRODUCTION

This thesis describes a simple method of evaluating and selecting, at the conceptual design phase of the project, the best processing option for Natural Gas conditioning.

Natural gas has been growing in importance due to the continuing growth in world energy demand and the need for an environmentally friendly fuel. Over the past decade or so, the purpose of processing raw natural gas and associated gas has been to yield a gas which meets the hydrocarbon dew point specifications or the required transportation pipeline specification. This is required before the gas is transported and purchased by the customer. When processing natural gas or an associated gas, valuable by-products such as Natural Gas Liquids (NGL), may be recovered.

The natural gas feed is a very complex mixture, containing as many as 20 or more different components. Each production well is likely to have different composition, pressure and temperature characteristics and these may change over the lifetime of the reservoir. As a result, every dew point control system is unique in terms of feed quality and quantity, operating pressure and temperature and product gas specification. Hence it is not usually possible to pre-select the best process.

For the wide range of different techniques available for conditioning natural gas, there is at the moment limited information available, on the relative economic benefits of the different techniques.

The conventional method of process selection is based on past experience, intuition or by preference for a particular technology on the part of the process engineer. This may lead to the selection of a process technology which is not the most appropriate, while some of the more novel process technologies may not even be evaluated.

The work, carried out in co-operation with BP Exploration, is about the development of a short-cut method for choosing between processes for the separation of the heavy hydrocarbon molecules from natural gas feed. The gas and condensate produced are required to meet the respective product specifications.

In developing a short-cut method of screening processes for natural gas conditioning, a number of questions concerning process modelling, performance and economic analysis have to be addressed. We may set these out as follows :

- How to evaluate the traditional and novel processing options on an equal footing ?
- How are these processes to be modeled ?
- What performance parameters are to be used and how do they affect the cost of separation ?
- How may the different processes be compared with each other ?
- Can this method be applied generically, for any given feed gas/product specification?

There are currently a number of gas processing techniques that could be employed for dew point control and NGL recovery. In order to ensure coverage of the majority of the processing options available, the following suite of technologies have been considered :

1. Joule-Thomson expansion,
2. Turbo-expansion,
3. Vortex tube,
4. Refrigeration,
5. Membranes,
6. Absorption,
7. Adsorption and
8. Refluxing condenser.

This is a highly complex problem with so many different processes which fall into many different categories. To simplify the problem the above listed processes are categorised as either equilibrium or as non-equilibrium methods. Methods 1 to 4, are classified as equilibrium methods, and 5 to 8, as non-equilibrium methods. A further complication in the case of non-equilibrium methods is that each technology separates the heavy hydrocarbons from gas stream, differently. Each technique has its own complexities and technical characteristics, for example, the adsorption and membrane processes are able to operate outside the two-phase region of the phase envelope but the equilibrium methods operate within it.

The next questions to be asked are: Can the method developed be used to illustrate why one process is better than the other? Under what circumstances, if at all, do the non-equilibrium processes have an advantage over the equilibrium processes ?

Having developed a method of evaluation which places all the processing schemes on an equal footing, it will then used to select the best processes for three typical feed gas compositions / product specifications.

Further applications of the methodology developed, is to determine the potential of combining two or more (hybrid) processes. This is potentially an important area for future development, to innovate and reduce cost of gas processing. However, to-date little or no information is available in the published literature, on the technical evaluation and the economic benefits of these processes. The aim here is to identify potential hybrid process schemes and determine whether they can compete with separate process techniques.

An economic evaluation was carried out. This involved obtaining cost information from a number of different sources, including the equipment vendors. To make sure that it was commercially viable, all the enquires were made under the auspices of BP Exploration Facilities Engineering. This gave comparisons of the capital costs of the main separation units, but to obtain a real comparison of each gas separation method the process as a whole must be considered, taking into account the reprocessing of the product streams.

The work is briefly described in the next section.

1.1 PROGRAMME OF WORK

This thesis describes the development of a methodology for evaluating processes for natural gas separation. A brief introduction to each chapter is set out below.

The Literature Review begins with an historical perspective and an overview of the natural gas business and the role it plays in today's world. It is followed by discussions of gas conditioning and the role of phase envelopes. A section of the review covers the merits of different selection methods that could be used to evaluate processes. The bulk of the review covers the alternative gas separation methods.

Chapter Three introduces the process design packages that can be used to model / simulate process schemes, and describes the architecture of the various process simulators. The chosen simulator is detailed and its capabilities and limitations set out.

Chapter Four discusses the method used to simulate alternative process schemes. The modules used to represent the real systems are briefly discussed.

Chapter Five covers the economic evaluation of the separation processes. This involved obtaining capital cost data for the alternative gas processing plant schemes. The data are then simplified and put into a generic format. For this purpose new cost equations in terms of the process design parameters had to be developed.

The natural gas separation selection methodology developed here divides, as we have seen, the processes into equilibrium and non-equilibrium methods. Chapters Six and

Seven cover the equilibrium and non-equilibrium methods respectively. The proposed methodology is discussed stage by stage, before being evaluated.

The next step is that the process selection methodology developed above is applied to hybrid processes. Chapter Eight begins with the problems associated with evaluating hybrid processes. The guide lines for overcoming these problems are then discussed. Also included are examples of some possible hybrid combinations that may be commercially viable. The chapter concludes with a demonstration of the method.

Chapter Nine discusses the proposed process evaluation methodology and how the principles might be used in application to solve the selection problem.

Chapters Ten and Eleven cover the main conclusions and future work respectively.

CHAPTER 2

2.0 LITERATURE REVIEW

2.1 NATURAL GAS

Natural gas is essentially a mixture of light saturated hydrocarbons in the range methane to propane, together with small quantities of heavier hydrocarbons up to C_{20}+. Nitrogen, carbon dioxide and hydrogen sulfide may occur in significant percentages, for example up to 30%. Natural gas has been growing in importance, largely due to the continuing growth in the world's energy demand, nuclear set backs and the 'green wave', spurred on by society's concern over greenhouse gas emissions and acid rain. Natural gas is a clean fuel, producing about half the carbon dioxide compared to that from oil or coal, virtually no SO_x and little NO_x. Carbon dioxide is the main greenhouse gas, while SO_x and NO_x are the principal contributors to acid rain (1).

The oil crisis of the 1970's brought to the attention of the world the fact these current energy supplies are not endless. As a result changes in fuel consumption have been made, driven by technological advances coupled with the economic and environmental policies imposed by governments. The current rate of world energy source depletion will become even more exaggerated due to a rapidly increasing population in the Third World.

In 1970, world gas consumption was 35 trillion cubic feet (TCF), by 1992, it had risen to 70 TCF. During this period the Third World energy use has doubled according to the United Nations report 'Two centuries of population growth, 1950-2150', and is generally projected to double again in the next 15 years, expanding six-fold by the year 2050.

Despite the uneven distribution of gas reserves throughout the world, as far as can be predicted global gas supplies are abundant and estimated gas demands can be met.

According to the latest 'BP Review of World Gas in 1993', proven gas reserves amount to nearly 138.3 trillion cubic meters. This translates into approximately 120,000 million tonnes of oil (mtoe) equivalent. Oil, with 136, 500 million tonnes, was ahead. However, proven reserves of natural gas are rising, for example in 1992 gas reserves grew by 10 per cent, whereas the oil reserves only grew by 0.8 per cent. Table 1, below, gives the World proven gas reserves by region in 1992 (2, 3, 4).

Table 1: World Gas Reserves by region 1992.

Region	percentage (%)
Asia & Australasia	6.9
Africa	7.1
Non-OECD Europe including Russia	40.2
Russia	34.4
OECD Europe	3.8
North America	5.4
Latin America	5.4
Middle East	31.0

Gas now has a reserves-to-production ratio of 65 years, compared with oil's 43 years. Geologists claim that there are much greater reserves still waiting to be discovered.

The use of natural gas does not solve all the problems associated with fossil fuels but it is an important step towards finding an environmentally friendly and economically stable energy source.

2.1.1 GAS CONDITIONING

Most natural gas is transported from the producer to the market by pipeline. However, before transportation, the raw gas is processed to yield a gas which either meets the pipeline or customers specifications (sales gas). The pipeline specification ensures single phase flow in the pipeline and takes into account pressure/temperature profiles in the pipeline. The sales gas specification is generally more severe and is usually stated as a dew point for the product gas. This typically involves drying, treating and natural gas liquid recovery.

2.1.2 SPECIFICATIONS

There is considerable variation in the specification of apparently similar gas products due to the many ways in which they are produced and the purposes for which they are required. Accordingly, in most cases the specification is negotiated between the customer and the producer. Any or all of the following quality parameters are normally included in the sales contract.

2.1.3 CALORIFIC VALUE

The calorific value is the amount of energy released by complete combustion of a unit quantity of a material. For natural gas, the heating value is usually expressed as the Gross or Higher heating value. Gross heating value is the energy evolved on burning a standard volume of gas when the total combustion gases are cooled to the reference temperature and the water of combustion condensed. The condensing water releases its latent heat which is theoretically available for heating purposes. This represents the maximum possible heat available.

The calorific value of a fuel gas mixture can be calculated as the sum of the calorific value of each of the components, each multiplied by the corresponding mole fraction. The answer obtained has to be corrected for the compressibility of the mixture (5,6).

2.1.4 WOBBE NUMBER

The Wobbe number is a number whose value is proportional to the heat input to a burner at constant pressure. Two fuels with the same heating value (within + or - 5%) will deliver about the same energy at the burner and are reasonably interchangeable. In Britain the Wobbe number is defined by:

$$\text{Wobbe Number} = \text{Gross Heating value} / \sqrt{Sg}$$

where Sg = Specific Gravity (dimensionless)

2.1.5 HYDROCARBON DEW POINT

The hydrocarbon dew point is usually stated as a specified pressure and temperature which will define a phase envelope for the product gas. This will vary depending on the climate and the customer.

2.1.6 WATER DEW POINT

This specification is the mass of water per unit volume of gas or the maximum allowable water dew point temperature at a specified pressure.

2.1.7 CONDENSATE (NGL)

The condensate specification is conventially measured in terms of Reid Vapour Pressure (RVP). Depending on the end use, the condensate mixture can contain varying amounts of butanes provided the allowable vapour pressure specification is not exceeded.

2.2 THE ROLE OF PHASE DIAGRAMS

Gas processing often involves the separation of a liquid-phase from a vapour-phase. A conventional phase separator is a pressure vessel which separates liquids from vapours. Understanding the phase characteristics, of a fluid stream enables the engineer to determine by calculation whether a fluid stream of known composition is vapour or liquid under certain conditions of pressure and temperature. The amount and compositions of vapour and liquid may also be determined if the stream is a mixture. If the vapour pressure of pure hydrocarbon is plotted against temperature, the resulting vapour pressure curve may be represented by a diagram such as that shown in Figure 1 (7).

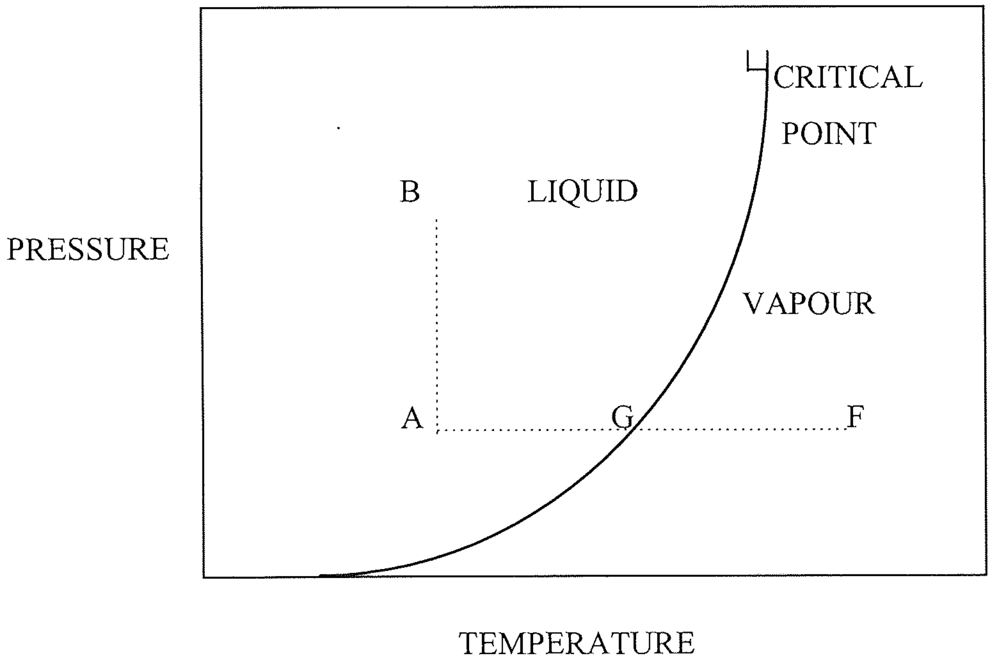


Figure 1 : Typical phase diagram for a pure hydrocarbon.

As the fluid pressure increases a critical point is reached where the properties of the liquid and vapour phases become identical and two separate phases cannot coexist. Above this point no phase separation can by definition take place.

At the conditions of point F the hydrocarbon is a superheated vapour. When the temperature is reduced to point G a liquid phase will appear. The temperature will remain constant at this point until all of the hydrocarbon vapour has become liquid after which sub-cooled liquid will be obtained by going to point A. An increase in pressure from point A to B will cause the liquid to increase in density but to remain in the liquid phase (8).

For a pure component, two phases can simultaneously exist only on the vapour pressure line. Elsewhere it exists only as one homogeneous fluid phase, either as a liquid or as a gas. For a multi-component mixture, another variable must be added to the phase diagram - composition. The location of the quality lines showing vapour-liquid splits with the phase envelope at different temperatures and pressures are dependent on composition. Instead of a single line there is a region in which two phases co-exist as shown in Figure 2. The two phase region of the phase diagram is bounded on one side by a bubble point line and on the other by a dew point line with the two combining at the critical point. The bubble point is the pressure at which the first bubble of gas comes out of a liquid, while the dew point occurs when the first drop of liquid condenses from a gas.

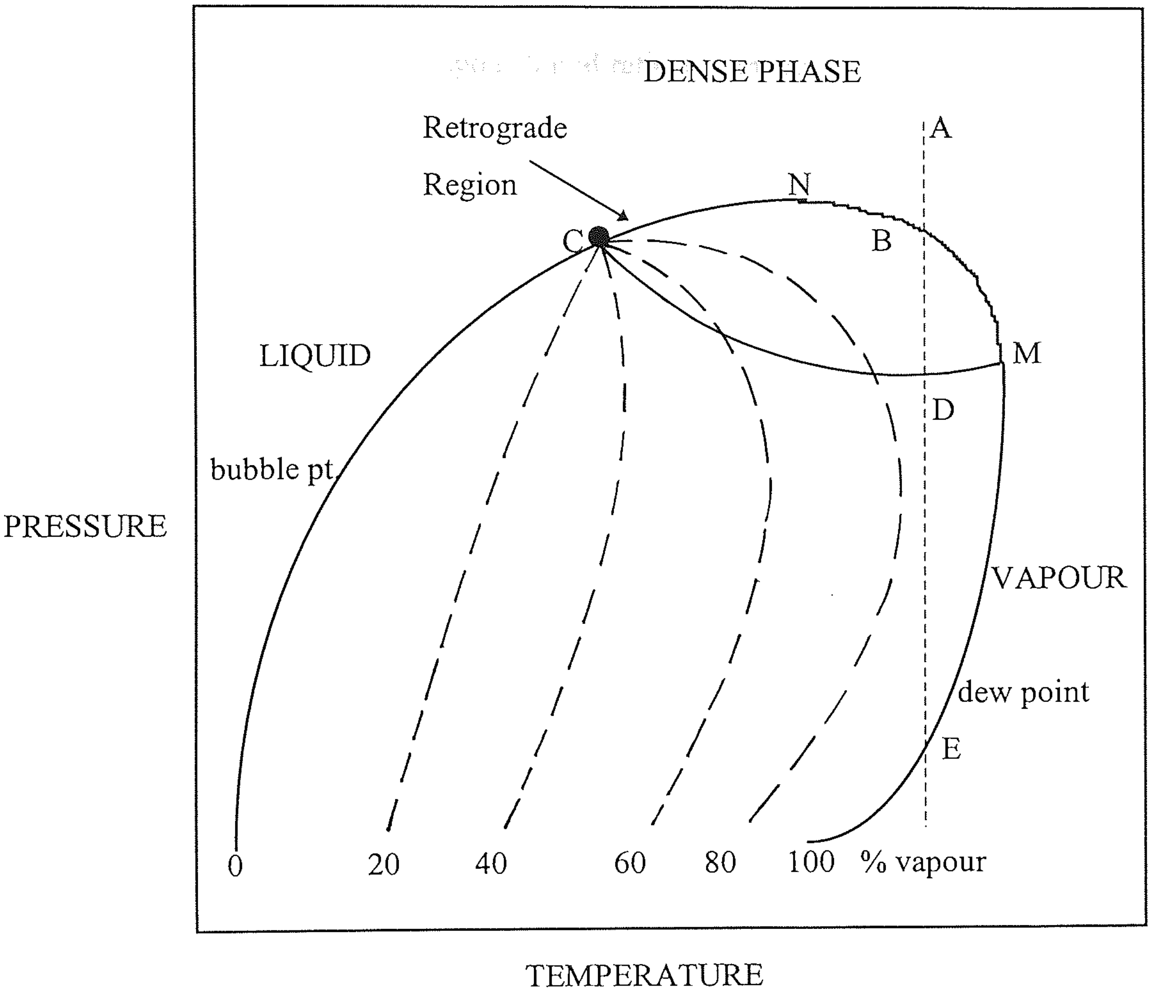


Figure 2 : Typical phase diagram for a multicomponent hydrocarbon mixture.

There are several terms used to define the locations of various points on the phase envelope.

Point N : Cricondenbar - the maximum pressure at which liquid and vapour may exist.

Point M : Cricondentherm - the maximum temperature at which liquid and vapour may co-exist in equilibrium.

Point C : The critical point - at which both phases become identical in composition-normally exists to the left of the cricondenbar for naturally-occurring hydrocarbon mixtures. The location of Point C is very important, for it fixes the shape of the quality lines which in turn governs the vapour-liquid ratio at a given temperature and pressure within the phase envelope.

Retrograde Region : The area inside the phase envelope where condensation of liquid occurs by lowering pressure or increasing temperature which is the opposite of normal behavior.

Quality Lines : Those lines show constant vapour percentages which intersect at the critical point (C).

Dense Phase : The area above the cricondenbar (point N) where the properties of the fluid are neither that of gas or liquid, i.e., it is denser than gas and more compressible than liquid.

Line ABDE represents a typical isothermal retrograde condensation process occurring in a gas/condensate fluid. Point A represents the single-phase fluid outside the phase envelope. As pressure is lowered, point B is reached where condensation begins. As the pressure is lowered further, more liquid forms because of the change in the slope of the quality lines. The retrograde area is governed by the inflection points of such lines. As the process continues outside the retrograde area, the liquid re-vaporises until the dew point is reached (Point E). Below E the fluid is 100% vapour.

2.2.1 EFFECT OF COMPOSITION

The composition has a significant effect on the shape and location of the phase envelope. For example, as a new natural gas stream is processed, the phase envelope begins to shift to the left. This is because the heavier components of the gas stream are progressively being removed, and the percentage of lights increases. It is therefore important to obtain reliable phase envelope predictions. The accurate prediction of bubble and dew point lines may be critical to both design and/or operation. The most difficult region to model is towards the top of the phase envelope, where dew point and bubble point curves converge. Quality lines are close together, making dew-point modelling difficult.

During the course of this project phase envelopes for different mixtures will be determined by computer simulation using the BP GENESIS simulator.

2.3 SELECTION SYSTEM

Over the last two decades considerable effort has been expended on development of systematic methods for choosing between processes. Nishida [1981] et al, (9) reviewed over 190 articles in field of process synthesis that had been produced in the previous decade alone. Most of this work has been focused on distillation column sequencing (separation trains), heat exchanger networks, control systems and chemical reaction paths (10). However, the more general problem of separation synthesis, using a number of different separation methods, has received little attention. Moreover, despite its importance, the synthesis of separation sequences for gaseous mixtures has been completely neglected in the process design literature. According to Barnicki [1992] et al, (11), "since the first proposals by Rudd [1968], (12), Masso and Rudd [1969], (13), advocating a systematic approach to separation system synthesis, not one article has appeared on any aspects of gas/vapour separation system synthesis."

The approaches used in the selection of separation methods can be divided into the following categories :

1. The Expert systems.
2. Case by case analysis.
3. The generic approach.

The expert system often takes the form of heuristics or rules-of-thumb which are used in solving the problem. These rules allow the process designers to locate one or several acceptable solutions. The method is based on the experience that has been built-up over a period of time with designing similar systems and in the understanding of unit operations, to form rules-of-thumb. The structure of the expert system is organised such that it emulates the approach that an expert process engineer follows.

The disadvantages of the expert system are that it is strongly dependent on the quality of the knowledge that has been incorporated. It may give no fundamental understanding of why the best method is the best, or why one process is better than another. Furthermore, the rules are based on past experience/expertise, as a result it is often difficult to evaluate new processes and thus recognize any advantages novel and hybrid processes may offer.

The case by case analysis is ideal for optimisation, taking into account the latest technological and processing improvements available (14). However, it is too time consuming and long-winded to be used for screening process options, for further analysis. Furthermore, it is a very complicated, expensive process and therefore not ideal for use at the conceptual design phase of the project.

A generic approach would be simple and less time consuming than the two approaches mentioned above, and would gives better understanding of why one process is better than another. This leads to a better understanding of the problem. Such a method would be equally adept at evaluating existing, novel and hybrid technologies . However, to-date no information is available in the published literature, on the generic approach for gas processing.

It is believed that the generic methodology presented here is the first available for choosing the best method of separation for gas dew point control.

2.4 ALTERNATIVE GAS SEPARATION TECHNOLOGIES

There are currently a number of gas processing technologies that could be employed for gas separation. In order to ensure coverage of the majority of processing options available the following suite of technologies have been covered.

2.4.1 JOULE-THOMSON EXPANSION

The Joule-Thomson principle, which has been applied for many years in oil and gas production, consists of adiabatic expansion of gas through a choke or valve from a high pressure to a low pressure (15). With appropriate heat exchange and a large pressure differential, cryogenic temperatures can be attained. However, as the expansion across the valve is considered adiabatic, no heat exchange between the gas and the surrounding occurs, and the First Law of Thermodynamics simplifies to:

$$\Delta H = 0$$

$$\Sigma\,(m_i h_i)_{\text{in}} = \Sigma\,(m_i h_i)_{\text{out}}$$

This means that the enthalpy of the stream after the expansion is equal to the enthalpy of the stream before the expansion. Therefore, the Joule-Thomson (J-T) valve is a constant enthalpy device that can produce cooling as a result of a reduction in pressure (16, 17). Dependant on the shape of the phase envelope and the initial starting conditions it is possible to either condense liquids or vaporise them.

This process is generally used when free pressure drop is available. It is advantageous to use this process if the following conditions exist.

- with low gas flow rates.

- if the gas is lean and high levels of liquid recovery is not economical.
- when it is required to operate under widely varying inlet gas rates (compared to turbo-expander).
- also, it is simple to operate. Therefore ideal for remote locations.
- minimum equipment requirement.

The disadvantages are that it is:

- inefficient at cooling the gas.
- not good for very low temperatures

2.4.2 TURBO-EXPANSION

The use of turbo-expanders in gas processing plants began in the early 1960's and by 1970, most new gas processing plants for ethane or propane recovery were being designed to incorporate them (18, 19).

The turbo-expander recovers useful work from the expansion of the gas stream. The process is an almost isentropic device that, like the J-T valve, can produce cooling as a result of a reduction in pressure (20, 21, 22). The expander removes enthalpy from the gas by converting it into shaft work that is used to drive a recompressor. In consequence turbo-expanders produce a larger temperature drop than obtained from the same expansion ratio across a choke valve, and are therefore able to recover more liquid from the gas (23, 24, 25).

The process can be analysed through the flow form of First Law of Thermodynamics (26):

$$\Delta H = Q - W$$

Since turbo-expansion is considered to be adiabatic.

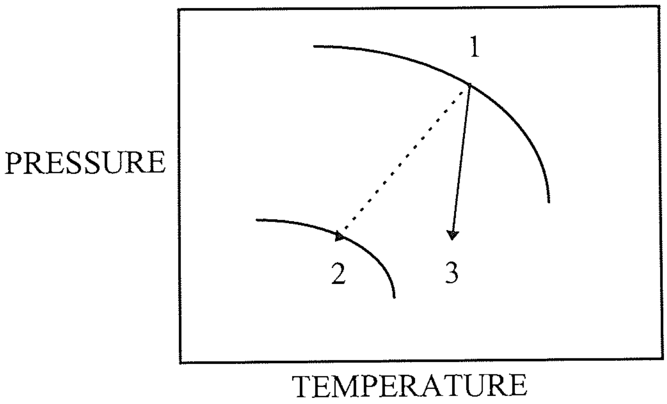
$$Q = 0$$

the First Law simplifies to

$$\Delta H = -W$$

i.e. work is obtained from the gas stream through the use of a turbine.

The diagram below illustrates the improvement in gas treatment that can be achieved with a turbo-expander over a J-T valve.



Assuming that at the expander inlet the gas is at its dew point, it flows through the expander, and its pressure-temperature path is shown by the dashed line from point 1 to point 2. Point 2 represents the outlet of the expander. If the gas had been expanded without doing any driver work, the expansion path would be from point 1 to point 3.

Turbo-expander applications are limited to expansion ratios of about 3 to 4. Greater ratio reduces expander efficiency to the extent that two-stage expansion may be advisable.

When turbo-expanders, first appeared great concern was expressed over the possibility of any liquid entering the machine because of the high speeds obtained {(10-40) x 10^3 RPM}. Some manufacturers will now accept small amounts of liquid in the feed gas and up to 40% wt of the feed as liquid in the outlet. But if processing the gas to very low temperatures, it is very important to make sure that the stream entering the turbo-expander is dry. Otherwise the formation of solid hydrates may result, which the turbo-expander cannot tolerate.

Selection of the turbo-expander process is indicated if two or more of the following conditions exist (27).

- pressure drop is available between supply and demand.
- lean gas.
- very low temperatures required (e.g. ethane recovery).
- compact plant layout.
- high utility costs.
- flexibility of operation.

2.4.3 VORTEX TUBE

The Vortex Tube is an emerging technology that may offer substantially higher efficiency than the Joule-Thomson process. The vortex tube separator (Figure 3, page 30) is a device which is used to extract Natural Gas Liquids (NGL's) from a natural gas stream. It consists of a tube in which a gas or gas-liquid mixture enters tangentially into the cylindrical tube and is allowed to expand to a lower pressure forming a high velocity vortex. If the pressure ratio between the feed and product streams is higher than the critical pressure ratio, then the inlet velocity will be the velocity of sound (sonic) (28, 29, 30, 31). If the pressure drop is increased, the inlet velocity and the volume flow rate will remain constant. If on the other hand the pressure drop ratio decreases below the critical pressure ratio, the performance of the vortex tube decreases.

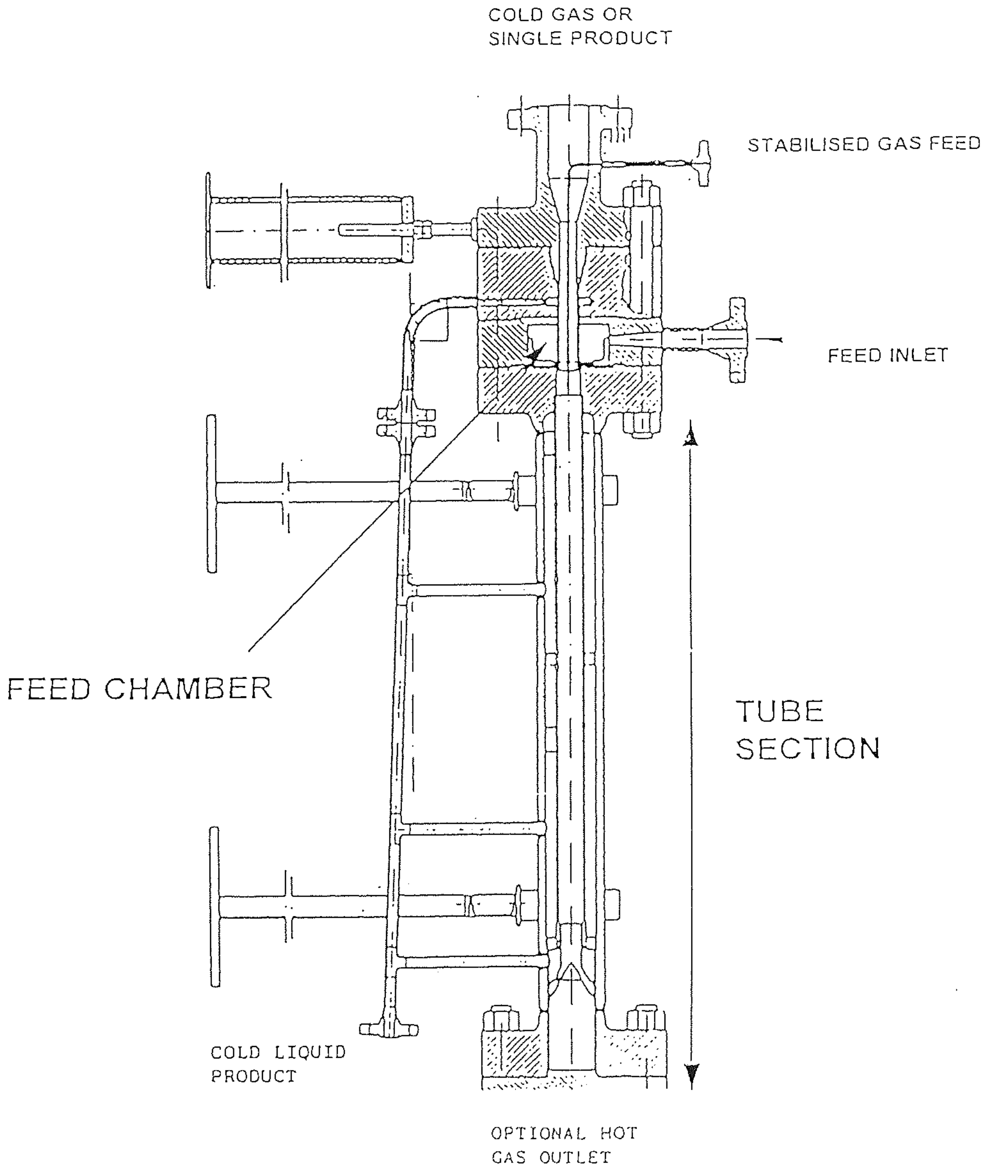


Figure 3 : Vortex tube separator.

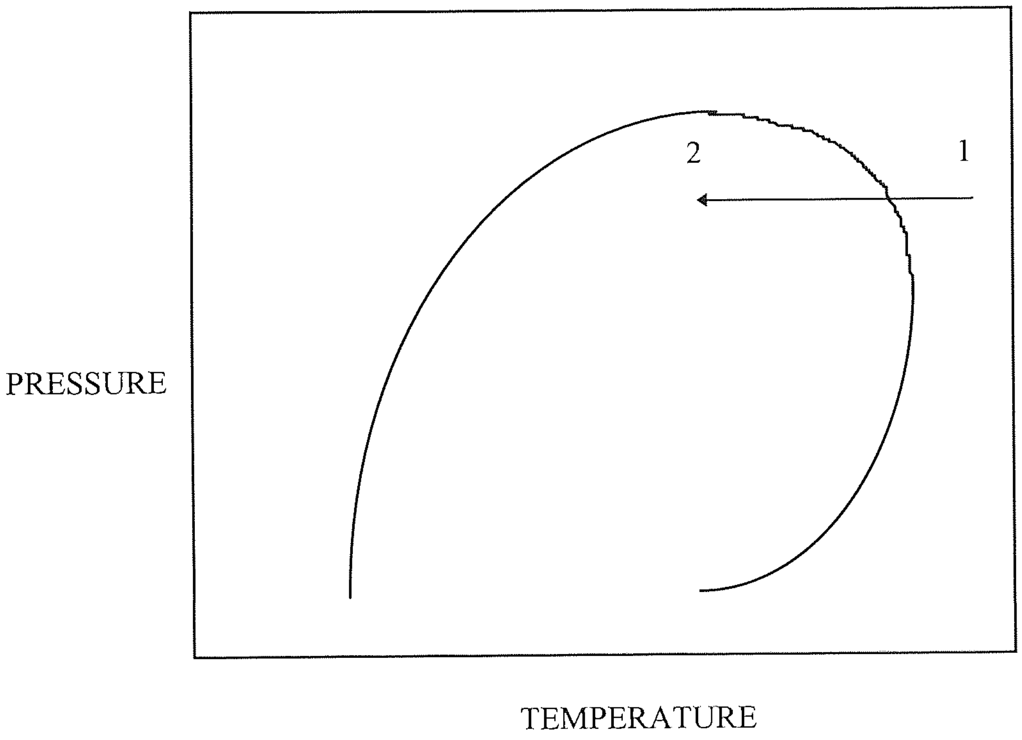
If the tube diameter is approximately four times that of the inlet nozzle, a fast-turning vortex is created in which the centrifugal force is a million times that of the earth's gravity. This high centrifugal force assures a perfect separation of liquid from the gas. The gas-liquid mixture enters, through a channel of gradually decreasing cross-section, tangentially into the cylindrical tube. The molecules that move towards the core have to pass through the high centrifugal force region. Only light specific gravity substances can move to the core. Since there is turbulence some heavies also move. Because of high centrifugal acceleration separation occurs by specific gravity. The liquid is thrown to the periphery and at a distance of one tube diameter, it is taken out. The gas vortex, stripped from its liquid content, continues in the tube where heat is transferred from its central part to the periphery, so that the core will be in the range of hundreds of degrees colder than the inlet, the periphery by about the same amount warmer. If the cold of the gas is utilised in the process, the cold gas is taken out through the diaphragm.

The advantages of using the vortex tube are:

- the gaseous product is superheated and the liquid product is subcooled so that no two phase flow is present in the product lines.
- low size / weight.
- no moving parts.
- insensitive to orientation and movement.
- little or no operating cost.
- can be used for low flow rates.

2.4.4 REFRIGERATION

Refrigeration systems are common in the natural gas processing industry. Refrigeration is an isobaric process, as a result the operating line is virtually horizontal, and cuts the phase envelope and the quality lines at right angles.



Refrigeration is achieved by the use of external cooling, produced by an independent Rankine-type thermodynamic cycles. For moderate negative temperatures, down to -40^{o}C a single propane or freon cycle can be used. Below -40^{o}C, two pure fluid refrigeration cycles operate in cascade. The cycle corresponding to the lower temperature uses a light hydrocarbon such as ethane. Even lower temperatures can be reached with mixed refrigerant systems. The diagram (figure 4) below illustrates a simple refrigeration system.

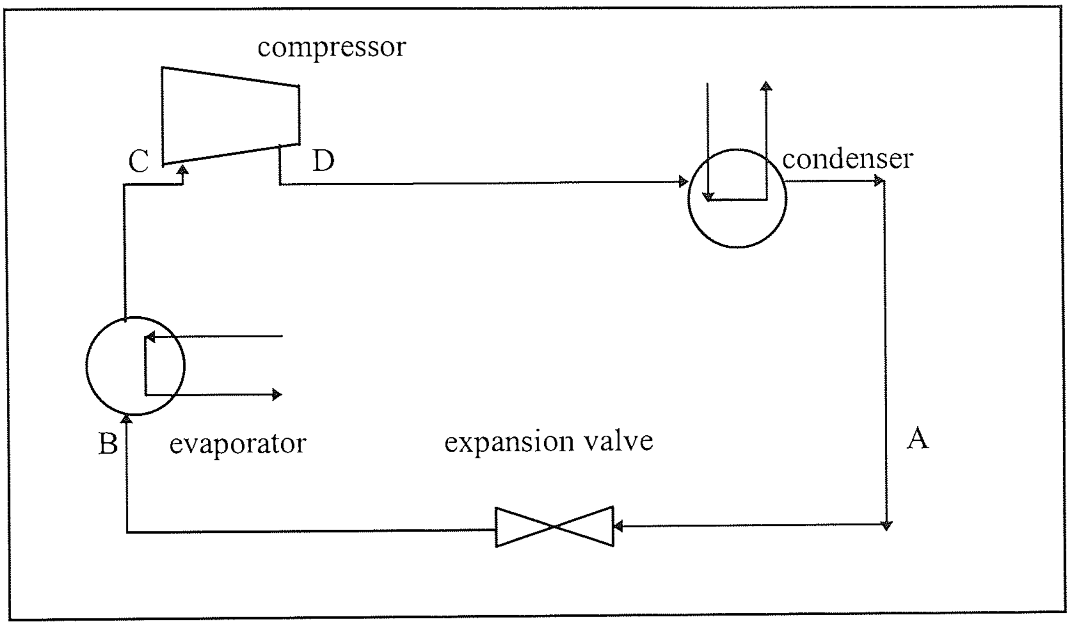


Figure 4 : A simple refrigeration system.

Saturated liquid at point 'A' expands across a valve. On expansion some vaporisation takes place. The refrigerant mixture enters the evaporator (i.e. the cooler) at 3-6^{0}C lower than the temperature to which the process stream is to be cooled. The liquid vaporises leaving at point 'C' as a saturated vapour at the pressure and temperature of the evaporator. This vapour is compressed and then enters the condenser as a superheated vapour. The refrigerant leaves the condenser as saturated liquid or slightly sub-cooled.

It is advantageous to use refrigeration if the process pressure corresponds to the product pressure and the dew point specification can be achieved at a reasonable temperature. The advantage is that no recompression of the product stream is required. The disadvantages are that an external refrigeration package is required and the process is not very suitable for very low temperatures (32).

2.4.5 ABSORPTION

Gas absorption is one of the oldest unit operations used in the processing industries. Rich gas enters the bottom of the absorber and flows upward through the absorber counter- currently to the lean oil. The lean oil preferentially absorbs the heaviest components from the gas and is termed 'rich' oil. The dry (processed) gas leaves the column at the top. Lean oil enters the absorber from the top and flows downward counter-currently to the rich gas. Rich oil leaves the bottom of the absorber and flows from there to the top of the stripper, where the absorbed components are removed by heating and/or stripping with steam. The lean oil from the bottom of the stripper is cooled and then recycled to the absorber and the absorbed material is removed overhead. The absorber-stripper combination (figure 5) are twins, they go together. The absorber recovers components while the stripper removes them, so that lean oil is in the proper condition before it is recycled to the absorber.

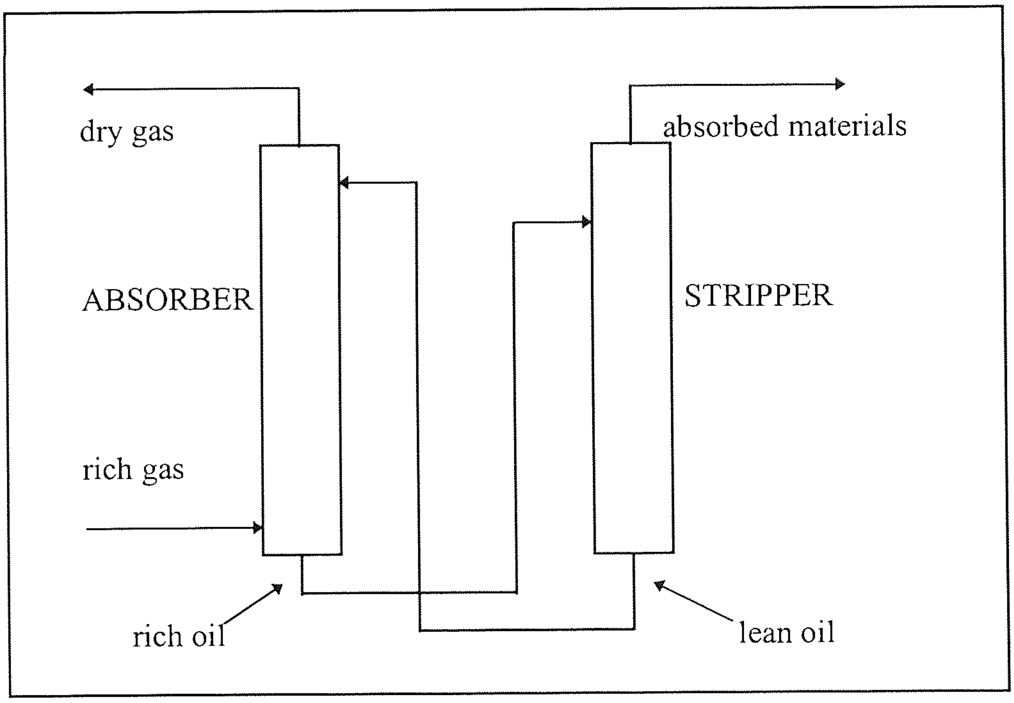


Figure 5 : The absorber - stripper combination.

A component in the gas phase is only absorbed by the oil when its partial pressure in the gas phase is greater than its partial pressure over the liquid phase. The oil acts as a means of lowering the vapour pressure of the liquified components by dilution. The leaner the oil is in the components to be absorbed the more efficient the absorption process. For a given gas rate and composition, the lean oil rate and number of equilibrium stages are the variables. Oil rate and number of stages are interdependent because as oil rate increases it remains leaner for a given absorption and therefore fewer contacts will be needed.

As components are absorbed, the temperature of oil and gas phases will increase due to the release of the heat of absorption. In magnitude, this is slightly greater than the latent heat of condensation. The heat released is proportional to the amount of gas absorbed. In some cases side coolers are used to limit the temperature rise in the absorber.

The lean oil used will have a molecular weight in the 100 to 200 range, depending on the absorber temperature. For ambient temperature absorbers, a heavy lean oil of 180 to 200 molecular weight is commonly used. For refrigeration temperatures, a lighter lean oil of 120 to 140 molecular weight is used. The lower the molecular weight of the lean oil the greater the number of molecules per gallon, resulting in a lower circulation rate. The disadvantage of lower molecular weight lean oil is that losses due to vaporisation are greater. A properly-stripped lean oil should contain few components lighter than pentane. In most circumstances, the lightest possible molecular weight oil should be used that conditions permit. This minimizes the mass to be pumped (33, 34).

However, since the advent of the turbo-expander plants, the absorption process has gone out of fashion due to the high energy requirements to chill the circulating lean oil and condense the recovered products. Furthermore, the ethane and propane recoveries obtained were much less then those with turbo-expander plants.

2.4.6 MEMBRANES

Membranes are essentially thin barriers that allow preferential passage of certain molecules. Gas separation membranes are normally comprised of polymeric materials. These membranes contain no holes or pores. The separation relies on the exploitation of the relative solubilities and diffusivities of different gases in a given membrane material. Therefore certain gases will pass through a membrane at a faster rate than other gases due to a difference in the solubility of the gas in the polymer and the rate at which the gas diffuses through the polymer membrane.

The driving force for the permeation is the difference in partial pressure across the membrane. Figure 6, below illustrates a simple gas separation process. The feed gas at a given pressure passes along one side of the membrane with the other side at a lower pressure. Due to the partial pressure differentials the hydrocarbons will permeate through the membrane to the low pressure side (35). Since the heavier hydrocarbons have higher permeation rates, they will permeate through the membrane, resulting in the low pressure (residual) stream being more concentrated with heavy hydrocarbons and the permeate stream being more concentrated with light hydrocarbons.

For a given temperature, pressure, and feed composition the permeability of an individual component through a polymer membrane is constant. This permeability is expressed in units of component gas flowrate per unit membrane surface area per component partial pressure difference across the membrane.

$$\text{Permeability} = \text{flowrate} / (\text{membrane area x component } \Delta P)$$

Various design parameters such as the feed and product composition, pressure differential, permeate pressure, permeation rates and membrane material will affect the performance of the membrane separation (36, 37).

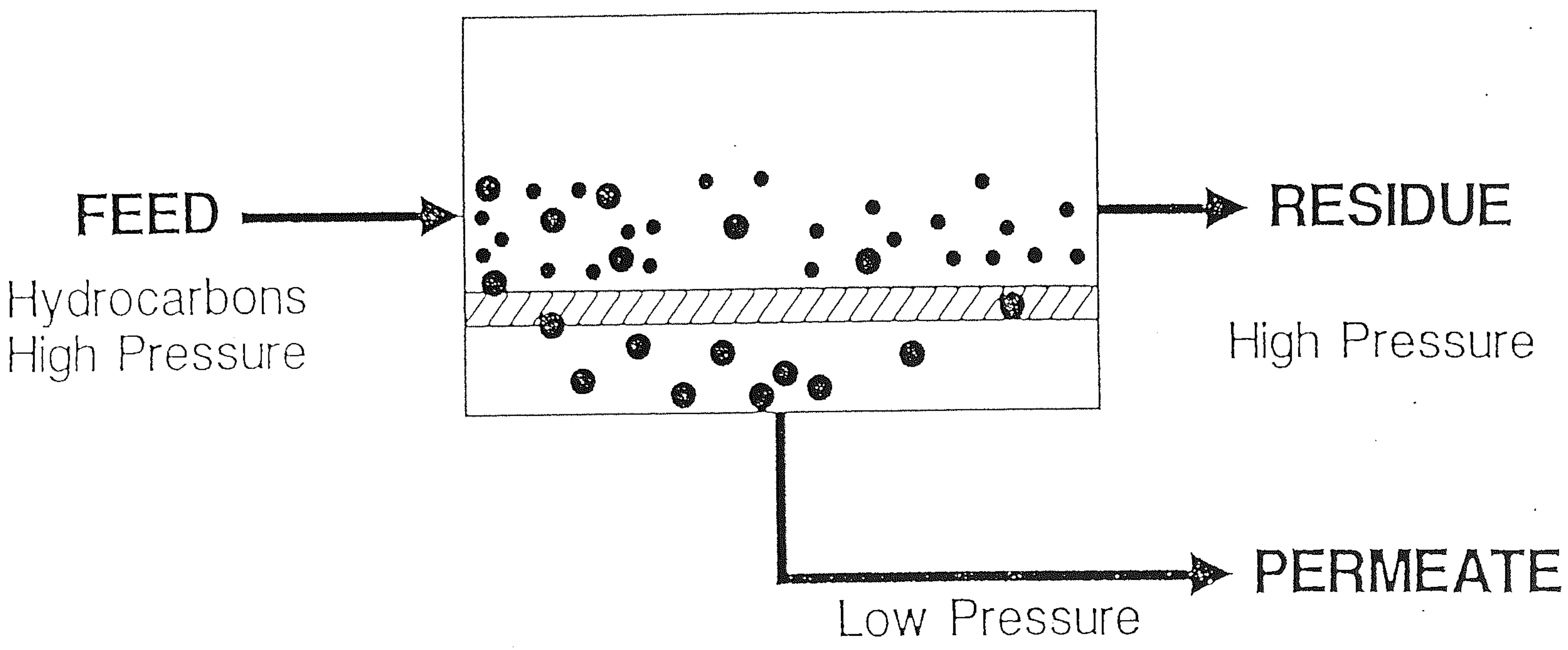


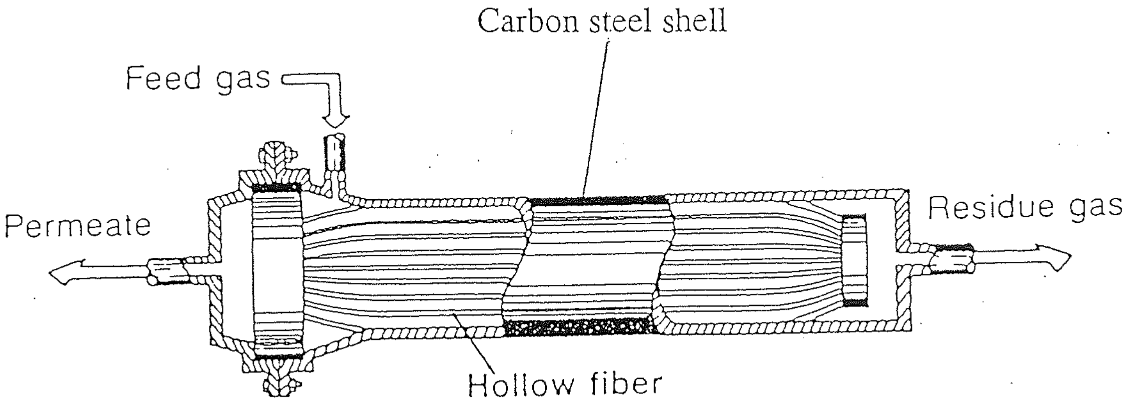
FIGURE 6 : Schematic of a Membrane Separation Process.

Membrane materials are available either as flat sheets or as hollow fibres, which are packaged into membrane modules suitable for the conditions of the gas stream to be processed. The most popular are shell and tube for hollow fibre membranes and spiral wound for the flat sheets (38). In hollow fibre modules (Figure 7, page 40), the membrane surface is in the form of tubes with the high pressure feed gas being introduced either on the shellside of the bundle or through the tube bore. The disadvantage of hollow fibre systems is that the cross-sectional area available for the permeate flow is very limited and for high pressure differentials the permeate has to be on the bore side of the fibre (39). Due to the small diameter of the bore (typically < 300 micron) pressure losses that occur when handling large volumes of high molecular weight permeate gas will greatly reduce the efficiency of the separation. A spiral wound module consists of a series of membrane envelopes that are constructed of two sheets of membrane glued back to back along three edges with the membrane facing outward. A spacer is inserted into the centre of this envelope to keep the membranes separated so that a channel is formed for the permeate gas. The open end of the membrane envelope is attached to a stainless steel tube that has been drilled with holes. Feed gas is introduced to one face of the membrane module where it travels between individual membrane envelopes. The feed channel spacer is designed to promote mixing of the gas as it travels along the membrane into the centre of the membrane envelope and then spiral around to the steel permeate tube and exit as the low pressure permeate stream. The product gas exits at other end of the module with essentially no pressure drop (40, 41).

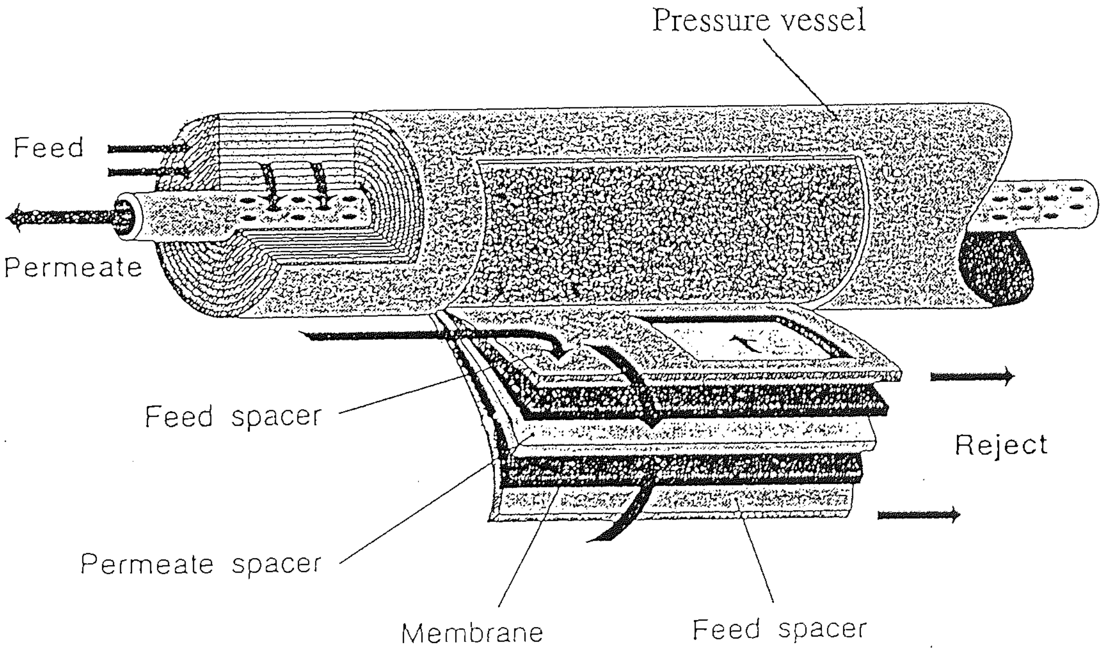
Membranes are able to treat feed gases at pressures higher than the cricondenbar. Feed gas is normally pre-treated before entering the membrane module to at least 10 °C above its dewpoint to ensure no liquid hydrocarbons will separate out in the membrane. The other advantages of the membrane process are:

- ease of installation.
- no operator requirement, since there are no moving parts.

FIGURE 7 : Illustrations of industrial gas membrane separators.



Hollow fibre module



Spiral wound module

- low weight and space requirements.
- modular design therefore highly flexible.
- no utility requirements.

2.4.7 ADSORPTION

Adsorption has traditionally been used for dehydration of natural gas. However, silica gel adsorbent also has the ability to adsorb hydrocarbons (due to its large pore diameter compared to, say, a molecular sieve) and it possesses a greater affinity for the heavier hydrocarbon components. To date the application of adsorption for hydrocarbon dew point control has been limited.

Adsorption is a process where molecules from the gas are held on the surface of a solid by surface forces. The solid desiccants used for adsorption generally have a large surface area due to the presence of capillaries or a crystalline type lattice in their interior. There are number of materials which have the above charateristics, such as alumina, molecular sieves, bauxite and silica gel. The selection depends on a rational compromise between cost and need. The life of these materials is between 3 and 5 years in the absence of poisoning. Service life is limited by a loss of capacity, dusting and breakage (42, 43).

The adsorbent bed may contain more than one adsorbent, such a composite bed will behave like two towers in series. In adsorption columns the flow normally is always downward because of the higher allowable velocity in this direction. The operating temperature preferably should below 50 °C Centigrade and the feed gas must enter the unit above its hydrate point. During normal operation, three zones exit in the bed:

1. Equilibrium zone.
2. Mass transfer zone.
3. Active zone.

In the equilibrium zone the desiccant is saturated with water (and heavy hydrocarbons). It has reached its equilibrium (water) capacity based on inlet gas condition and has no further capacity to adsorb water. Figure 8, shows the mass transfer zone. As the gas enters a dry desiccant bed, all of the absorbable components are adsorbed at different rates. After the process has proceeded for a short period of time, a series of adsorption zones will appear. These zones represent the length of the tower involved in the adsorption of any component. Behind the zone all of that component entering has been adsorbed on the bed. Ahead of the zone the concentration of that component is zero. These zones form and move down through the desiccant bed. Water would be the last zone formed.

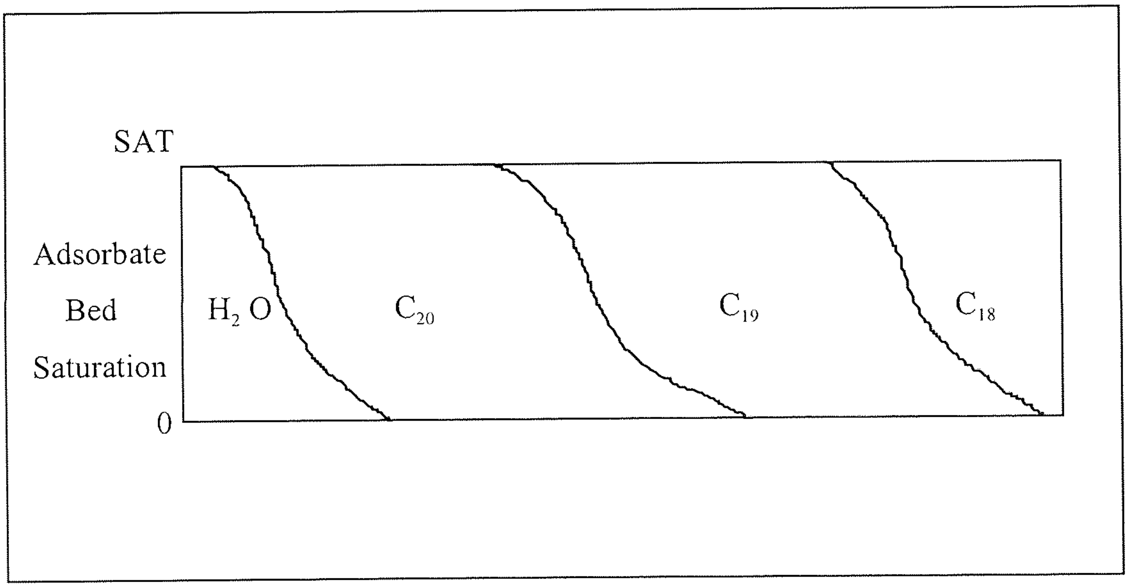


Figure 8 : Schematic portrayal of bed saturations.

All of any component is adsorbed until the front of its zone reaches the outlet of the bed. When the back of its zone reaches the outlet of the bed, no more adsorption of that component will take place. It will furthermore be displaced almost entirely by the component in the zone following it down the bed if the cycle is continued. If the process continues long enough none will remain on the bed.

For hydrocarbon and water dew point control the length of the adsorption cycle is short. In this case, octane breakthrough time represents the maximum cycle time. Therefore the process is limited to relatively lean feed gases otherwise the beds are too large. Breakthrough time depends on gas velocity and bed length.

The adsorption performance is independent of the feed gas thermodynamics, thereby eliminating any risk due to uncertainty in the phase envelope. As a result adsorption is able to process gases even if the process pressure is higher than the cricondenbar.

Most large desiccant units for natural gas processing normally contain more than two towers to optimise the economics. The design variables are the number of towers, cycle time, gas flowrate, unit capacity, dewpoint specification, and regeneration requirements. These are not independent variables. Therefore, some have to be fixed for calculation purposes (44, 45).

2.4.8 REFLUXING CONDENSER

A refluxing condenser, sometimes called a dephlegmator, is a vertical plate-fin heat exchanger in which both heat and mass transfer takes place to effect separation. The feed gas is cooled and partially condensed as it flows up through the feed passages. The resulting condensed liquid runs back down in the same passages as reflux, counter-current to and in intimate contact to the rising vapour (feed gas). The refrigeration for the process is provided by the cold product gas and/or the external refrigerant unit flowing in adjacent passages. In the feed passages, the descending liquid becomes enriched with the heavier components from the vapour stream while the more volatile components are revaporised and stripped from the liquid by the upflowing vapour. Thus the liquid produced is largely stable, thereby reducing the stabilisation load.

The feed cooling process corresponds to a large number of partial condensation steps in series, in which liquid formed at each step becomes partially stripped or stabilised as it flows back down the passages. The exchanger may be designed and operated with very close temperature differences throughout its length. These factors result in high thermodynamic efficiency and, in turn, minimize energy consumption.

The concept of a heat exchanger as a multistage rectification device is not new. However, only in the last 15 or so years has accurate design of reflux condenser, become feasible, helped by the reliable and accurate prediction of thermodynamic data. At present reflux condensers are used for a wide variety of roles such as air separation, recovery of hydrogen, ethylene and natural gas liquid recovery (46, 47, 48, 49, 50).

The refluxing condenser is a constant pressure process and offers advantages if no free pressure drop is available. Furthermore it has excellent turndown capabilities, while being able to tolerate variations in either feed or product specifications. The added advantage of this process are that if it is used for offshore processing, it can give reductions in both space and weight requirements, thereby reducing cost.

Note, for gas processing the plate-fin heat exchanger is usually a brazed aluminium type.

The next chapter looks at process design and simulation.

CHAPTER 3

3.0 INTRODUCTION TO PROCESS DESIGN AND SIMULATION

“Make sure that you take the ‘Star Tours’ ride when you visit Disneyland” an American friend insisted. “It’s as good as the real thing, and a lot cheaper!” I can echo this tribute to simulation (51).

Simulators of one kind or another have been used since the 1950s to solve mathematical models of processes in chemical engineering (52). In recent years, as a result of computers becoming significantly more powerful, the newer simulators can handle steady state and dynamic operations in the same package.

The driving force behind these developments has been market forces. To increase safety and reduce environmental impacts, achieve greater efficiencies through process modifications, conserve energy and material resources, size equipment and study the effect of variation in the design and operating parameters on the performance of the plant, there is no alternative to the use of a process simulator.

Today no major chemical or petrochemical plant would be built without first simulating the process (53, 54). Simulators are produced either for general purpose use or as special-purpose one to handle only very specific problems.

In the following sections various aspects of steady-state process simulation will be introduced and some of the issues involved in the use of a typical simulator will be looked in brief.

3.1 PROCESS SIMULATION ARCHITECTURES

The basic features of a process simulation are that a mathematical model of the process is required and a means of solving it. The mathematical models consist of mass and energy balances related to process and equipment parameters (Fig. 9). Such models can be solved in two ways, depending on what is known and what has to be calculated. In a "design mode" the model inputs and/or design parameters are calculated from the output specifications (figure 10). The simulation carried out in "design mode" is numerically less stable than that in the "analysis or performance mode" (55). The "performance mode" calculates the outputs from the given inputs and operating conditions (figure 11). Hence, the majority of the simulation packages use "performance mode" in process simulation.

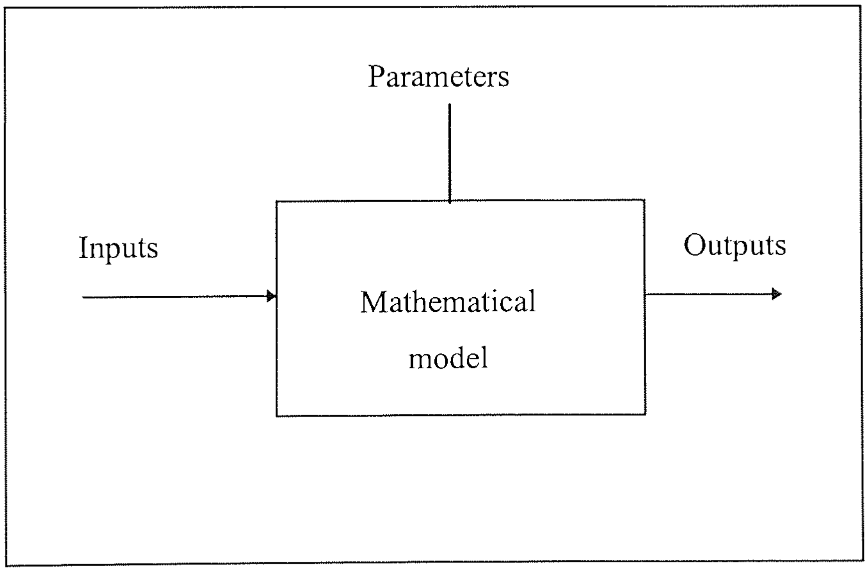


Figure 9 : Mathematical model of a unit operation.

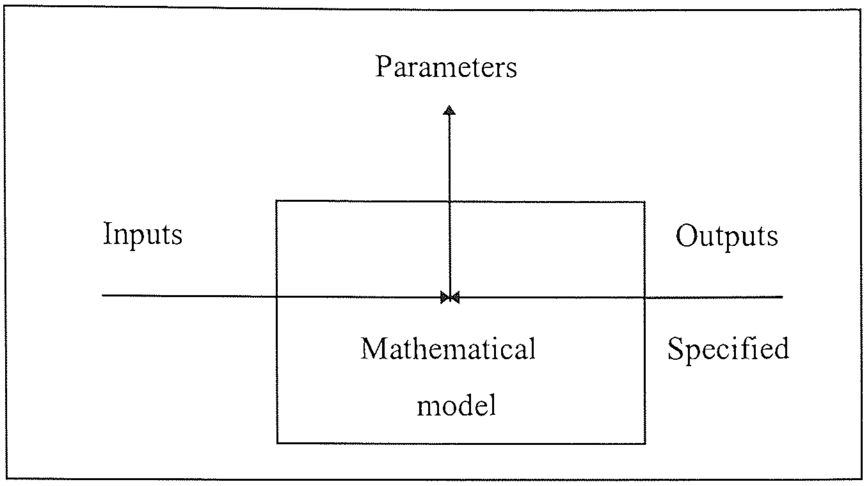


Figure 10 : Design Mode.

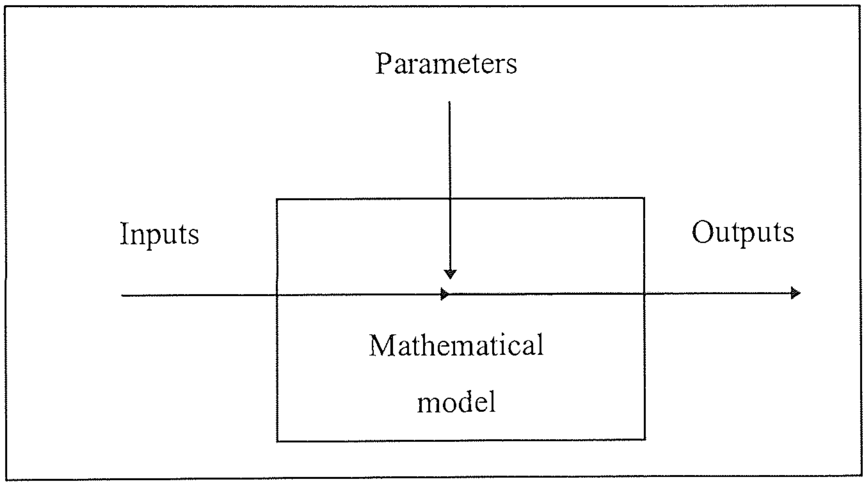


Figure 11 : Performance Mode.

Two main methods of simulating complete process flowsheets will now be discussed

3.1.1 SEQUENTIAL-MODULAR APPROACH

The sequential-modular method approach implements the unit operation modules as computer subroutines that calculate the output variables as functions of the input variables. The calculation moves from module to module, starting from the feed streams until products are obtained (figure 12).

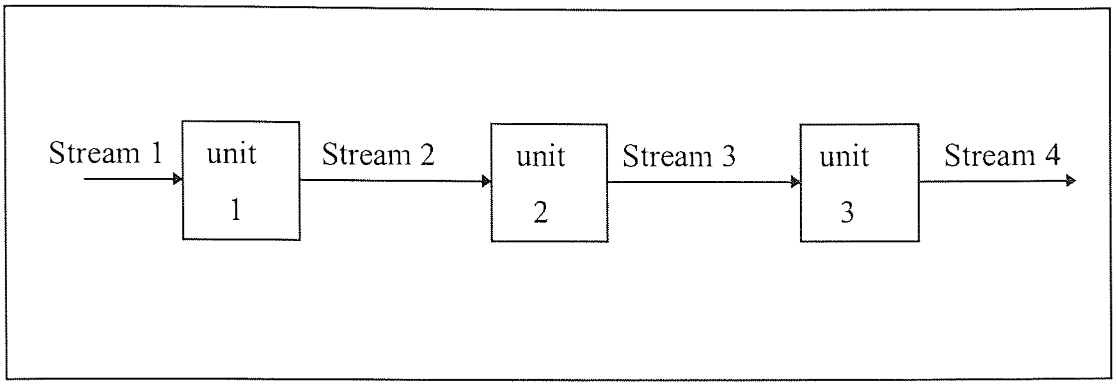


Figure 12 : A simple series process.

Unfortunately, processes involving recycles of mass and/or energy present a problem, since it represents an interlinking network of modules. To solve this problem, the system needs to partition the flowsheet, select "tear" streams, nest convergence of the tear streams, and determine the computational sequence. For example (figure 13), the value of the recycle (stream 7) is estimated, thus tearing the recycle stream. The model is then be solved sequentially to give a result for the recycle stream. The result is then used to solve the model again. This iterative procedure continues until the process converges.

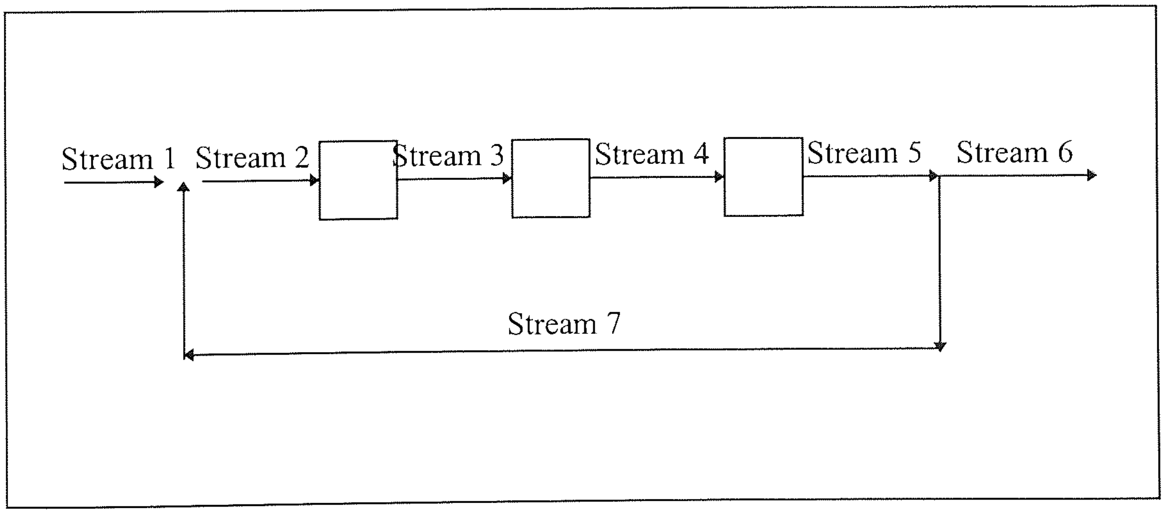


Figure 13 : A simple recycle process.

A general purpose sequential modular program, used by a majority of the steady-state process simulator applications (including BP GENESIS), has the following structure:

- A store of physical and thermodynamic properties for pure chemical components and their mixtures.
- A system working data-base derived from the data obtained from the store as well as the problem input.
- Individual module data base derived from the working data base.
- Modules or subroutines for individual process units, computational procedures including input / output for each module and internally iterated variables, if any.
- Flowsheet topology to be used by the system executive to order the computation sequence.
- Routines for recycle calculations and convergence analysis.
- Other mathematical service routines.
- An interface to the user, reporting and printing facilities.

3.1.2 EQUATION-SOLVING APPROACH

In the equation-solving approach all the equations describing the flowsheet are collected together into a global model and solved simultaneously. The advantage of this approach is its flexibility in that any number of constraints can be added as equations to the set defining the problem. The corresponding model can be used to analyze the steady-state or dynamic behaviour, or optimise the design and operation, of a process.

3.2 CHOICE OF THE SIMULATOR

There are many commercially available simulation packages on the market that may be adequate for modeling chemical engineering processes. However, to select the best package, the user needs to consider carefully the simulator's capabilities for meeting the requirements of modeling the respective processes. Points to consider are:

- The cost of using the package for research purposes.
- Unit operations available.
- Quality of the component data base.
- Thermodynamic and physical property estimation methods.

User friendliness is another important factor that should be considered. Similar thought should be given to the level of user support the supplier is willing to offer, since it is likely that the user will at some time or other require assistance.

A list of some of the simulation packages which are currently available is presented in Table 2, below.

Table 2: Process Simulation and Design packages.

Package	Supplier	Comments
HYSIM	Hyprotech Ltd	steady state simulation
HYSIS	Hyprotech Ltd	steady state & dynamic simulation
CHEMCAD III	Chemstations Inc.	steady state simulation
PRO/II	SimSci	steady state simulation
GENESIS	BP	steady state simulation

The simulation package chosen for this work is BP's in-house generalised simulation system, GENESIS. It can be used for a wide variety of applications from simple flash calculation to the simulation and optimisation of complex flowsheets with multiple recycle streams. The convenience of using this package is that it is compatible with BP's proprietary process engineering software applications. This is of particular importance when modeling novel technologies such as membranes, vortex tubes, etc.. The added advantage of using BP GENESIS was that, an active back-up and development team to provide day to day training and consultation to users was available.

The features of the BP GENESIS simulation package are briefly described in the next section.

3.3 GENESIS FLOWSHEET SIMULATION

The starting point for flowsheet simulation is to draw a process-flow diagram of the relevant process of the type shown in figure 14, below.

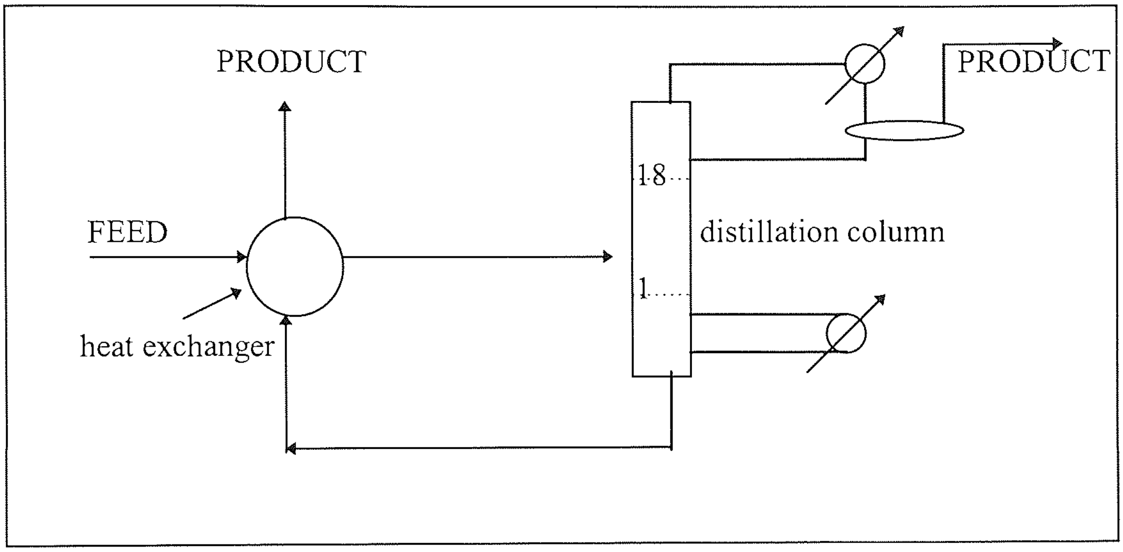


Figure 14 : Process flow diagram.

The next step is to construct a model of the process (figure 15), by modeling each equipment item or processing step in the process in terms of the GENESIS modules that represent the required operations. The stream connections between the modules must be specified.

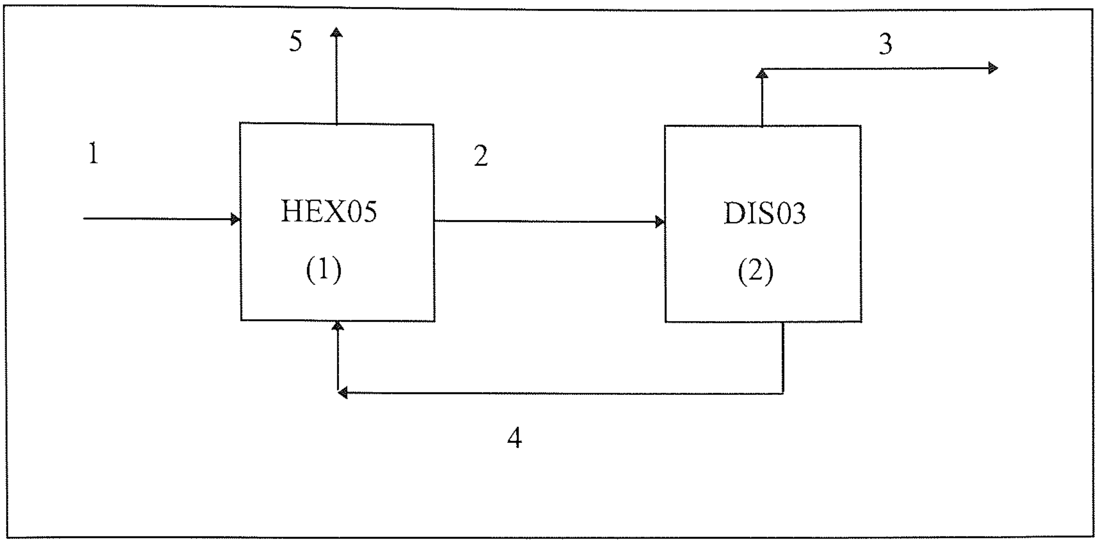


Figure 15 : Model of the process.

The next step involves specifying the components in the system to be simulated, the mathematical models to be used for the K-value and thermal property calculations, the composition and condition of the feed streams. The simulator then asks for the module input data.

On specifying the module inputs, the interactive input sequence for the flowsheet has been completed, and the flowsheet is ready to be run.

The output from the simulation run includes tables with stream data, operational data of the different equipment modules which is not available in the stream summaries, error messages and warnings about certain difficulties are printed.

3.3.1 CAPABILITIES AND LIMITATIONS

The extraordinary processing power and speed of today's computers allow the understanding and manipulation of highly complex, multi-variable systems, a task that otherwise would be near impossible. Computer simulation permits processes to be modelled and optimized simultaneously for various variables, such as profitability and environmental factors. Although widely used, simulation has not reached its full potential in many areas yet. To improve reliability, significant advances need to be made in the field of physical property prediction, steady-state and dynamic modeling of systems and across the board modeling of unit operations. The simulation results produced depend on the model supplied and are therefore only as accurate as that model.

The next chapter looks at the models used to simulate the process technologies.

CHAPTER 4

4.0 PROCESS MODELLING

4.1 INTRODUCTION

In this chapter how the different gas separation processes being considered for evaluation are modelled will be discussed.. The calculations have been carried out within the BP GENESIS simulator. The models used are a virtual representation of the real system(s). They can therefore, be used in conceptual process design (53), for the feasibility analysis of traditional and novel process technologies, technical and ergonomic evaluation of design alternatives. They may also be used for strategic planning, correlation of cost data from the economic evaluation, etc.

Note that before any simulation can be carried out by the GENESIS process simulator, a description of the flowsheet in terms of the equipment to be simulated (membrane, flash drums, etc.), the stream connections between them and the process conditions is required. In the flowsheet each piece of equipment corresponds to the BP GENESIS module that will represent the operations required.

4.2 VLE CALCULATIONS WITHIN THE TWO PHASE REGION OF THE PHASE ENVELOPE

Flash calculations are essential to almost all the process modules in BP GENESIS (56). The purpose of the flash calculation is to determine the amounts of vapour and liquid in the feed and their corresponding properties. There are essentially three modes of operating the flash module:

- Flash at a given temperature and pressure,
- Calculate temperature for specified conditions and

- Calculate pressure for specified conditions.

For modes 2 and 3 the flash calculation can be carried out at a given value of one of the following:

fraction vaporised
fraction liquified
enthalpy rate
entropy rate
vapour flowrate
liquid flowrate
specific enthalpy
specific entropy

Thus flash calculations, including adiabatic, isothermal and isentropic, can be performed by selecting the appropriate mode. In addition calculations can be performed for bubble point or dew point conditions.

The schematic view of the equilibrium separation modules used for the calculations in the two-phase region of the phase envelope is shown in figure 16, page 56. The main pieces of equipment are the slug catcher and the two flash drums. All three pieces of equipment are modelled using the flash module. As the feed gas enters the slug catcher it removes any condensed liquid present. The gas then enters the flash drum where it is flashed at a given temperature and pressure, using module 1. The condensate streams 3 and 5 are combined, before they enter the flash drum to remove any remaining gas.

Once all the relevant data has been entered, the simulation run is iniated. At the end of each simulation run, a summary of all the streams in the flowsheet is produced. Examples of typical stream results for a simulation run are given in appendix A1-2. If however a customised report is needed, then the reporter module, RPT01, can be used at the end of the flowsheet.

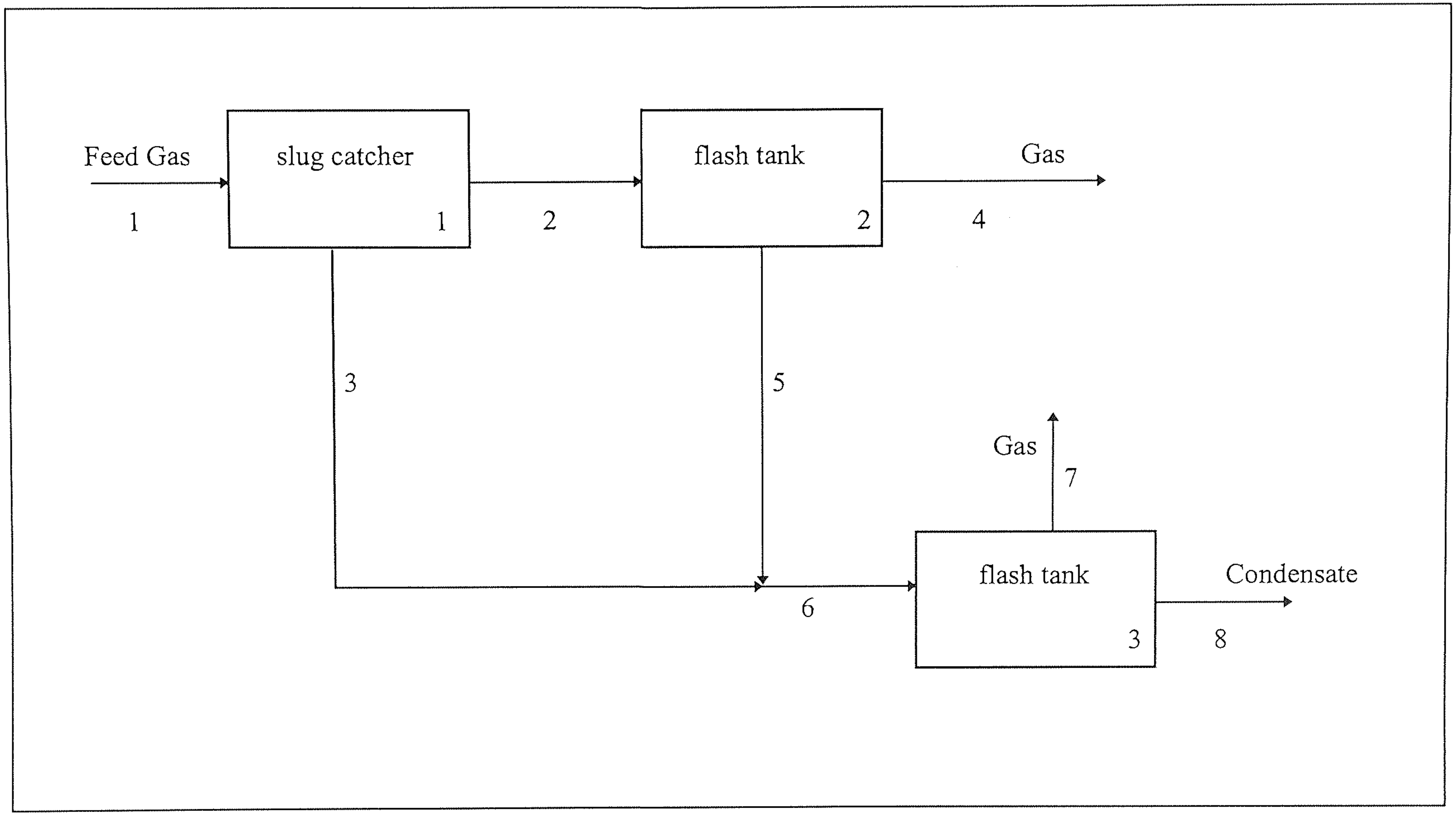


Figure 16: Schematic view of flash modules for VLE calculations.

Using this method the amounts of liquid and vapour at any pressure and temperature condition within the two-phase region of the phase envelope may be calculated.

4.3 PHASE ENVELOPE GENERATOR

The phase generator module is usually located at the end of the flowsheet to calculate a phase envelope on any of the streams in the flowsheet. The phase envelope is generated by carrying out a series of flash calculations over the entire range of conditions for the particular composition in the stream. The initial pressure point and the interval between flashes of pressure may be specified.

The final results show the phase envelope, temperature, pressure, the total mass and energy balance and compressibility factor for each flash calculation.

An example of the output is given in appendix A1-3.

4.4 JOULE-THOMSON PROCESS

The flow scheme used to simulate the Joule-Thomson process, is shown in figure 17, page 58. The feed gas from the plant inlet facilities enter the gas/gas exchanger, where the cold product gas is used to cool down the feed gas. Any condensate from the cooled feed stream is removed in the Knock Out drum (KO). The cooled gas is then expanded through the Joule-Thomson valve. The expanded gas then flows to the cold gas separator, where the condensed heavy hydrocarbons are separated from the product gas. The product gas is then used to cool the feed gas. The condensate streams 3, 6 and 9 form the feed to the condensate stabilisation section.

Approach temperatures of 5°C were used in the gas/gas exchanger in simulation.

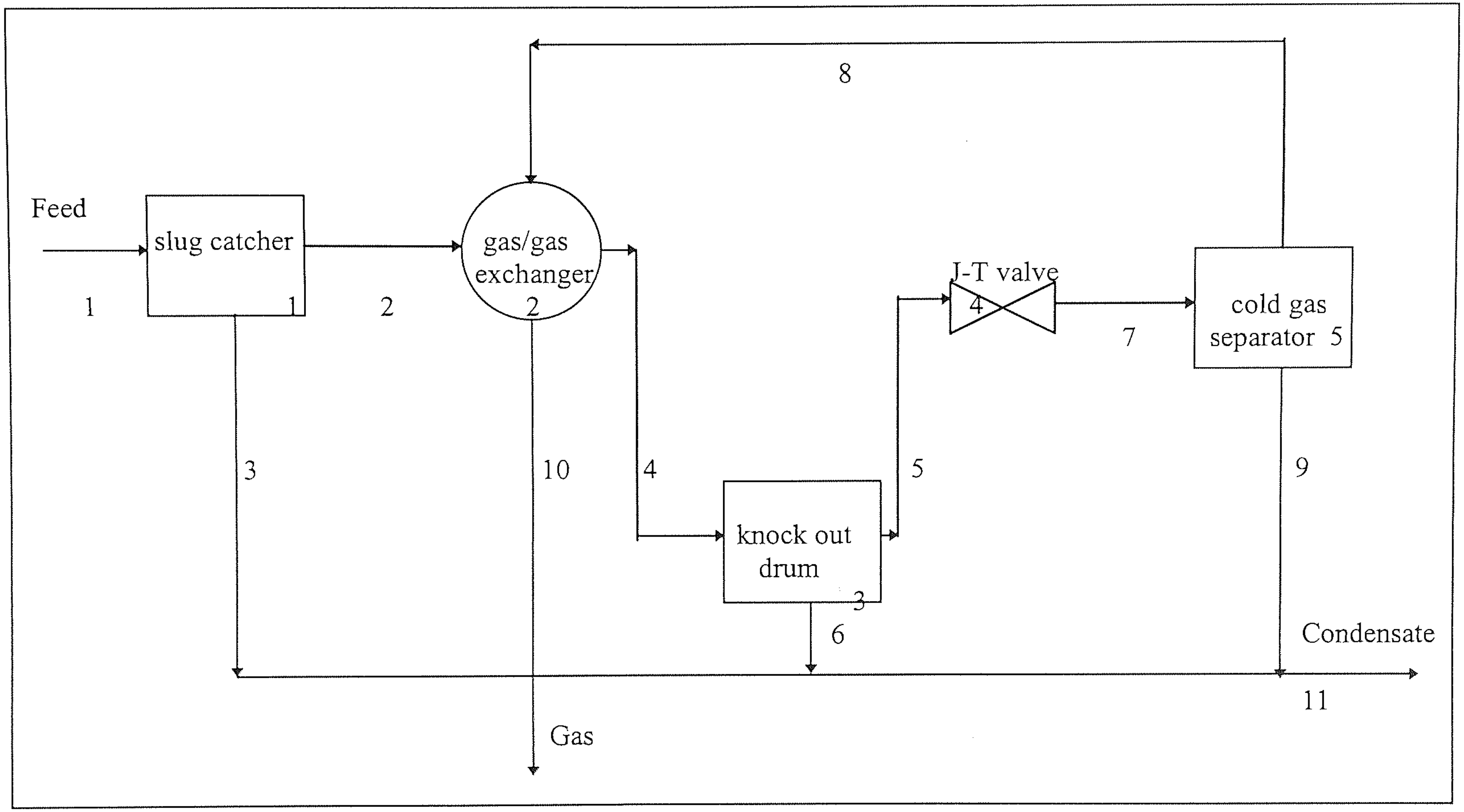
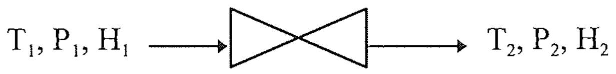


Figure 17: Joule-Thomson process flow scheme.

Modules 1, 3 and 5 are modelled as flash tanks at the feed inlet temperature and pressure condition. Note, that module 1, the slug catcher is considered a part of the process's inlet facilities.

Module 4, simulates a valve by an isenthalpic flash calculation of its input feed stream. For the simulation either the pressure drop or the downstream pressure is supplied. The computer solution to the model is as listed below:

Since the process is considered isenthalpic (adiabatic), no heat exchange between the gas and the surroundings occurs, the First Law of Thermodynamics simplifies to:



$$\Delta H = 0$$

thus

$$H_1 = H_2$$

The following calculation method is used to determine the outlet conditions:

- Determine the enthalpy of the feed stream.
- Assume the outlet temperature T_2.
- Do a flash calculation at P_2 and T_2 to find relative amount and analysis of each phase.
- Find the total enthalpy at point 2 from the above flash at the assumed temperature T_2.
- If $H_1 = H_2$, then the assumed temperature T_2 is correct.
- If not, steps 2-5 are repeated until the enthalpies are equal, within the desired limits of accuracy.

4.5 GAS/GAS HEAT EXCHANGER

The heat exchanger module HEX05 is used to simulate any gas/gas exchangers used in the process flowsheet. The model assumes a shell-and-tube heat exchanger with 1 counter-current shell having 2 tube passes, the shellside calculated first.

The performance of the heat exchanger can be modelled either as a total duty or as any of the following outlet stream temperature conditions:

- Temperature,
- Bubble-point,
- Dew-point,
- Mole fraction vaporised and
- Weight fraction vaporised.

For all cases evaluated, the outlet stream temperature is specified, with a minimum temperature difference of 5°C between inlet and outlet streams. Note that a pressure drop of 0.5 bar for each side is also specified.

An example of the output file is presented in Appendix A1-4, it shows the calculated heat duty, corrected LMTD and the LMTD correction factor.

4.6 TURBO-EXPANDER PROCESS

The simulation model (figure 18, page 61) of the turbo-expander only differs from that of the Joule-Thomson process in that module 4, the J-T valve, is replaced by an expander-compressor module.

As with the Joule-Thomson process, the feed gas entering the gas/gas exchanger is cooled by the product gas from the cold gas separator. The condensed heavy

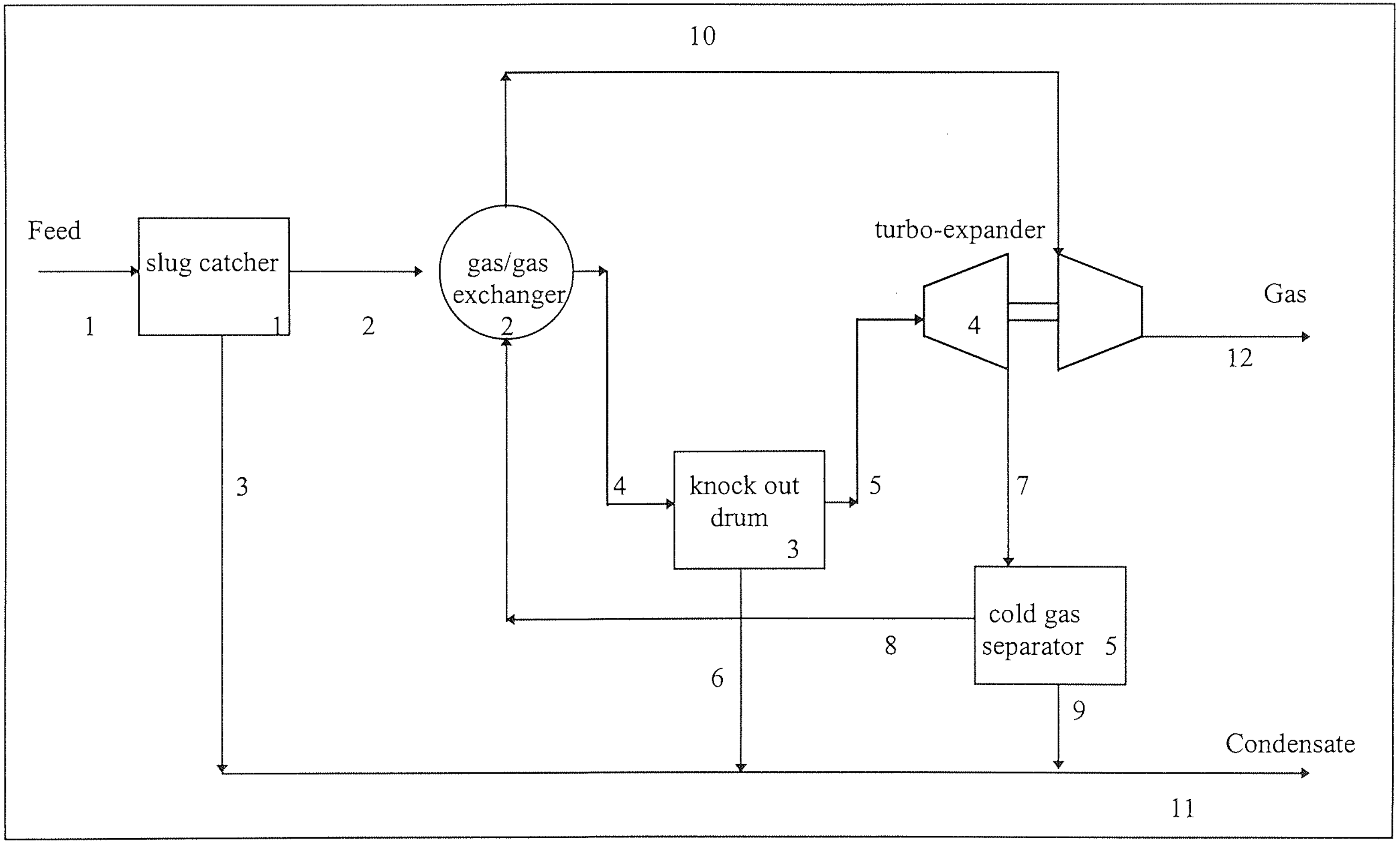


Figure 18: Turbo-expander process flow scheme.

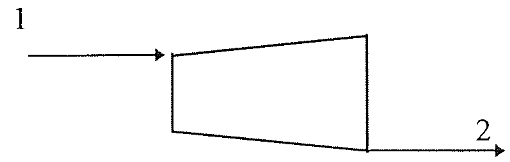
hydrocarbons are removed from the stream in the expander knock-out drum. Even if no condensation takes place the knock-out drum is included to makes sure that no liquid enters the expander. The cooled gas is then expanded through the Turbo-expander. The expanded gas flows to the cold gas separator, where the condensed liquids are separated from the product gas. The gas is then used to cool the incoming feed in the gas/gas exchanger.

The work produced by the feed gas in the expander can be used to recover partially the pressure of the product gas or to generate power. In the simulation, if the pressure of the gas in stream 10 is below the required limit, than the compressor side of the Turbo-expander is used to recompress the product gas after it leaves the gas/gas exchanger.

The turbo-expander (and compressor) module 4, is simulated as an isentropic operation (57). The required input specification can be any of the downstream pressure, the pressure drop (increase in pressure for a compressor) and the shaft work. The calculation results will show the output pressure, the shaft work and the stream conditions.

Turbo-expansion is an isentropic and adiabatic process. Thus the First Law of Thermodynamics reduces to :

$$H_1 - H_2 = -W$$



where H represent enthalpy and W represents work recovered by the turbine.

The following calculation procedure is used to evaluate the work required and determine the outlet stream conditions:

1. Determine the enthalpy H_1 and entropy S_1 of the inlet stream.
2. Assume the outlet temperature T_2.
3. Do a flash calculation at the assumed T_2 and known P_2 to establish if liquid is formed.
4. Calculate enthalpy H_2 and entropy S_2.
5. If $S_1 = S_2$, then the assumed temperature is correct. If not, then steps 2-4 are repeated until $S_1 = S_2$.
6. Once $S_1 = S_2$, then $H_1 - H_2 = \Delta H$ for isentropic process.
7. Calculate actual $\Delta H_{act} = (E) (\Delta H_{theor})$
8. Calculate actual work $W_{act} = m (\Delta H_{act})$, where m is the flow rate.

4.7 REFRIGERATION PROCESS

The process model of a refrigeration process used, is shown in figure 19, page 64. Here the slug catcher is used to remove any condensed liquids from the feed stream. The feed gas then enters the gas/gas exchanger where it is cooled by the cold product gas. The cooled gas then passes to the gas chiller, where further cooling takes place. The propane refrigerant unit is used to achieve a low chiller temperature. The cooled gas flows to the cold gas separator, where the condensed liquids are separated from the product gas. The gas is then used to cool the incoming feed in the gas/gas exchanger.

Module 1 and 4 are essentially flash tanks, while module 2 represents a gas/gas heat exchanger. The calculation method of these modules have already been set out above.

The refrigeration unit is represented by the gas chiller in the flowsheet. The refrigeration unit (Figure 4, page 34), is simulated by the cooler module in BP GENESIS, to carry out a rating calculation. The performance of the module may be specified either as a total duty or as an outlet stream temperature condition. In every case the calculation proceeded until a 5°C approach temperature was achieved in the gas/gas exchanger to obtain the maximum benefit.

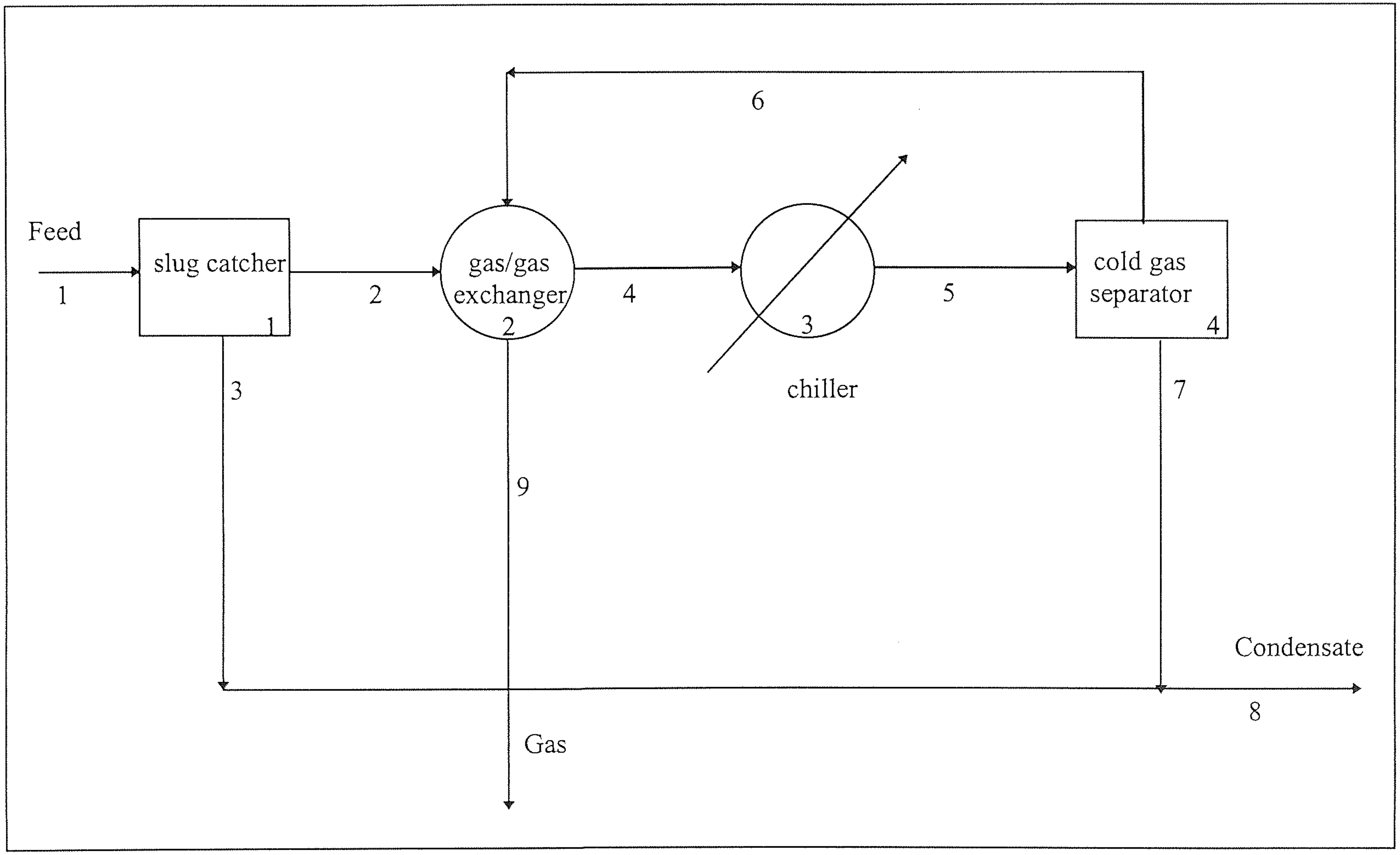


Figure 19: Refrigeration process flow scheme.

Refrigeration is an isobaric process and as such the calculation method used for the refrigeration unit (chiller) is much more straight forward than for either the J-T valve or the turbo-expander. The temperature of the process stream in the cooler is fixed so that single flashes suffice.

1. Determine the enthalpy H_1 of the inlet stream.
2. Set the outlet temperature T_2.
3. Do a flash calculation at the set T_2 to establish if sufficient liquid is formed.
4. Calculate enthalpy H_2.
5. Calculate cooler duty $Q_c = H_1 - H_2$.

4.8 MEMBRANE PROCESSES

The aim was to model a solubility membrane which preferentially allowed the more soluble heavier hydrocarbon molecules to pass through whilst restricting the passage of lighter less soluble molecules through the membrane. The flow diagram of the whole process is shown in figure 20, page 66.

Feed gas entering the membrane passes through the filter coalescer to remove any liquids. The dried gas then enters the membrane module, where the heavy and the light components are separated into the residue and the permeate streams. The residue (product gas) stream, containing mainly the light components, goes to the next stage, for example the sales gas terminal. The permeate stream is flashed to separate the gas from the liquid. The gas stream is then compressed and condensed before joining the liquid stream which has been pumped up to the same pressure. The combined stream then passes to the condensate stabilisation section.

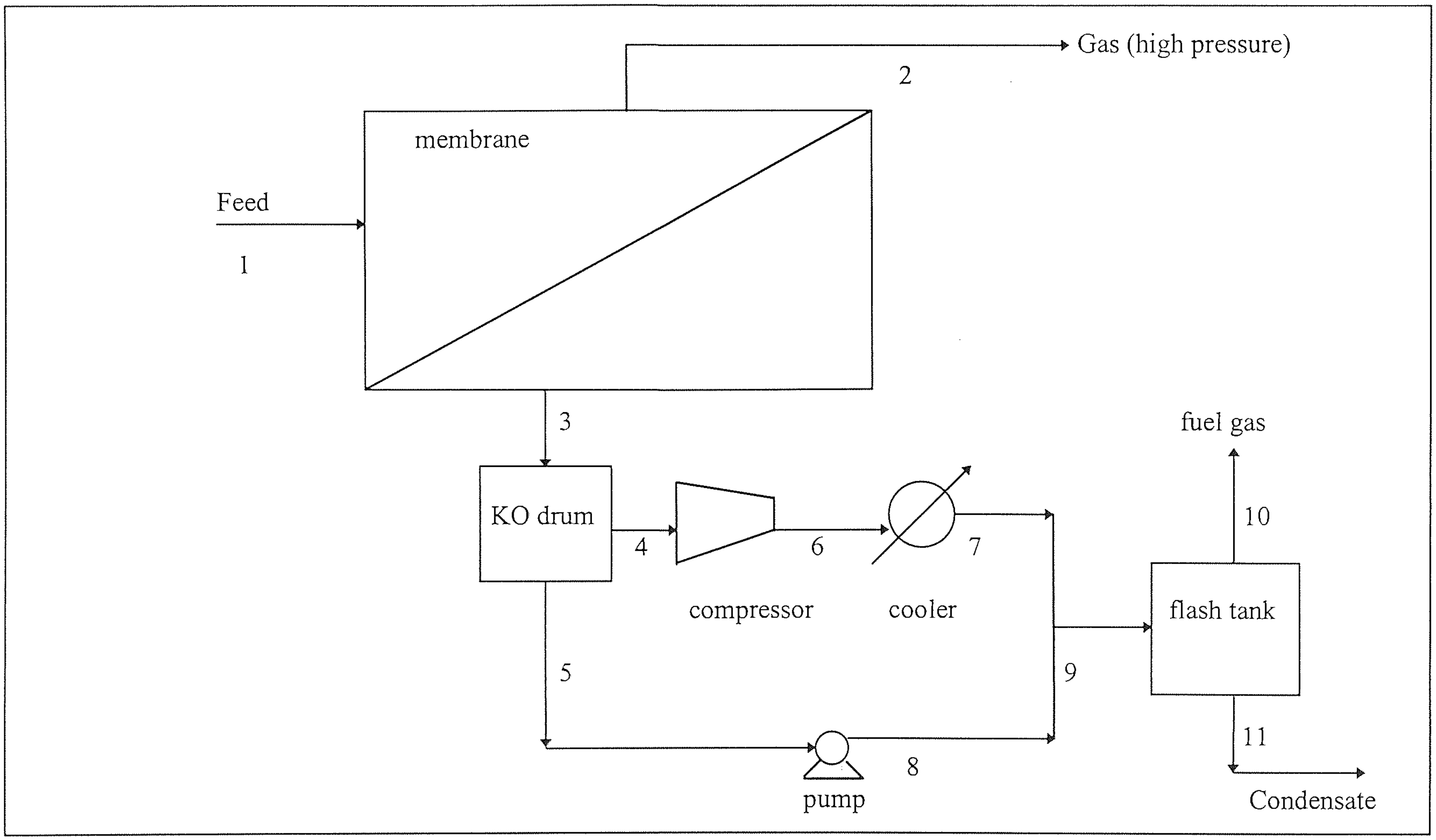


Figure 20: Membrane process flow scheme.

Note, the membrane module is simulated with a permeate pressure of 2 bar. Thus the permeate stream needs to be compressed before it passes to the condensate stabilisation section.

The permeation rate data for the model were supplied by BP Exploration. The data were limited to those for methane, ethane, propane, butane, pentane and nitrogen. The data for the hydrocarbons were analysed and hence, extrapolated for components up to nC_6. Further extrapolation was not necessary because the membrane module in the BP GENESIS is limited to a maximum of 30 components. Accordingly the simulation of the membrane process is not as accurate as it is for the other processes such as refrigeration, and the turbo-expander process. This is because not only the isomer components are coupled in with the normal components but the nC_6+ components are added to the nC_6. This mean that the calorific value, Wobbe number and the phase diagram of the product stream may well be affected. Furthermore, due to the proprietary nature of the membranes, the permeation rate data used will not been disclosed in this work.

The membrane module is able to calculate the area and the operating efficiency or amount and composition of the permeate gas, if the inlet conditions and the permeation coefficients are known. The following inputs are required before running the module.

1. Permeation coefficients.
2. Permeate pressure.
3. Estimation of the membrane area.
4. Specification of key component(s) in the product stream.

The module varies the area until the key component specification is obtained. The permeability of the membrane for each component is assumed to be independent of pressure and composition. The model also assumes no pressure drop on either side of the membrane, isothermal flow and an ideal gas mixture.

4.9 VORTEX TUBE PROCESS

The flow diagram of the Vortex tube with its gas/gas exchanger is shown in figure 21, page 69. Unlike some of the processes above, no cold gas separator is required to separate the condensed heavy hydrocarbons from the product gas.

The Vortex tube is modelled using Isenthalpic and Isentropic principles, with the addition of control modules incorporating design equations.

Cheriyan (58) suggests that the cold end temperature of the vortex tube is proportional to the ratio of the feed to the cold end flow rates and the log of the ratio of the inlet to outlet pressure, $\log_{10} (P_{in} / P_{out})$. The two equations below form the heart of the vortex tube model.

$$K = (T_{in} - T_c) / (T_{in} - T_{isen})$$
$$K = -0.285052\ Fc + 0.086506\ \mathrm{Log}_{10}(P_{cold} / P_{in}) + 0.539034$$

where:

K	Performance parameter
T_{in}	Inlet temperature
T_c	Cold product temperature
T_{isen}	Isentropic temperature for expansion
P_{in}	Inlet pressure
P_{out}	Outlet pressure
F_c	Fraction of cold end gas

But for propriatory reasons further details of the model cannot be released.

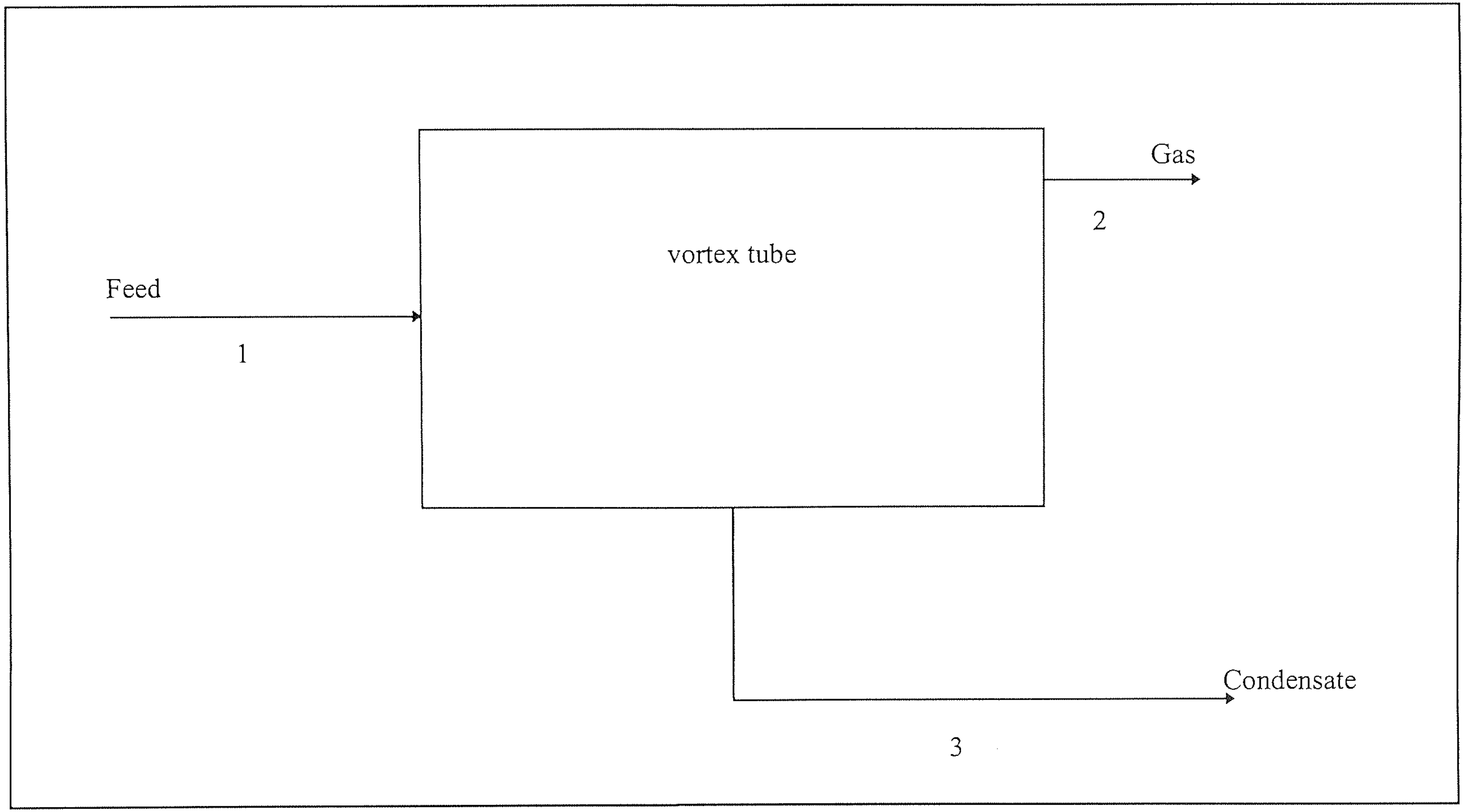


Figure 21: Vortex tube process flow scheme.

4.10 ADSORPTION PROCESS

Unlike the above processes, there is no gas Adsorption model within BP GENESIS that could be used to simulate the process. Review of a number of other simulation design packages, such as HYSIM, ChemCad, etc., revealed that a computer simulation model for the process was not available. This may be due to the limited application of adsorption to hydrocarbon dew point control to date.

The literature review (page 42) on the characteristics and performance of the process states that "All of any component is adsorbed until the front of its zone reaches the outlet of the bed. When the back of its zone reaches the outlet of the bed, no more adsorption of that component will take place (59)". After numerous discussions about the process and its characteristics with engineers at BP Exploration, it was decided that due to the very selective nature of the process at separating the heavier molecules from the feed gas stream the process could be modelled simply by removing the required components from the feed gas to represent the product gas stream. For example, if the product specification is achieved by the removal of nC_8+ components, then the product gas stream will contain feed gas minus the nC_8+ components. The resultant condensate stream will contain only the adsorbed nC_8+ components. The product streams are then used as the basis for calculating the phase envelope, the calorific value and Wobbe number.

4.11 ABSORPTION PROCESS

The simulation model of the absorption process, is shown in figure 22, page 71. Rich feed gas enters the bottom of the absorber and flows upward through the absorber counter-currently to the lean oil. The lean oil preferentially absorbs the heaviest components from the gas and is termed 'rich' oil. The dry (processed) gas leaves the column at the top. Lean oil enters the absorber from the top and flows downward counter-currently to the rich gas. Rich oil leaves the bottom of the absorber and flows

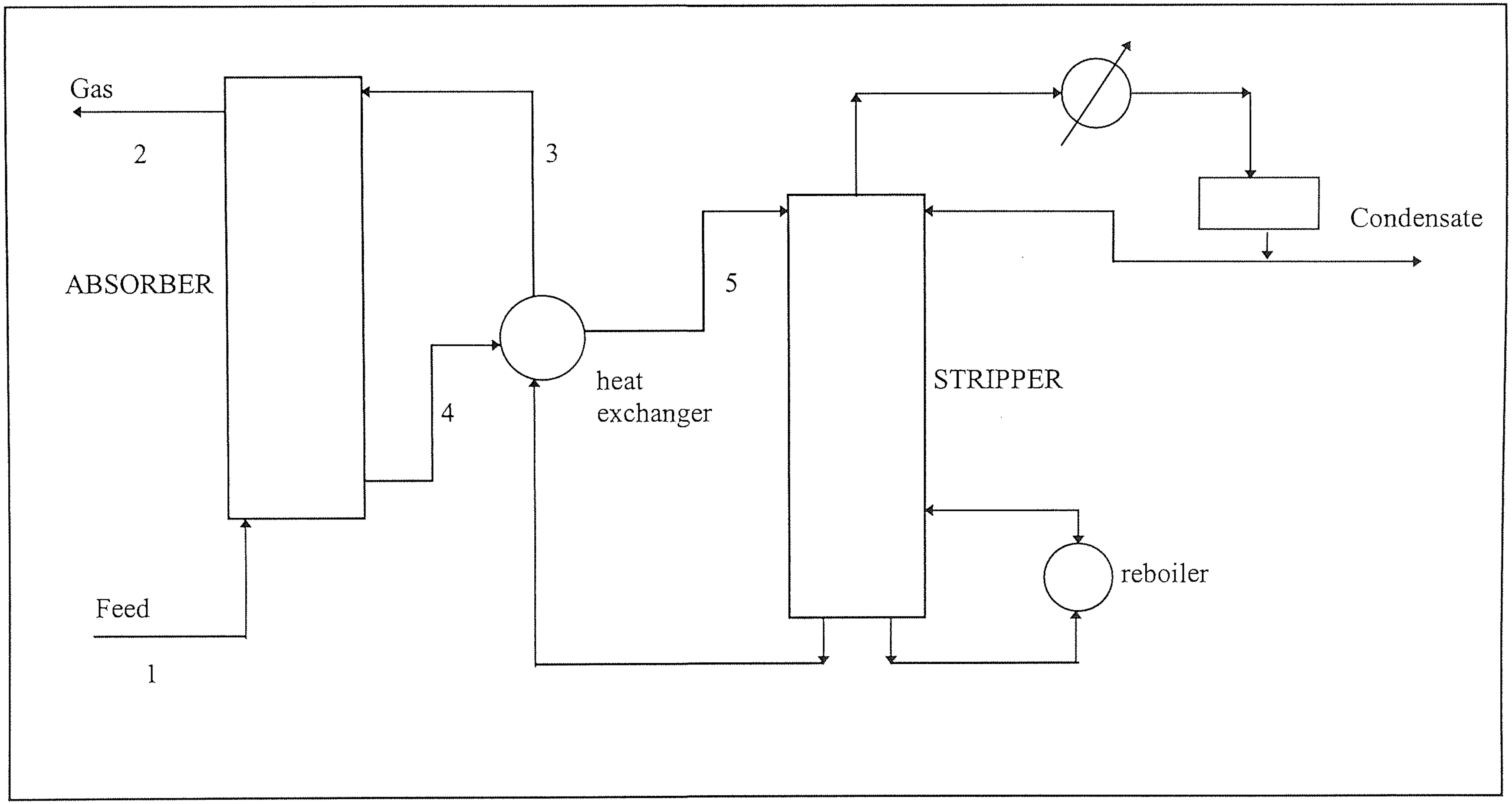


Figure 22: Absorption process flow scheme.

from there through the rich oil/lean oil heat exchanger to the top of the stripper, where the absorber components are removed by heating and/or stripping with steam. The lean oil from the bottom of the stripper is cooled in the rich oil/lean oil exchanger. The lean oil then passes through the gas chiller, where further cooling may take place (if required), before being recycled to the absorber and the absorbed material is removed overhead from the stripper.

In simulating the absorber column, a light lean oil of 120 to 140 molecular weight is used. The lean oil, usually either nC_{13} or nC_{14}, is used because the column is operated at refrigeration temperatures. Note that the absorber operates at higher pressure and lower temperature, whereas the stripper operates at lower pressure, between 20 to 30 bar, and higher temperature.

The absorber is modelled using a rigorous distillation method within the BP GENESIS module DIS03. The column has no heater (reboiler) or coolers (condenser) and no sidestreams. The program calculates the internal temperature, liquid and vapour profiles and the product rates, heat contents and compositions.

The computational procedure generally followed is:

1. Assume top tray temperature.
2. Make dew-point calculation on the estimated exit gas composition.
3. Find composition of liquid leaving top tray using K-values at the top tray pressure and temperature.
4. Find composition of vapour leaving plate 2 from liquid leaving plate 1.
5. Run enthalpy balance around top plate using assumed vapour temperature from plate 2. If balance checks, continue a similar calculation for plate 2. If not, repeat steps 1 - 5.
6. Continue the calculation down the column until reaching the bottom plate. Check the overall heat balance. If it checks, solution is complete. If not the procedure is repeated until the overall heat balance checks.

The stripper is simulated using the distillation module DIS01, which uses a short-cut calculation method, using the Fenske, Underwood and Gilliland equations. The method also assumes constant molar overflow.

For a simple column rating calculation, component purity and recovery specifications are used. The duties of the condenser and reboiler are thus based on the overall enthalpy balance, which is based on the overhead and bottoms product specifications.

4.12 GAS COMPRESSION

Gas compression is required if the produced gas pressure is below the specified gas pressure. The duty of a centrifugal gas compressor for natural gas is estimated, using the BP GENESIS flow sheeting package. The same procedure that is used to model the turbo-expander, above, is used here for the compressor. In the case of the turbo-expander the specification is a drop in pressure, whereas with a compressor, it is a rise in pressure.

Note that, for gas compression a driver size of about 25% above the calculated total power is specified to allow for mechanical and thermodynamic inefficiencies.

4.13 CONDENSATE STABILISATION

The condensate stabilisation facilities process raw condensate by removing the light front end components from the feed to meet the specified Reid Vapour Pressure (RVP) of no greater than 14.5 psi.

The process is thus modelled by assuming that it essentially removes all the butane and lighter components from the feed. The stabilised condensate from the bottoms can then be stored at atmospheric temperature and pressure. The off-gas from the condensate stabilisation facilities is assumed to be used as a fuel, if the feed to the stabilisation system remains below 100 kmol per hour. However, if the raw condensate feed flowrate rises above 100 kmol per hour, a compressor will be required for the stabilisation system to recompress all the off-gas, up to the main stream pressure.

The feed to the stabiliser is usually a combined raw condensate from the plant inlet slug catcher and from the gas dew-pointing units. The column typically has 10-20 trays and is operated at approximately 10 bar.

4.14 REFLUXING CONDENSER

The model used for simulating the refluxing condenser is a modified BP model and, as such, the exact details of how the process is modelled cannot be released. The model is based on the characteristics of the refluxing exchanger, in that it is directly analogous to the multistage rectification column. For further details see chapter 2 (literature review).

The flow diagram of the reflux exchanger process used is shown in figure 23, page 75.

In the next chapter these models are used to correlate the cost data obtained from vendors.

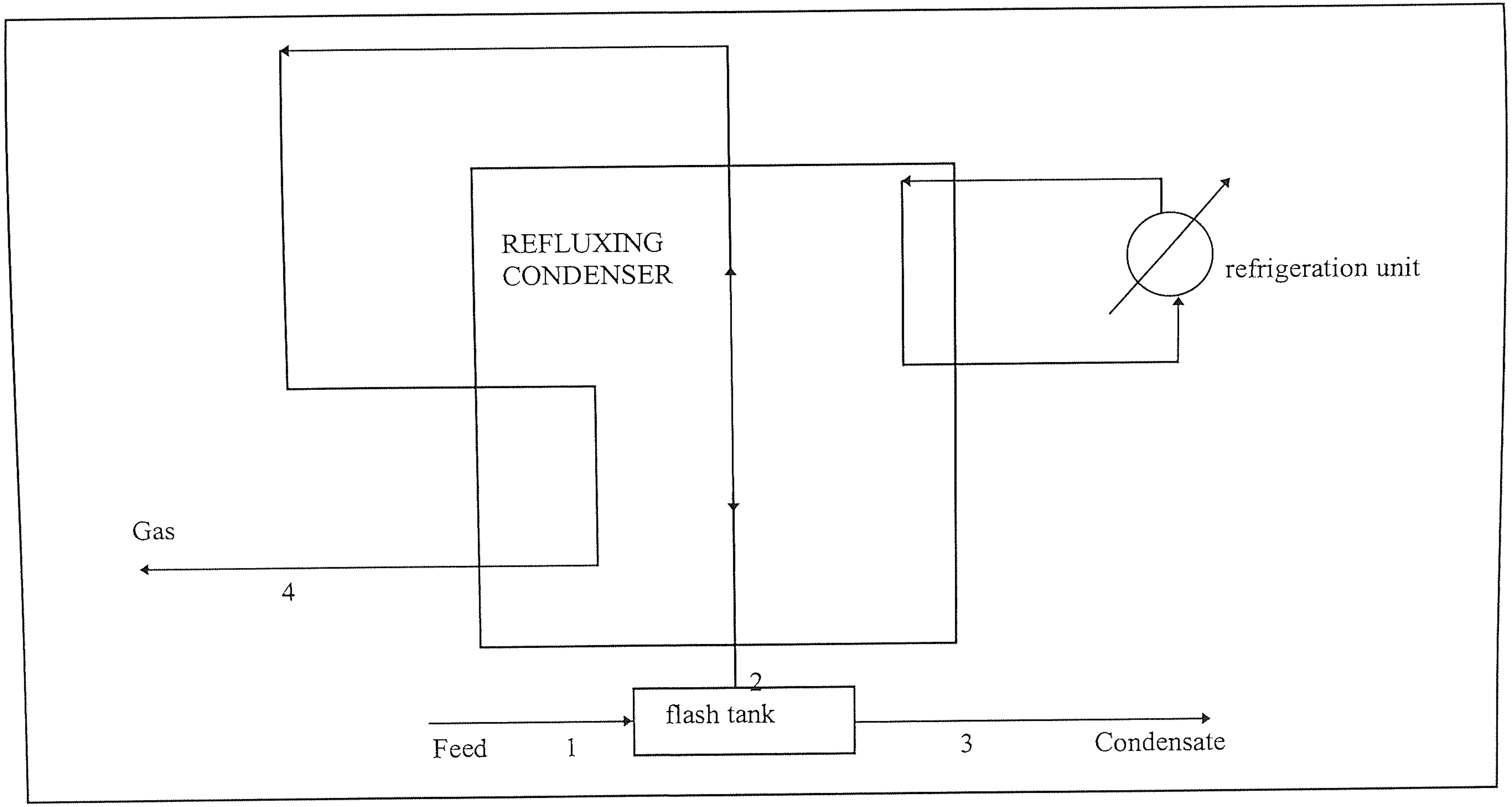


Figure 23: Reflux condenser process

4.15 INTERIM OBSERVATIONS

1. These models can from process specification, produce estimates of equipment size and power requirements.
2. The models presented here represent actual process method schemes.
3. The models reviewed in this chapter provide a suitable basis for correlation of the cost data obtained from the sources available.

CHAPTER 5

5.0 ECONOMIC EVALUATION

5.1 INTRODUCTION

Economic evaluation was carried out to determine the capital costs of gas processing plants. Using the capital expenditure (CAPEX), competing process technologies can be compared on an economic basis. All the processing technologies evaluated are commercially available. Variations of the individual technology will also be discussed.

Unless otherwise stated, the cost estimates cover the actual equipment only, thus excluding auxiliary equipment, such as installation, piping, labour or materials. This will make the costs independent of location, since the installation factors vary greatly depending on whether the plant is located offshore, or onshore. The costs are of order of magnitude or study estimate quality. An accuracy of no better than ± 20 percent should therefore be assumed.

The cost of any equipment can normally be related to a characteristic design variable. The chosen design variable is dependent on the nature of design equation. The design equations for most processing equipment may be fairly complex. To overcome this, the factors influencing the CAPEX of the process have to be understood.

Due to the nature of this project, the cost relationships have to be kept generic, taking into account different types of feed gases, lean gases and rich gases, variations in feed temperature and pressure. The simplified cost relationships may then be used at the conceptual design phase of a project to make quick preliminary estimates of the capital cost of equipment.

The cost information was obtained from number of sources. To make sure that it was commercially viable, all the enquiries were made under the auspices of BP Exploration

Facilities Engineering. The enquiries were kept very broad to acquire as much information as possible.

The economic evaluations are based on an inlet design pressure of 100 bar. Costs required for units designed to different inlet pressures can be determined by using the pressure adjustment factors given in appendix A1-5.

5.2 SOURCES OF INFORMATION

In most cases, the process equipment prices are based on actual project data obtained from BP Exploration and vendors. For Joule-Thomson, Turbo-expander, Refrigeration, Gas Compressor, Adsorption, condensate stabilisation and Absorption processes, the prices have been escalated from 1991 to the end of 1995 using the Chemical Engineering Plant Cost Index published in Chemical Engineering Journal. For the Membrane, Vortex tube, and Refluxing Exchanger processes, the cost were obtained at the end of 1995.

To honour commercial confidentiality the suppliers of each particular set of cost data have not been identified.

The processing units were to be designed and built to the current ASME standards.

5.3 SEPARATION PROCESSES

5.3.1 JOULE-THOMSON VALVE

The cost data obtained are based on actual projects and are in generic form with flowrate being the cost variable. The data obtained are mainly based on natural gas hydrocarbon dewpointing plants. They may therefore be only used for the purpose of estimating the costs of hydrocarbon dewpointing plants. The data include only the major components of the plant.

The emphasis is to cost only the major pieces of equipment. As seen in figure 17 on page 58, the main pieces of equipment are the gas/gas exchanger, the cold gas separator and the Joule-Thomson valve. The Joule-Thomson valve is not a major piece of equipment, but due to its importance in the process scheme, it is necessary to include its cost.

The cost data obtained are presented in Table 3(a) on page 106. The capital cost of the whole process scheme is given. The data represent a standard cost of a dew point control system with a feed gas inlet pressure of 100 bar. For a feed gas flow rate of 186.57 Mmscfd the corresponding costs of gas/gas exchanger, cold gas separator and Joule-Thomson valve are £671 445, £79 164, and £12 236. As expected the cost of the Joule-Thomson valve is minor compared to the other two major pieces of equipment. By assuming the cost ratios between the different pieces of equipment to be the same over the whole range of the cost data obtained, the corresponding cost for each piece of equipment could be calculated.

When sufficient 'free' pressure drop is available then the gas/gas exchanger may be omitted from the process scheme. Using the cost ratios, the cost of the process scheme without the gas/gas exchanger is calculated. Table 3(b) on page 106, suggests that savings of 88 % are possible for the same processing conditions.

5.3.2 TURBO-EXPANDER

The prices for the turbo-expanders have mainly been based on actual natural gas hydrocarbon dew point control projects. The data given include equipment only costs and they have been updated to the end of 1995. As was the case with the Joule-Thomson process the cost obtained is a function of the gas throughput. The estimates produced should be considered as being order-of-magnitude quality. The cost estimates are based on the turbo-expander plants meeting the hydrocarbon dew point with a typical pressure drop of between 15 to 35 bar. The costs for an NGL recovery would be higher due to higher pressure drop in the expander.

The capital cost of the turbo-expander is based on the flow diagram shown in figure 18 on page 61. The gas/gas exchanger, the expander knock-out drum, the expander-compressor and the cold gas separator are the main pieces of equipment that are taken into consideration.

The work produced by the feed gas in the expander can be used to recover partially the pressure of the product gas or generate power. The cost data assume that the compressor side of the turbo-expander is used to recompress the product gas after it leaves the gas/gas exchanger. If instead power is generated, the cost of the generator or the value of the power are not considered.

The cost data obtained are presented in Table 4(a) on page 107. They represent the total capital cost of the turbo-expander process scheme. All the major pieces of equipment are costed together. The data may only be used to estimate the capital cost of a turbo-expander plant required for hydrocarbon dew point control.

For a feed gas flow rate of 261.19 Mmscfd, the corresponding cost of the gas/gas exchanger is £892 830 and the rest of the equipment is £1 198 496. The total cost therefore is £2 091 326. The cost of the gas/gas exchanger is 42.69 % of the total cost. By assuming the percentage cost of the gas/gas exchanger to be the same over the whole range of the cost data obtained, the corresponding cost can be calculated as a funtion of feed flow rate.

Table 4(b) on page 107, represents the capital costs of the turbo-expander scheme without the gas/gas exchanger. The gas/gas exchanger may be omitted from the process scheme if sufficient 'free' pressure drop is available.

5.3.3 REFRIGERATION

The prices for the refrigeration systems have mainly been based on actual natural gas hydrocarbon dew point control projects. The data given include equipment only costs and they has been updated to end of 1995. The cost was obtained as a function of the gas throughput.

For dew point control, only moderately low temperatures are required. For this purpose either a propane-refrigerant unit or a freon-refrigerant unit can be used. The difference in cost between the two different units is not considered to be significant. It does not affect the overall cost of the refrigeration system.

Note that, the cost estimates obtained are based on plants processesing lean gas compositions. Therefore, to be able to use the cost estimates at the conceptual design phase of the project for any given lean or rich gas feed, the cost data obtained for the Refrigeration process have to be correlated and extended to make the refrigeration duty rather than the gas flow rate the cost function, since the flowrate does not take into account whether the gas feed is rich or lean.

However, the cost analysis recieved, states that rich gas plant would cost approximately 30 - 40 percent more for the same gas throughput, since the duty performed by the refrigeration unit would certainly be higher. Knowing this fact the cost data may be evaluated with a typical lean and a typical rich gas, so that they could be correlated with respect to duty required. Further details on how this was done for the Refrigeration system is given below in the correlation section. The estimates produced should be considered as being order-of-magnitude quality.

The capital cost of the refrigeration system includes the cost of the refrigeration unit. The process diagrams of the refrigeration unit and the refrigeration system as a whole are shown in figures 4 and 19 on page 64. The main pieces of equipment considered for costing are, the gas/gas exchanger, the cold gas separator, and, from the refrigerant unit, the expansion valve, the compressor, the gas chiller (evaporator) and the condenser.

The costs presented in Table 5(a) on page 108, are obtained for the refrigeration systems where the refrigerant compressor is either driven by a motor or by a gas turbine. A refrigeration system using a motor to drive the compressor is limited to a heat duty of roughly 1100 kW. This means approximately a lean gas throughput of up to 250 million of standard cubic feet of gas per day. The turbine-driven system does not have the same limitation, but its costs are between 30 to 68 per cent higher than the corresponding motor-driven system.

5.3.4 MEMBRANE PROCESSES

Vendors were asked to carry out order-of-magnitude cost evaluation for seven different cases. The bids were kept simple, so as to obtain generic information. In each case it is assumed that the permeate pressure (rich gas stream) is at 2 bara and the inlet temperature for cases 1 to 6 is 4°C ; for case 7 it is 0.49°C. The specified cases are:

Case	1	2	3	4	5	6	7
Flowrate [MMSCFD]	50	100	200	300	300	300	250
Inlet Pressure [barg]	71	71	80	71	71	71	90
Hydrocarbon Cricondentherm [°C]	-1.7	-1.7	-1.7	-1.7	-5	-8	-5

Feed gas composition 1 is used for cases 1 to 6 ; composition 2 for case 7, see appendix A1-1.

The analysis carried out by the vendor shows that for all cases, it is possible to meet the dew point with the feed gas flowing through a filter coalescer and then through the membrane skids. The estimated order of magnitude cost of these units, skid mounted and built to ASME standards is as shown in Table 6(a) on page 110.

The cost so obtained is only a part of the total capital cost. This can be seen by looking at the process flow diagram (Figure 20, page 66). The added units are necessary because the permeate pressure needs to be increased from 2 bar to 10 bar. Part of the stream also needs to be liquified before flowing to the condensate stabilization section of the plant.

As the emphasis is to cost only the major pieces of equipment, only the additional cost of the compressor will be included. The cost of the rest of the equipment is minor.

Therefore the main pieces of equipment are the filter coalescer, the membrane module and the compressor. The cost of the compressor will be calculated separately. Tables 6(b) and 6(c) on page 110, show the corresponding cost data obtained for the membrane module and the compressor. Since the duty for the compressor is likely to be below 1200 Horsepower, the cost represent a low to medium power centrifugal compressor.

5.3.5 ADSORPTION PROCESS

To date the application of Adsorption schemes has been limited. It is for this and proprietary reasons that little or no capital cost data are available for the hydrocarbon dew-point control process. But the silica gel and the more expensive molecular sieves have traditionally been used for gas dehydration. The difference between dehydration and hydrocarbon dew-point processes is that for the same size adsorber beds, the cycle times are shorter for hydrocarbon dew pointing. The process units for dehydration and dew-point control are exactly the same.

The other main difference between the dehydration and dew-pointing scheme is the choice of the adsorbent. For dew-point control silica gel is the preferred option, because it has a higher capacity for hydrocarbons than molecular sieves, used for dehydration. Therefore, by basing the capital cost on molecular sieve we are likely to be slightly over estimating the cost.

The cost data obtained and shown in Table 7 (on pages 111-114), are based on natural gas dehydration projects. The data provided are specific to certain cases and they need to be correlated to simplify them into a generic format which should consider the flowrate and the amount of adsorbent needed. The data given includes equipment-only costs and they have been updated to the end of 1995.

The proposed flow scheme consists of the regeneration knock out drum, three silica gel adsorber beds, the regeneration gas heater, the fines filter and the cooler/condenser. The size of the adsorber beds is affected by the gas flow rate and the amount of hydrocarbons to be removed.

5.3.6 VORTEX TUBE

To acquire the capital cost data for the Vortex process a number of different cases were sent to the vendor for evaluations. The inlet temperature for case 1 to 5 is 4°C ; for case 6 it is 62.2°C. The cases specified are:

Case	1	2	3	4	5	6
Flowrate [MMSCFD]	50	100	200	200	300	365
Inlet Pressure [barg]	71	71	60	80	71	35.5
Hydrocarbon Cricondentherm [°C]	-1.7	-1.7	-8.0	-1.7	-1.7	4.45

The feed gas for cases 1 to 5 was composition 1; for case 6 composition 3, see appendix A1-1.

The results of the bid enquiry are summarised in Table 8(a) on page 115.

The capital cost of the Vortex tube is based on the flow diagram shown in figure 21, on page 69. It is the only piece of equipment that is taken into consideration. The analysis carried out by the vendor shows that for cases 1 to 5, it is possible to meet the dew point for all the five cases using a single vortex unit. The estimated budget price of this unit, skid mounted and built to ASME standards with instrumentation would be £106, 670. Analysis of the evaluation carried out by the vendor suggests that for feed flow rates of upto 300 mmscfd, a single-unit Vortex tube is able to meet the required dew point specifications. If the feed flow rate is above this, than two process units are required. This was confirmed in further discussions with the vendor.

For case 6, in order to meet the dew point specification it was necessary to pre-cool the incoming feed gas to the Vortex unit using a gas/gas exchanger. As a result of low feed gas pressure and high volume, it would be necessary to employ vortex unit of the same type as used above in parallel to perform this duty. The estimated budget price of these two units, skid mounted and built to ASME standards with instrumentation would be £ 200, 000.

Table 8(b) on page 115, includes the cost of the gas/gas exchanger and the corresponding cost of the Vortex process scheme. Unlike some of the above processes, no cold gas separator is required to separate the condensed heavy hydrocarbons from the product gas.

5.3.7 ABSORPTION PROCESSES

The Absorption process is a constant pressure process, which maintains the gas at the inlet pressure level. Therefore when costing it is important to note that the absorber will be operating at high pressure whereas the stripper will be operating at low pressure and high temperature.

The capital cost of the Absorption process is based on the flow diagram shown in figure 22 on page 71. The absorber, the stripper, the lean oil/rich oil exchanger, the condenser, the reboiler, and the refrigerant unit are the main pieces of equipment taken into consideration.

The costs of the absorber, the stripper and the refrigerant unit were obtained separately. The cost of the absorber is based on a de-methanising column made of 9% nickel steel. With 9% nickel steel, the material factor is taken into account, since the maximum operating pressure of the absorber for the cases considered is 80 bar.

The cost of the stripper is based on a de-ethaniser with typically 30 trays. The de-ethaniser essentially performs the same function as the stripper, in that it removes an ethane-rich stream from an NGL stream. The operating pressures and temperatures are also similar. The cost includes the lean oil/rich oil exchanger, the condenser and the reboiler.

The cost data available for the de-methaniser column and the de-ethaniser were of very generic type, based on the gas flow rate. It was felt that if this were used for the Absorption process, we are more likely to over estimate the capital cost than to under estimate it.

Tables 9(a), 9(b) and 9(c) on pages 116-117 show the capital cost data.

5.3.8 REFLUXING CONDENSER PROCESS

The refluxing condenser consists of a brazed aluminum plate-fin heat exchanger in which both heat and mass transfer takes place to affect the condensation of the heavier hydrocarbon molecules. This process is relatively new to hydrocarbon dew point control. It is a constant pressure process and therefore offers advantages if no free pressure drop is available.

In order to obtain quality cost estimates a number of different vendors including IMI Marston and Costain were approached. However, due to the propriatory nature of the process none of the vendors approached were willing to supply any information about cost. The next step was to search the literature to see if there were any published data about the refluxing condenser process. Again no data were found. At this stage there were two options, either to leave out the refluxing condenser from the process evaluation or to estimate the capital cost.

Note that to overcome this problem the cost of the refluxing condenser process is based on the heat duty, since heat transfer is the dominant process. The capital cost of the refluxing condenser can therefore be estimated by combinimg the cost of the plate-fin heat exchanger and the refrigeration unit. To take into account the likelihood of under-estimating the equipment cost, the plate-fin heat exchanger would be over designed by 20 percent.

The capital cost data for the plate-fin heat exchanger were obtained by sending a number of different cases to the vendor for evaluations. The inlet temperature for cases 1 to 3 is 4°C ; for case 4 to 6 is 62.2°C. The inlet pressure for cases 1 to 3 is 90 bar ; for case 4 to 6 it is 89 bar. The respective feed flowrates are for cases 1 to 3 is 9899.23 kmol/h or 173252 kg/h ; for case 4 to 6 it is 9960.16 kmol/h or 218262 kg/h. The cases specified are:

Case	1	2	3
Heat Load [kW]	1413	2936	6000
Product gas Outlet Temperature [°C]	-6	-16	-34
Cooling gas Inlet Temperature [°C]	-12	-21	-40
Cooling gas Outlet Temperature [°C]	-2.55	-1.45	-2.94
Mass flowrate Cooling gas [kg/h]	172690	172152	170739
% Liquid [mass basis]	0.14	0.44	1.13

Case	4	5	6
Heat Load [kW]	5700	7350	9100
Product gas Outlet Temperature [°C]	23	13	3
Cooling gas Inlet Temperature [°C]	15	0	-20
Cooling gas Outlet Temperature [°C]	57	56	55

Mass flowrate Cooling gas [kg/h]	211813	204808	188981
% Liquid [mass basis]	1.85	3.367	5.504

The feed gas for cases 1 to 3 was composition 1; for cases 4 to 6 was composition 3, see appendix A1-1.

The results of the bid enquiry are summarised in Table 10(a) on page 118. The specifications for the heat exchangers are as follows : design pressure , 100 barg ; material type, 316L stainless steel.

The material requirement in our case is for brazed aluminium rather than stainless steel. The vendor was made aware of this and suggested that for brazed aluminium the cost would be half that of stainless steel.

For dew-pointing, only moderately low temperatures are required. The cooling duty is shared between the cooled product gas and the refrigerant unit. The refrigerant unit for this purpose can be either a propane-refrigerant unit or a freon-refrigerant unit.

The prices for the refrigeration units obtained have mainly been based on the actual projects. The data given include equipment-only costs and they have been updated to the end of 1995, Table 10(b) on page 118.

The process diagram is shown in figure 23 on page 75.

5.3.9 GAS COMPRESSOR

The cost data obtained comprises equipment only costs for centrifugal gas comprssors for natural gas service and it has been updated to end of 1995. The cost is obtained as a function of the driver power, (KW). The cost data is based on the compressor being driven by the gas turbine driver. The gas turbine driver is more economical since the fuel gas is readily available on site. The cost estimates produced should be considered as being order-of-magnitude quality.

Note, vendor data suggest that fuel gas usage at 25% gas turbine efficiency can be estimated by assuming 0.34 NM^3 of gas are needed per kW of power.

The cost data obtained are presented in Table 11 on page 119.

5.3.10 CONDENSATE STABILISATION

The prices for the condensate stabilisation systems have mainly been based on actual natural gas hydrocarbon dew point control projects. The data given include equipment only costs and they have been updated to the end of 1995. The cost was obtained as a function of the total raw condensate produced.

To simplify the calculation it has been assumed that if the raw condensate flow rate is below 100 kmol per hour than the off gas produced from the stabilisation process can pass directly to the plant fuel gas sysytem. However, if the raw condensate flow rate is above 100 kmol/h than the produced off-gas will be compressed and put back into the main product gas stream.

5.4 CAPITAL COST CORRELATION

In the previous section, the capital cost data have been collected for the alternative gas processing plant schemes. Some of the data are already in generic format, the remainder need to be simplified and put into a generic format, so that they could be used to make a preliminary estimate of the equipment cost at the conceptual design stage.

For estimates of this quality it is customary to present the costs as a function of one or two readily obtained design or sizing variables. The major variables which determine the cost of the system must therefore be defined.

Parameters which affect the capital cost are:

1. Gas flow rate: this affects the size of the main equipment items, such as adsorber and stripper columns and the membrane flow area.

2. The quantity of the heavy hydrocarbons to be removed: this particularly affect the size of, for example, the heat exchangers.

3. Pressure: higher pressure either increases the wall thickness of the equipment or needs a more expensive material.

To obtain the capital cost data a number of different cases were sent to the vendors for evaluation. Upon receiving the cost evaluations, exactly the same cases were modelled using the BP GENESIS flowsheeting package. The respective process models are presented in chapter 4. From the results of modelling, the cost data may then be correlated with the design or a sizing parameter, or both.

Most of the data can be presented either as power-functions or linear equations, with the regression coefficients, r, included. However, the cost equations are valid for only certain ranges of a given sizing variable. Any extrapolation beyond the specified range can introduce significant error into the cost estimates.

5.4.1 JOULE-THOMSON VALVE

The cost data for the Joule-Thomson valve are based on the plant meeting the hydrocarbon dew point duty. From Table 3 on page 106, it is clear that the gas/gas exchanger makes up the major part of the cost. The major factor which affects the size of the gas/gas exchanger is the gas throughput. Increasing the gas flow rate will also affect the amount of condensate and therefore the size of the cold gas separator. The cost data obtained reflect this. It is also important to note that the cost of processing decreases with increasing gas throughput.

The end of 1995 prices for the Joule-Thomson processes are represented by the following power-function equations. The accuracy of the fits is greater than 99.9%.

$$\text{Cost of the plant with the gas/gas exchanger (£1000)} = 20.76\, F^{0.692}\ PF$$

$$\text{Cost of the plant without the gas/gas exchanger (£1000)} = 2.487F^{0.692}\ PF$$

[F = gas flow rate/mmscfd]
[PF = pressure factor]

These equations are suitable for gas feed flowrate of 111 - 600 mmscfd.

5.4.2 TURBO-EXPANDER PROCESS

The turbo-expander process can be used with the gas/gas exchanger or without one. The cost savings of over 42% can be made if the gas/gas exchanger is not included. For the dew point control plant the major factor affecting the equipment cost is the gas flow rate. The feed gas composition is not considered to have much effect on the cost. Table 4 on page 107 shows the cost of processing decreases with increasing flow rate.

The following cost equations are obtained by curve-fitting the data. In each case the accuracy of the fit is greater than 99.9%.

Cost of the plant with the gas/gas exchanger (£1000) $= 45.59\ F^{0.687}\ PF$

Cost of the plant without the gas/gas exchanger (£1000) $= 26.127\ F^{0.687}\ PF$

These equations are suitable for gas feed flowrate of 111 - 600 mmscfd.

5.4.3 REFRIGERATION PROCESS

Table 5(a) on page 108 shows the equipment cost of the refrigeration scheme. The cost is based on the plant meeting the lean gas hydrocarbon dew point. From the bid evaluations, we are told that the equipment cost is 30 - 40 percent higher for a typical rich gas compared to the lean gas. This is because the quantity of the heavy hydrocarbons to be condensed is higher for a rich gas for the same dew point specifications.

The cost data can be simplified further and the 30 - 40 percent increase in the cost for the typical rich gas taken into account by making the equipment cost a function of feed gas composition.

To meet the same hydrocarbon dew point, for instance a cricondentherm of - 2^0C, for the lean and rich gas. The difference is the number of molecules to be condensed. The more molecules condensed, the larger the heat duty required of the refrigeration system. The larger the heat duty, the bigger the size of the refrigeration scheme and therefore the higher the cost of the equipment.

The refrigeration duty was calculated by using the refrigeration model in chapter 4, within BP GENESIS, for the entire range of gas flow rates shown in Table 5(a) on page 108, using composition 1. The resulting heat duty data are shown in Table 5(b). The heat duty can now be correlated with the capital cost.

By curve fitting, the following cost equations are obtained.

Cost of the plant with motor driver (£1000) $= 1005.6 + 0.9238\ H\ PF$

This equation is suitable for heat duty requirements of 386 - 1200 kW.

Cost of the plant with turbine driver (£1000) $= 820.46\ H^{0.172}\ PF$

This equation is suitable for heat duty requirements of 386 - 6100 kW.

[H = heat duty, kW]

To determine whether we can apply the same principles of using the heat duty to calculate the equipment cost for the rich gas feed, it is assumed that composition 2 represents a typical rich gas feed. The refrigeration scheme was simulated to calculate the duty with the rich gas to meet the hydrocarbon dew point of -2^0C. The results are shown in Table 5(c) on page 109.

The calculated cost for processing the rich feed gas is between 34.41 to 41.25 percent higher than the comparative cost for the lean feed gas. The calculated cost is either within or very close to the range specified by the contractors.

It is important to note that motor driver cost equation can only be used if the heat duty is below 1100 kW, or if the feed gas flow rate is below 250 mmscfd and it is a lean gas.

5.4.4 MEMBRANE PROCESS

Economic evaluation data for the membrane module were received from the vendor for a number of different cases. The aim is to correlate the data so that they could be used to evaluate the membranes for any given feed and product specifications. To simplify the problem of correlating the data, all the cost information shown was obtained from one supplier. This is because different suppliers have different types of membranes, with different selectivity and permeability properties.

The ideal situation would be to obtain the design procedure along with the membrane selectivity and gas permeability data, but due to its proprietary nature this is not possible. The approach is to correlate the data in Table 6(a) on page 110, by using the membrane model presented in chapter 4, to simulate the different cases. For each case exactly the same inlet and conditions as in those cases sent to the vendor were used.

From the simulations, a corresponding area of the membrane is calculated for each case. The results are shown in Table 6(b) on page 110.

The capital cost of the membrane module from the supplier can now be plotted against the area. The corresponding linear cost equation is obtained by curve fitting.

$$\text{Cost of the membrane module (\$1000)} = 237.37 + 18.436\,A$$

$$[A = \text{area of the membrane, m}^2]$$

This equation is suitable for membrane area of 30 - 150 m^2.

To calculate the total capital cost of the membrane plant, the cost of the compressor for the permeate stream has to be included. The following equation fits the cost data shown in Table 6(c) on page 110.

Cost of the compressor (\$1000) = $1.0783\ H^{0.794}$

[H = horse power, HP]

This equation is suitable for 40 - 4000 HP.

The total cost of the membrane plant (\$1000) = $237.37 + 18.436\ A + 1.0783\ H^{0.794}$

The equation can be used to estimate the capital cost of the membrane process plant for any given feed or product gas specifications.

Note that the estimated cost is valid only if the membrane model is used to calculate the membrane area.

5.4.5 ADSORPTION PROCESS

The costs of the main pieces of equipment for the adsorption process are presented in Table 7(a) on page 111. The feed gas flow rate and the amount of hydrocarbons to be removed are the factors which affect the equipment size and therefore the cost. The cost of processing a rich gas to the same specifications as a lean gas is much higher.

The data have to correlated in such a way so that they can be used to estimate, order-of-magnitude cost of the adsorption process scheme for any given feed or product gas specifications.

There are four sets of data, each set with a different rate of hydrocarbon recovery (condensate). The graph of cost *versus* flow rate were plotted and the equation of the lines obtained by curve fitting.

For case 1,

$$\text{Cost (£1000)} = 139.8 + 2.305\ F \qquad (1)$$

For case 2,

$$\text{Cost (£1000)} = 144.33 + 3.763\ F \qquad (2)$$

For case 3,

$$\text{Cost (£1000)} = 195.44 + 4.254\ F \qquad (3)$$

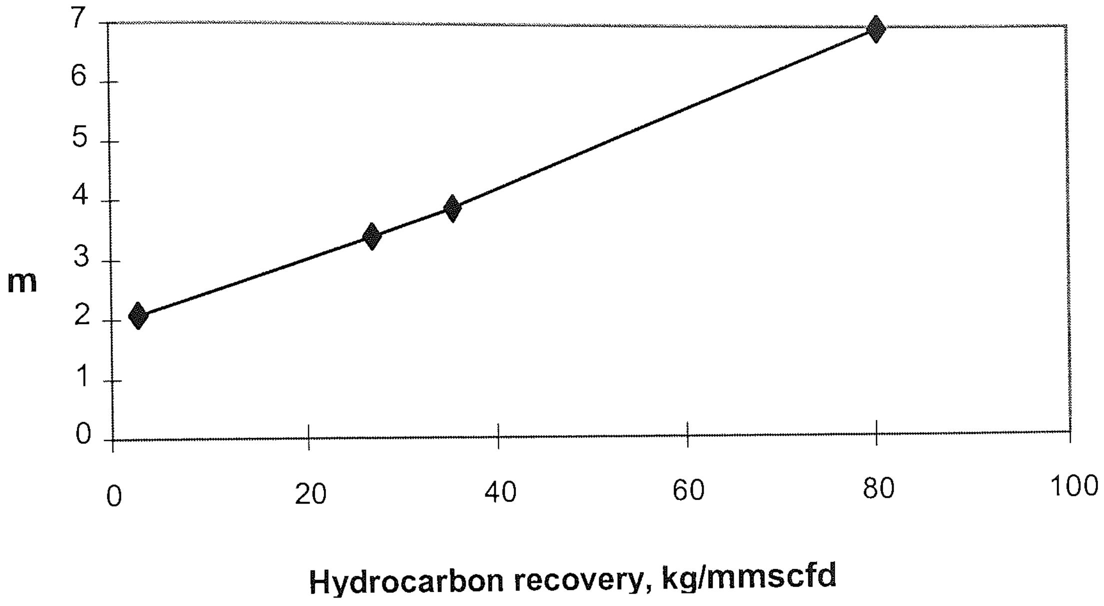
For case 4,

$$\text{Cost (£1000)} = 401.67 + 7.691\ F \qquad (4)$$

The theoretical point of origin of the data lines is determined by adjusting the axis. The location of the point is (-27.985 , 70).

The next stage is to determine the common/average gradient, which can used to represent all the hydrocarbon recovery rates. This is done by plotting the gradient (m) values, calculated in equations 1 - 4, against the corresponding hydrocarbon recovery rates as shown in figure 24, below. The following equation of the line was obtained by curve fitting.

Figure 24 : Determining the average gradient.



$$m = 1.96 + 0.07\,R \qquad (5)$$

[R = rate of hydrocarbon recovery, kg/mmscfd]

Equation of the straight line is

$$Y_c = mX_f + C \qquad (6)$$

[where Y_c = cost and X_f = gas flow rate (F)]

and if equation 5 is substituted into equation 6,

$$\text{Cost } (£1000) = \{1.96 + 0.07\,R\}\,F + C$$

taking into account the point of origin, therefore

$$\text{Cost (£1000)} = \{1.96 + 0.07\,R\}\,F + 70$$

but need to adjust for the axis

$$\text{Cost of the Adsorption process (£1000)} = [\{1.96 + 0.07\,R\}\,\{F + 27.985\} + 70]PF \quad \text{-------- (7)}$$

This can equation can only be used if the condensate recovery rate is between 2.6 - 80 kg/mmscfd and the corresponding feed gas flowrate is between 150 - 523 mmscfd.

To evaluate the accuracy of the above cost equation the equipment cost of the adsorption process was calculated using the equation for all the cases. The calculated values were then compared to the actual bid values, in Table 7(b) on page 113. For cases 1 to 4, the equation over-estimates the cost between 0.93 to 16 percent.

Considering that the equation will be used at the conceptual design phase of the project to estimate order of magnitude cost, the degree of error is acceptable.

The rate of hydrocarbon recovery is based on the model presented in chapter 4.

5.4.6 VORTEX TUBE

From the economic evaluation, we can conclude that irrespective of the dew point specification, the pressure drop, the cost of the vortex process remains the same. Only if the feed gas flow rate increases above 300 mmscfd, the cost is roughly doubled, because two units are required.

Table 8(a) on page 115 shows the cost obtained for the process scheme. Table 8(b) shows the cost of the vortex tube and the gas/gas exchanger. The cost of the gas/gas exchanger is taken from the turbo-expander process economic evaluations. The cost of the gas/gas exchanger compared to the vortex tube is very high and it varies with the feed flow rate.

The prices of the gas/gas exchanger and the vortex tube are combined to give the total process cost. The total cost of the process is represented by the following equation. The accuracy of the fit is 99.63%.

$$\text{Cost (£1000)} = 25.831\ F^{0.66}$$

5.4.7 ABSORPTION PROCESS

Unlike some of the other process schemes considered above the costs of the major pieces of equipment were calculated separately and then combined together. The data in Table 9(a) on page 116, include the cost of the absorber column and the pump. The cost is based on the feed gas flow rate, with the maximum flow rate of 466 mmscf/d. The vendor prices for the absorber are represented by the following linear cost equation:

$$\text{Cost of the absorber (£1000)} = \{265.65 + 1.5808\ F\}\ PF \qquad (8)$$

In Table 9(b), the cost of the stripper, the condenser, the reflux pumps, the reboiler and the lean/rich oil exchanger are included. The cost is based on the liquid flowrate of 1000 - 5000 tonne/d. The cost equation is:

$$\text{Cost of the stripper (£1000)} = 3.0651\ L^{0.666} \qquad (9)$$

[L = liquid flow rate, tonnes/d]

The cost of the refrigeration system used for cooling the condenser also need to be included in the cost equation. The cost equation for the refrigeration system is:

$$\text{Cost of the refrigerant unit } (£1000) = 100.65\ D^{0.3606} \qquad (10)$$

[D = duty, kW]

The suppliers suggest that for generic purposes a quick estimate of the required duty can be made by assuming 5.4 kW required per tonne/day of the top product of the stripper.

Equation 8, 9, and 10 are combined to give the total capital cost of the absorption process.

$$\text{Cost } (£1000) = \{265.65 + 1.5808\ F\}\ PF + 3.0651\ L^{0.666} + 100.65\ D^{0.3606}$$

The method of estimating the absorption cost is satisfactory for the purpose of determining the order-of-magnitude cost.

5.4.8 REFLUXING CONDENSER PROCESS

The capital cost of the refluxing condenser is estimated by combining the costs of the plate-fin heat exchanger and the refrigeration unit. The cost of the refrigeration units is the same as that represented by equation 10 for absorption, above.

However, the capital cost data obtained for the plate-fin heat exchanger need to be correlated before they could be used to calculate the price of refluxing condensers, when combined with the cost of refrigeration unit for any given feed and product specifications. The cost data presented in Table 10(a) on page 118, are correlated by using the model presented in chapter 4. The same inlet and outlet conditions were used when calculating the heat exchanger area. Table 10(b) shows that the difference between the calculated and actual area is less than 5 percent for all the six cases. Note, both the calculated and actual heat exchanger area have been over sized by 20%.

The vendor's prices for plate-fin heat exchangers are represented by the linear cost equation ($r^2 = 0.998$):

$$\text{Cost (1000)} = 15.413 + 0.3567\,A$$

[where A = area, m^2]

The following cost equation is obtained by combining the costs of plate-fin heat exchangers and the refrigeration unit to represent the estimated cost of the refluxing condenser:

$$\text{Cost (1000)} = 15.413 + 0.3567\,A + 100.65\,D^{0.3606}$$

This equation is suitable for cost calculation if the area of the exchanger is between 180 - 650 m^2 and refrigerant duty of 500 - 15000 kW.

5.4.9 GAS COMPRESSOR

For the gas compressor the cost equation is otained by curve-fitting the data in Table 11 on page 119.

It is:

$$\text{Cost of the compressor (£1000)} = 1208.3 + 125.75\,P + 9.98\,P^2$$

[where P = driver power, kW]

This equation is suitable if the required driver power is between 1 - 22 MW.

5.4.10 CONDENSATE STABILISATION

The cost data for condensate stabilisation are based on those of stabilisation systems meeting the RVP of 14.5 psi. The cost increases with increasing amount of condensate feed. The cost of the stabilisation process with a compressor is higher as a direct result of the compressor being included to boost the pressure of the produced off-gas. Note, the cost of the atmospheric tanks for the stabilised condensate storage is not included in the cost total.

The end of 1995 prices for the stabilisation processes are represented by the following linear equations.

Cost of the process without a compressor (£1000) = 124.3 + 3.235 C

Cost of the process with a compressor (£1000) = 466.23 + 9.327 C

[where C = condensate feed, m^3/h]

These equations are valid if the condensate flowrate is between 4 - 60 m^3.

In the next chapter, the selection methodology developed is explained.

5.5 INTERIM OBSERVATIONS

At this point it is possible to draw some interm conclusions as to the relative costs of the different processes examined, and to note some of the limitations of the analysis.

1. From the cost information supplied by the venders, cost correlations have been obtained in terms of cost parameters. The parameters used to describe the systems are:

Process	Parameters
Joule-Thomson	pressure and feed flowrate
Turbo-expander	pressure and feed flowrate
Refrigeration	pressure and heat duty
Membrane	pressure, area and power
Adsorption	pressure, feed flowrate and condensate recovery
Vortex tube	pressure and feed flowrate
Absorption	pressure, feed flowrate, heat duty and liquid flowrate

2. For an equivalent separation the cost of the Joule-Thomson valve is generally half that of both the turbo-expander and motor driven refrigeration. In contrast the cost of the vortex tube is roughly a fifth of the Joule-Thomson valve process.

3. The cost for refrigeration is higher if the feed is a rich gas and if the process is turbine driven. The turbine-driven system is used for high flowrates.

4. To calculate the total membrane cost, the cost of the permeate compressor has been added to the cost of the membrane module and the impact of different types of membranes from different suppliers has not been considered.

5. Like the refrigeration process the cost of the adsorption increases with increasing recovery of condensate.

6. The capital cost of the absorption process is higher than that of the adsorption process because the cost of the refrigerant unit is included in the absorption cost.

7. The processes were modelled as described in chapter 4 for correlating the vendors data.

8. The economic analysis carried out allows comparisons between different process systems to be made in terms of the process as a whole.

9. The correlations obtained can be used to estimate the capital costs of process technologies for other specifications.

10. Any extrapolation of the cost data beyond the specified range can introduce significant error into the cost estimates.

Table 3 : The capital cost (uninstalled) for Joule-Thomson valve process schemes.

a) : with the gas/gas exchanger

Flow rate [mmscfd]	Cost [£1000]	Cost/mmscfd [£1000]
111.94	542.02	4.84
186.57	762.84	4.09
261.19	977.04	3.74
335.82	1184.51	3.53
410.45	1338.31	3.26
485.07	1485.60	3.06
559.7	1632.78	2.92

b) : without the gas/gas exchanger

Flow rate [mmscfd]	Cost [£1000]	Cost/mmscfd [£1000]
111.94	64.94	0.58
186.57	91.40	0.49
261.19	117.06	0.45
335.82	141.92	0.42
410.45	160.34	0.39
485.07	177.99	0.37
559.7	195.62	0.35

Table 4 : The capital cost (uninstalled) for Turbo-expander process schemes.

a) : with the gas/gas exchanger

Flow rate [mmscfd]	Cost [£1000]	Cost/mmscfd [£1000]
111.94	1143.10	10.21
186.57	1649.68	8.84
223.88	1922.50	8.59
261.19	2091.33	8.01
335.82	2480.97	7.39
410.45	2857.81	6.96
485.07	3156.59	6.51
559.7	3468.28	6.20

b) : without the gas/gas exchanger

Flow rate [mmscfd]	Cost [£1000]	Cost/mmscfd [£1000]
111.94	655.09	5.85
186.57	945.40	5.07
223.88	1101.75	4.92
261.19	1198.50	4.59
335.82	1421.80	4.23
410.45	1637.75	3.99
485.07	1808.98	3.73
559.7	1987.60	3.55

Table 5 : The capital cost (uninstalled) data for Refrigeration process.

a) : The equipment cost for a lean gas feed.

Flow rate [mmscfd]	Gas Turbine Driven Cost [£1000]	Motor Driven Cost [£1000]
100	2286.54	1362.03
200	2576.04	1718.13
250	2676.84	1896.19
300	2762.12	2074.24
400	2902.22	-
500	3015.78	-
600	3111.85	-

b) : heat duty correlations

Flow rate [mmscfd]	Heat Duty kW	Gas Turbine Driven Cost [£1000]	Motor Driven Cost [£1000]
100	386	2286.54	1362.03
200	771	2576.04	1718.13
250	964	2676.84	1896.19
300	1157	2762.12	2074.24
400	1542	2902.22	-
500	1928	3015.78	-
600	2313	3111.85	-

c) : the application of the heat-duty correlation to a rich gas feed for a refrigeration system using a gas-turbine driven compressor.

Flow rate [mmscfd]	Heat Duty kW	calculated Gas Turbine Driven Cost [£1000]	% increase in cost with respect to a lean gas as in (a)
100	1210	2783.59	34.41
200	2420	3136.26	34.42
300	3630	3362.93	38.71
400	4840	3533.62	41.25
500	6050	3671.96	39.70

Table 6 : The capital cost (uninstalled) for the Membrane process scheme.

a) the membrane module.

Case	Flow rate [mmscfd]	Cost [$1000]
1	100	900
2	200	1300
3	200	1400
4	300	1925
5	300	2475
6	250	2695
7	300	2950

b) correlations with the respective membrane area.

Case	Area m^2	Cost [$1000]
1	31	900
2	62	1300
3	66	1400
4	93	1925
5	122	2475
6	127	2695
7	149	2950

c) the cost of the permeate compressor.

Horsepower HP	Cost [$1000]
40	20
100	42
300	100
800	215
1500	360
3000	620
4000	775

Table 7 : The capital cost (uninstalled) of the Adsorption process.

a) : the equipment cost.

Case : 1

Hydrocarbon recovery rate : 2.68 kg/mmscfd

Flow rate [mmscfd]	Cost [£1000]
149.25	485.48
223.88	650.66
298.51	828.09
373.13	1005.53
447.76	1173.13
522.39	1340.74

Case : 2

Hydrocarbon recovery rate : 26.80 kg/mmscfd

Flow rate [mmscfd]	Cost [£1000]
149.25	709.84
223.88	985.87
298.51	1261.91
373.13	1547.76
447.76	1830.53
522.39	2111.42

Case : 3

Hydrocarbon recovery rate : 35.376 kg/mmscfd

Flow rate [mmscfd]	Cost [£1000]
149.25	838.03
223.88	1143.54
298.51	1458.99
373.13	1782.06
447.76	2101.15
522.39	2421.35

Case : 4

Hydrocarbon recovery rate : 80.04 kg/mmscfd

Flow rate [mmscfd]	Cost [£1000]
149.25	1547.76
223.88	2121.03
298.51	2702.24
373.13	3274.29
447.76	3845.45
522.39	4416.51

b) comparison of the actual and calculated capital cost data.

Case : 1

Flow rate [mmscfd]	calculated Cost [£1000]	% Error between the calculated and bid data presented in part (a)
149.25	450.64	2.49
223.88	610.91	3.67
298.51	771.18	2.82
373.13	931.44	2.28
447.76	1091.71	2.75
522.39	1251.98	3.10

Case : 2

Flow rate [mmscfd]	calculated Cost [£1000]	% Error between the calculated and bid data presented in part (a)
149.25	749.89	16.64
223.88	1036.16	16.04
298.51	1322.43	15.71
373.13	1608.69	14.76
447.76	1894.96	14.30
522.39	2181.23	14.06

Case : 3

Flow rate [mmscfd]	calculated Cost [£1000]	% Error between the calculated and bid data presented in part (a)
149.25	856.29	12.82
223.88	1187.36	14.64
298.51	1518.43	14.91
373.13	1849.49	14.59
447.76	2180.56	14.59
522.39	2511.63	14.53

Case : 4

Flow rate [mmscfd]	calculated Cost [£1000]	% Error between the calculated and bid data presented in part (a)
149.25	1414.89	0.93
223.88	1981.16	3.13
298.51	2547.42	4.09
373.13	3113.69	5.00
447.76	3679.96	5.66
522.39	4246.23	6.16

Table 8 : The capital cost (uninstalled) of the Vortex tube process schemes.

a) : without the gas/gas exchanger

Flow rate [mmscfd]	Cost [£1000]
100	106.67
200	106.67
300	106.67
365	200.00

b) : with the gas/gas exchanger

Flow rate [mmscfd]	Cost of the gas/gas exchanger [£1000]	Total cost [£1000]
111.94	488.01	594.68
186.57	704.28	810.95
261.19	892.83	999.50
335.82	1059.18	1259.18
410.45	1220.06	1420.06
485.07	1347.61	1547.61
559.7	1480.68	1680.68

Table 9 : The capital cost (uninstalled) data for the Absorption process.

a) : cost of the absorber.

Flow rate [mmscfd]	Cost [£1000]
37.31	293.91
111.94	436.10
149.25	516.68
223.88	635.19
298.51	763.16
335.82	815.31
373.13	862.72
410.45	905.37
466.42	967.00

b) : cost of the stripper.

Liquid flow rate [tonnes/d]	Cost [£1000]
1000	306.06
2000	476.52
3000	635.36
4000	759.33
5000	894.93

c) : cost of the refrigerant system.

Heat Duty kW	Cost [£1000]
500	1104.13
1000	1246.08
2500	1703.51
4000	2018.97
5000	2239.80
7000	2618.36
10000	3107.33
11000	3233.51
12500	3407.02
15000	3643.62

Table 10 : The capital cost (uninstalled) data for the Refluxing condenser process.

a) : the cost of the plate-fin heat exchanger.

Case	Heat load [kW]	Area [m²]	Weight [kg]	Cost [£1000]	Cost / 2 [£1000]
1.1	1413	183	5000	160	80
1.2	2936	452	12300	350	175
1.3	6000	732	19900	550	275
2.1	5700	880	24000	650	325
2.2	7350	788	21500	600	300
2.3	9100	645	17600	500	250

b) : comparison of actual and calculated plate-fin heat exchanger area.

Case	Calculated area [m²]	Actual area [m²]	% Error between the calulated and actual area
1.1	179.43	183	-1.95
1.2	451.69	452	-0.07
1.3	750.00	732	2.46
2.1	890.63	880	1.21
2.2	807.69	788	2.50
2.3	622.22	645	-3.53

Table 11 : The capital cost (uninstalled) data for the gas compressor.

Power [MW]	Cost [£1000]
1	1296.24
2	1488.36
4	1872.60
5	2136.48
7	2639.97
9	3197.55
11	3768.38
12	4117.29
13	4512.56
15	5244.60
16	5728.21
17	6395.10
19	7130.49
20	7683.61
21	8280.95
22	8833.01

Table 12 : The capital cost (uninstalled) data for a condensate stabilisation system.

Condensate flow rate [m^3]	No compressor [£1000]	With compressor [£1000]
4	132.50	493.54
10	149.39	562.00
15	172.13	630.01
20	194.88	
25	211.11	682.02
30	227.45	740.43
35	240.37	785.92
40	253.29	841.12
45	269.52	893.02
50	285.75	948.22
55	298.78	980.69
60	315.01	1013.26

CHAPTER 6

6.0 A SHORT-CUT METHOD OF PROCESS EVALUATION

It should now be clear that selecting the best method for gas processing is a complex problem. At least ten alternative methods of processing are available, the composition, pressure and temperature of the feed gas varies widely and there are several different specifications for the product gas.

A further complication is that after an initial separation, further processing of the product streams is usually required. The gas product may need to be recompressed and the liquid product may need stabilisation to remove any high vapour pressure components.

Note that all product specifications, such as the dew point, calorific value and Wobbe number can be translated into a single cricondentherm temperature, which may be represented on the Phase Diagram. Further explanation on the cricondentherm and the product specifications is given in chapter 2.

The methodology takes into account the value of the products, the reprocessing costs and the capital costs of the separators. The methodology recognises that the value of the products is dependent on the composition of the products, which in-turn is dependent on the efficiency of the processing method. Therefore, how the gas and liquid are separated not only affects the total value of the products but it also affects the cost of liquid stabilisation and gas recompression.

The methodology described below chooses between the different processing options on the basis of comparative cost. This is defined as:

$$\text{Comparative Separation Cost} = [\,TPV_{\text{base case}} - TPV\,] + AR + AC$$

$$TPV = V_G + V_L + E$$

$$AR = C + S$$

where,

TPV	=	Total product value,
V_G	=	Value of the product gas,
V_L	=	Value of the liquid product,
AR	=	Amortised reprocessing cost,
C	=	Cost of gas compression,
S	=	Cost of stabilising the liquid,
AC	=	Amortised capital cost of separator, and
E	=	Value of the recovered energy.

This is further illustrated in figure 25, which presents a simplified flow sheet for any given separation process.

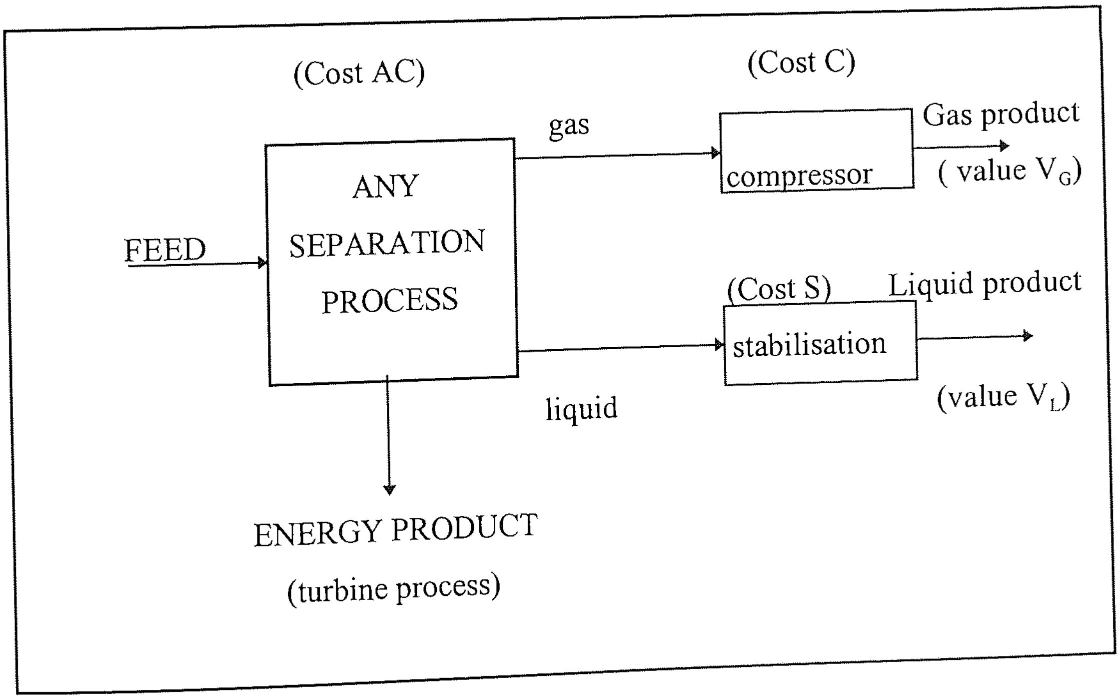


Figure 25 : Simplified generalised gas separation flow sheet.

In order to obtain the comparative separation cost of processing of using the alternate processing methods a base-case method is chosen. Any of the processes listed below may be used as a base-case. The total product value of any other processes being considered is then subtracted from the TPV of the base-case. Any loss (positive number) in the total product value with respect to the base case is taken as a penalty cost, whereas any gain with respect to the base case is taken as extra value (negative number). To this are added the amortised reprocessing cost and the amortised capital cost of the separator.

This method of comparative cost estimation is very simple and quick. Also, by taking into account the efficiency of each process and its effect on the values and costs means that the competing process are compared on equal footing. In addition the dominant effect of the total product value is eliminated by taking a base case method. As a result this methodology will clearly demonstrate the total cost of using one process over another.

Even though the cost data has been obtained under the auspices of BP Exploration, there will be errors within it. However it is important to remember, that this methodology is not a data base and hence, if a more accurate cost database is available then it can be used instead.

In developing the methodology for choosing the best process, note that the separation methods may be classified into Equilibrium process methods or Non-equilibrium process methods. The equilibrium methods are those which depend on changing the temperature and sometimes the pressure of the feed gas such that the mixture may be separated into a vapour phase and a liquid phase each in equilibrium with the other. These processes are thus limited to the two-phase region of the phase envelope. This classification is set out below.

Equilibrium process methods

1. Joule-Thomson valve
2. Vortex Tube
3. Turbo-expander
4. Refrigeration

Non-equilibrium process methods

5. Membrane
6. Absorption
7. Adsorption
8. Refluxing Condenser.

Note that all of the non-equilibrium processes and the Refrigeration process do not require any reduction in the feed gas pressure thus the need for recompression of the gas is removed. However the pressure-reducing equilibrium processes are still candidate processes because of their simplicity and either their low cost or the possibility of energy recovery. Thus the new method of process evaluation starts by considering the equilibrium processes.

6.1 CHOOSING BETWEEN EQUILIBRIUM PROCESSES

Any equilibrium process may be represented as in figure 26. The process may also be illustrated on the phase diagram, a Pressure - Temperature chart (figure 27), which includes a two phase region.

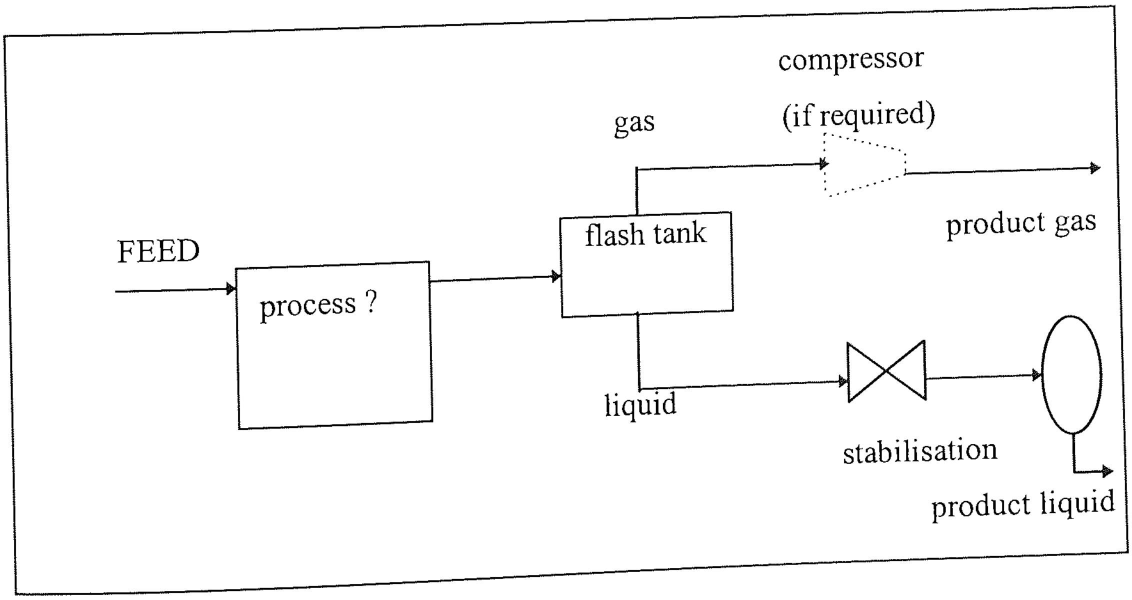


Figure 26 : Equilibrium separation process flow sheet.

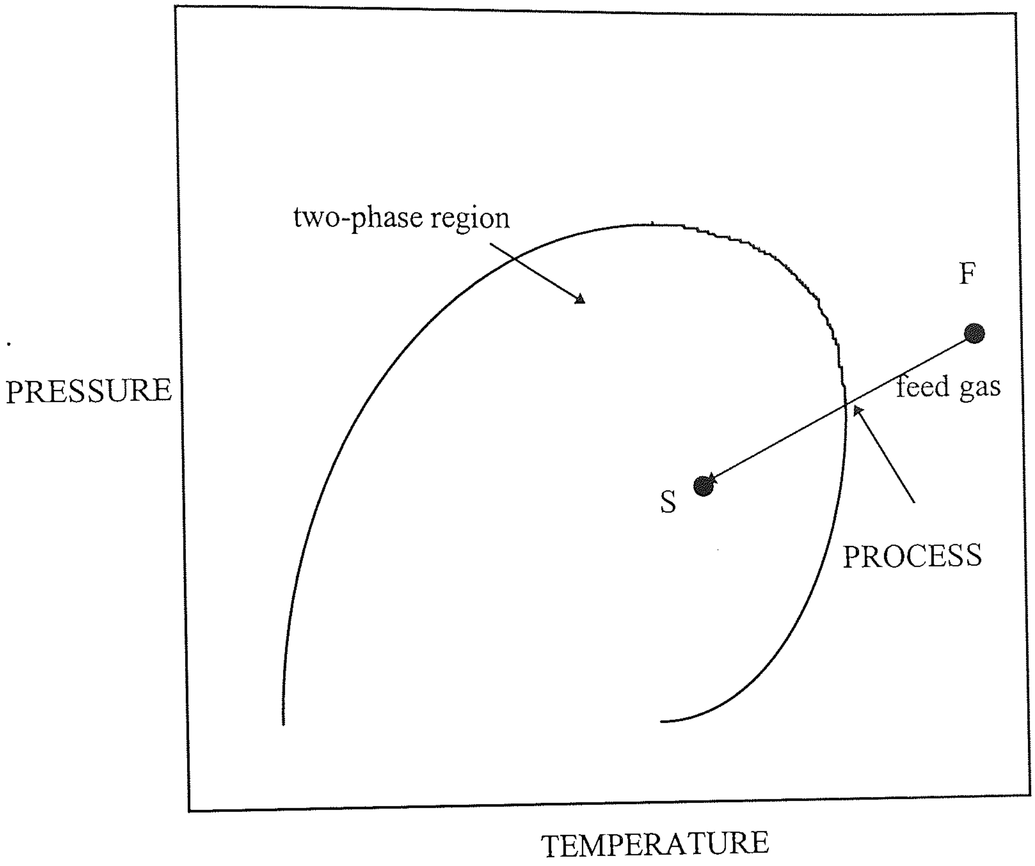


Figure 27 : Typical phase diagram.

Here the feed gas at conditions F is caused by the process to go to conditions S within the two phase region, and so is separated into:

a) a gas stream for recompression and

b) a liquid stream for stabilisation.

The combined composition of the gas and liquid at point S is the same as the feed gas composition at F. The composition and quantity of the vapour and the composition and quantity of the liquid at S are all fixed by the pressure and temperature at S and by the feed gas composition, and related one to another by the mass balance and the laws of thermodynamics.

Thus if the boundary limits of the two-phase region and the feed composition are known, then for any point, S, in the two-phase region the composition and quantities of the gas and liquid are known. It then follows that at any point, S, for a given quantity of feed and product specifications:

(i) the cost of recompressing the gas,

(ii) the cost of stabilising the liquid and,

(iii) the value of the final products may all be calculated.

A specimen calculation for a point(s) in the two-phase region is given in Appendix A1-6.

These costs and values may be calculated for many points, S, in the two phase region and presented as cost and value lines. These are illustrated in figures 28 to 42, on pages 130-144. The value of the gas and the liquid condensate may be combined to give the total product value (TPV); and the amortised reprocessing cost (AR) can be obtained from the combination of the liquid stabilisation and compression cost lines.

Regardless of whether the feed is a lean gas or a rich gas, the producer has to determine whether natural gas itself would be marketed as a gas or as liquid with or without the removal of any or all of the heavy components, or may form the feed stock to a chemical process. There are three main categories of product gas specifications, the pipeline, the sales gas and the maximum recovery of natural gas liquids, of which only the first two will be considered.

To illustrate the evaluation methodology 3 different feeds and product specifications are considered. These cases are provided by BP Exploration to cover the range of conditions met in practice and are given below.

Feed 1

Pressure : 35.50 bar
Temperature : 62 °C

Composition

Components	mol%
Nitrogen	0.742
Carbon Dioxide	5.241
Methane	76.067
nC2	10.373
nC3	4.227
iC4	0.935
nC4	1.264
C5+	1.152

Feed 2

Pressure : 80 bar

Temperature : 4 °C

Composition

Components	mol%
Nitrogen	1.757
Carbon Dioxide	0.400
Methane	92.004
nC2	3.987
nC3	0.861
iC4	0.151
nC4	0.201
C5+	0.641

Feed 3

Pressure : 118 bar
Temperature : 0.49 °C

Composition

Components	mol%
Nitrogen	0.815

Specification 1

Cricondentherm : 10 °C

Product gas pressure: 84 bar

Condensate specification

Reid Vapour Pressure: 14.5 psi

Specification 2

Cricondentherm : - 1.7 °C

Heating value

Calorific value: 975-1115 Btu/scf
Wobbe number: 1272-1351 Btu/scf

Product gas pressure: 70 bar

Condensate specification

Reid Vapour Pressure: 14.5 psi

Specification 3

Cricondentherm : - 22 °C

Heating value

Calorific value: 975-1115 Btu/scf
Wobbe number: 1272-1351 Btu/scf
Product gas pressure: 70 bar

Carbon Dioxide	2.360
Methane	81.750
nC2	8.903
nC3	3.529
iC4	0.556
nC4	1.053
C5+	1.019

Condensate specification

Reid Vapour Pressure: 14.5 psi

A complete list of the compositions is given in Appendix A1-1.

Figures 28-32, on pages 130-134 refer to feed 1 and transport pipeline specification. Here only partial processing is required to avoid condensation during transportation to the central processing facilities. The dew-point depression required is therefore small, since only a very small percentage of the heaviest components need to be removed.

Figures 33-37, on pages 135-139 refer to feed 2 and sales gas specification. This specification is far more stringent than the pipeline specification above. The required dew-point depression is greater and the produced gas must also be within the necessary heating value range specified by the customer.

Figures 38-42, on pages 140-144 refer to feed 3 and sales gas specification. The category of the specification is the same as above but due to the different feed gas conditions the duty is in fact far higher. Here the main restrictive factor is the specified heating value range.

In this work the construction of figures 28 to 42 was made, with some difficulty, by selecting a number of points in the two-phase region and calculating compositions, flow rates and costs for each point by the method illustrated in Appendix A1-6. Should this method of process selection be adopted, a computer program may be written and used to generate the various figures for the feed gas to be evaluated at the beginning of the conceptual analysis.

The advantage of this approach will now be explained first by a qualitative evaluation from inspection of the various figures and then by a quantitative evaluation of each of the processes which will be illustrated on the charts.

6.2 QUALITATIVE EVALUATION BY INSPECTION OF FIGURES

Inspection of figures 28 to 32 for case 1, figures 33 to 37 for case 2 and figures 38 to 42 for case 3 leads to the following deductions.

Case 1: Irrespective of the chosen process, the produced gas will require recompression. This is because the feed pressure at 35.50 bar is lower than the specified transport pipeline pressure of 84 bar. Therefore the processing cost is minimised if the small percentage of heavy hydrocarbons are recovered, without substantial pressure loss. Of the equilibrium methods only refrigeration meets this criteria.

Case 2: Here the feed pressure at 80 bar is close to the specified product pressure of 70 bar. Therefore there is not enough free pressure drop available. This means that the cost of recompression is significant for all points in the two phase region below 70 bar as shown in figure 35 on page 137. Thus methods which maintain the gas pressure (refrigeration or non-equilibrium methods) are favoured.

Case 3: Unlike case 1 and case 2, here free pressure drop is available. The feed pressure at 118 bar is higher than the specified product gas pressure of 70 bar. Therefore the cost of recompression is incurred only if the process operating conditions drop below the 70 bar level. Figure 42 on page 144, illustrates that for the equilibrium processes, to meet the specified cricondentherm temperature the process pressure must fall below 80 bar. This will bring the operating point into the two-phase region since the feed point is above it. Thus the simple pressure drop processes such as a turbo-expander or a vortex tube may have an advantage.

Figure 28, Case 1: Value of the gas

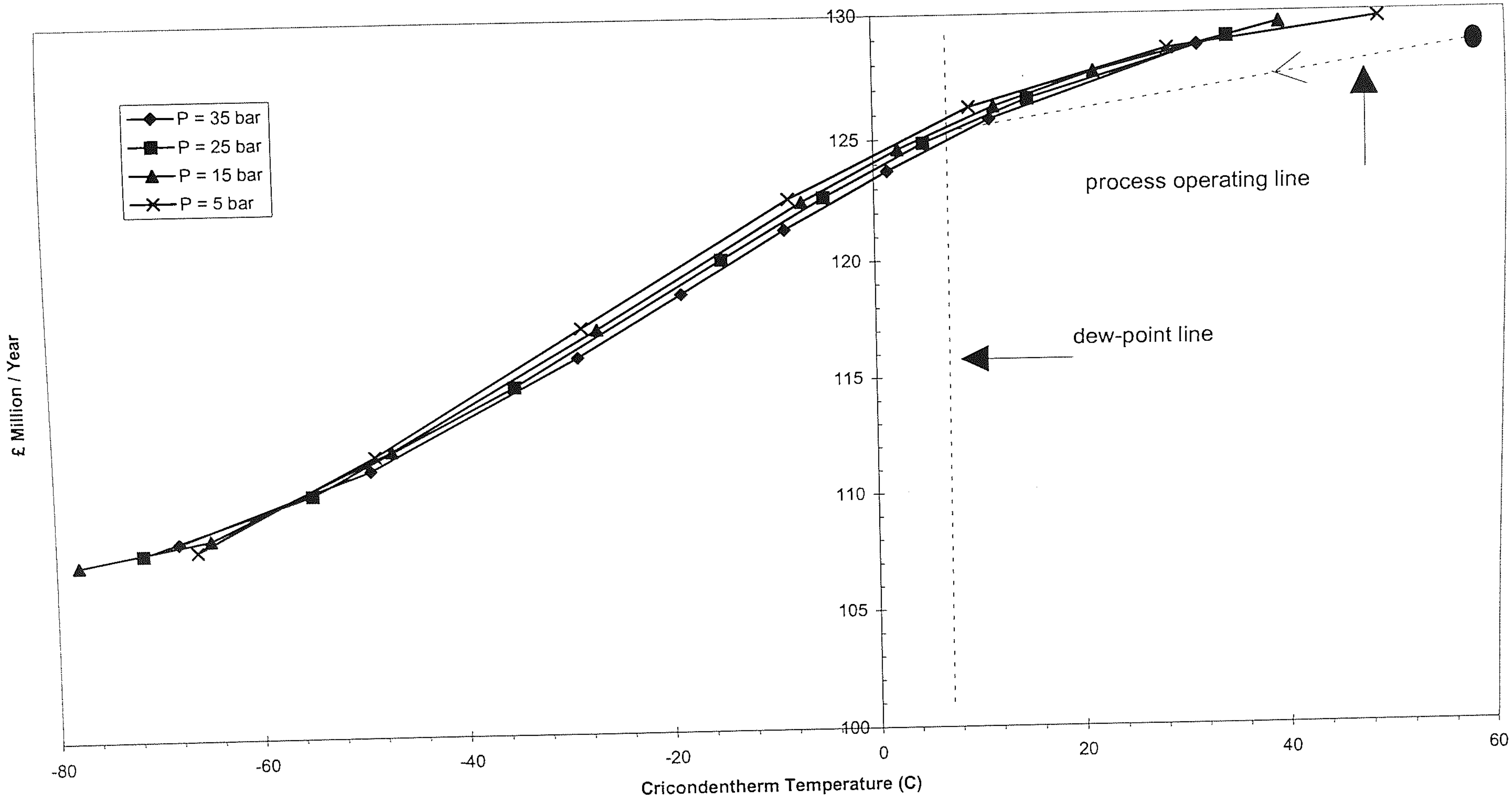


Figure 29, Case 1: Value of the liquid condensate

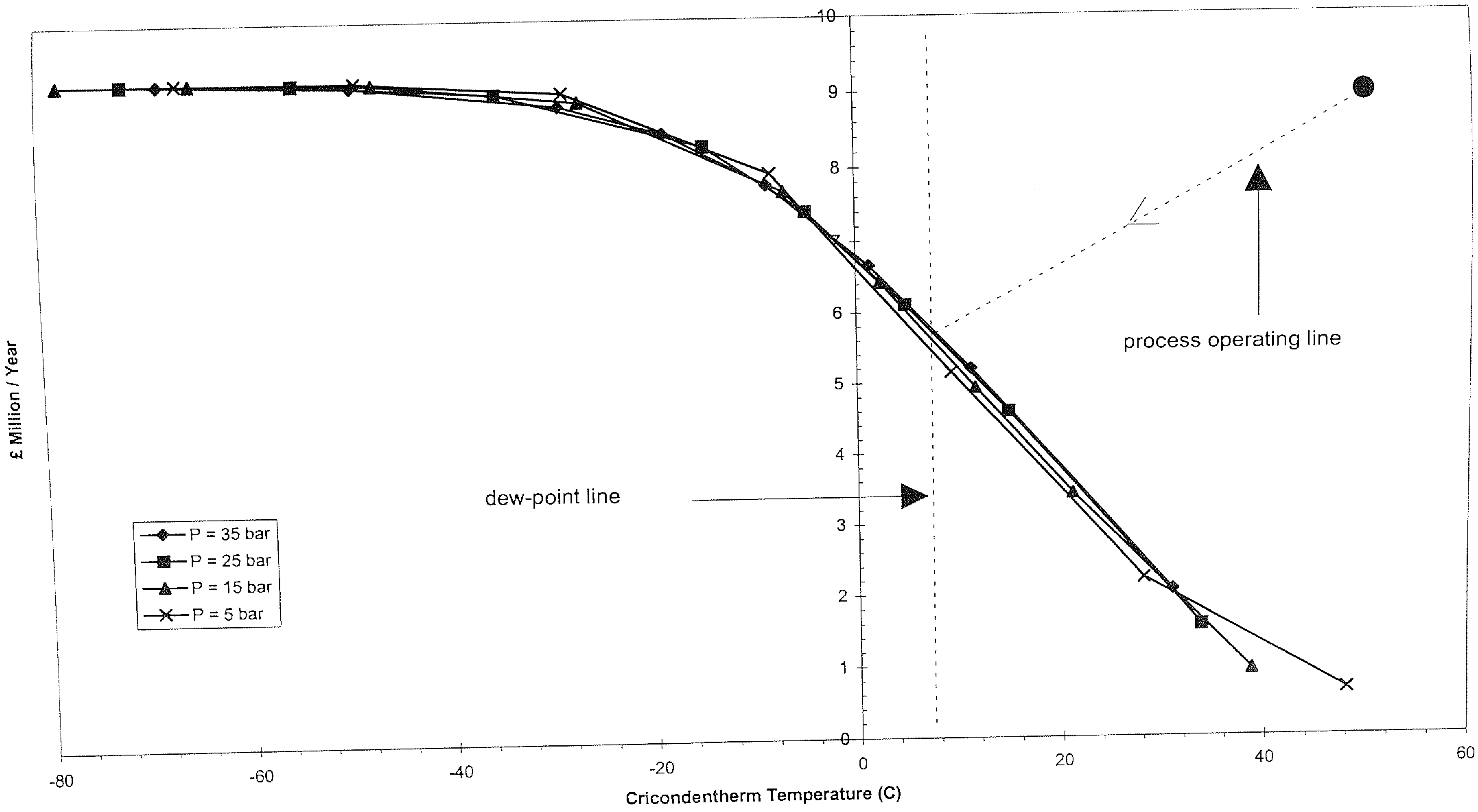


Figure 30, Case 1: Cost of liquid stabilisation

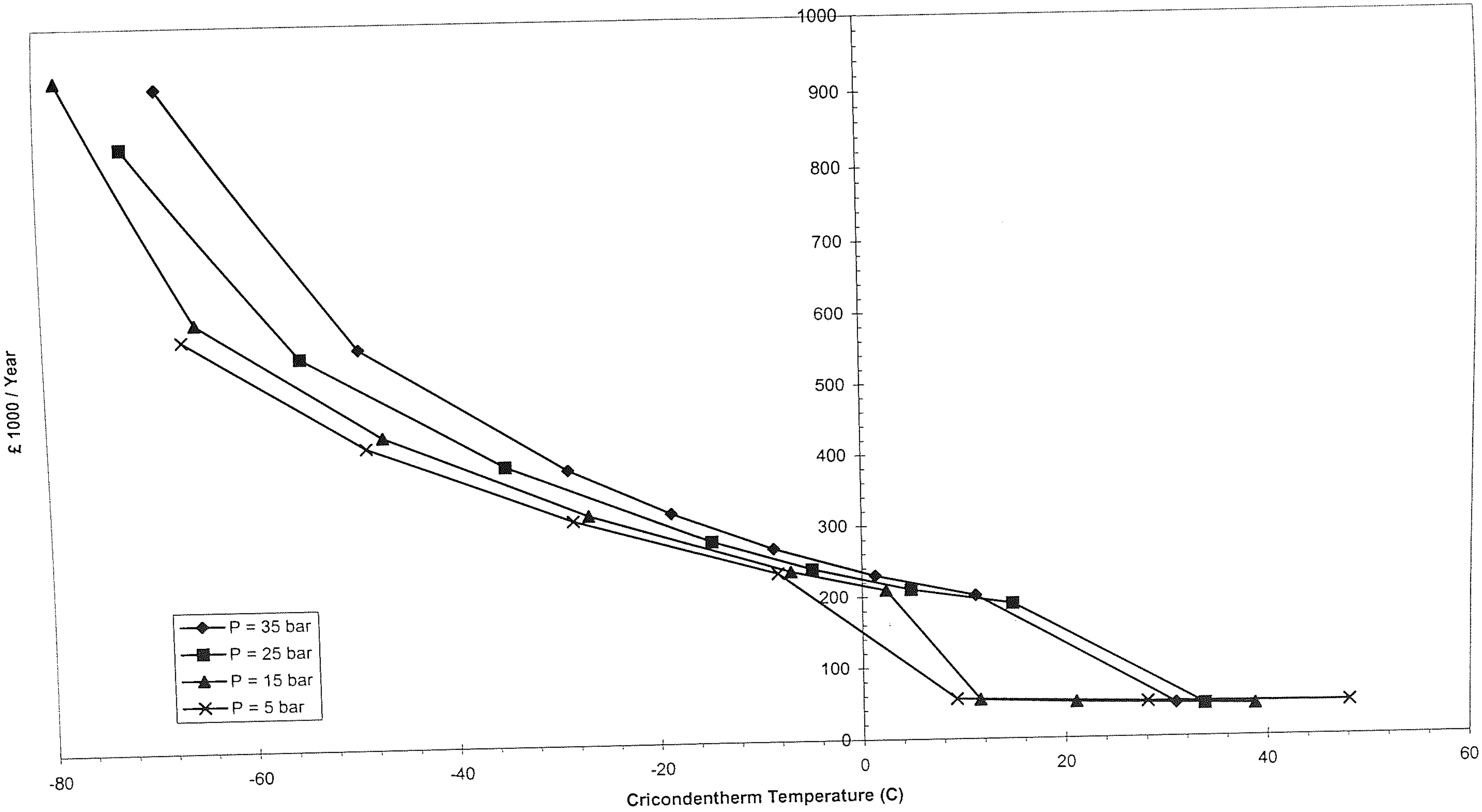


Figure 31, Case 1: Cost of gas recompression

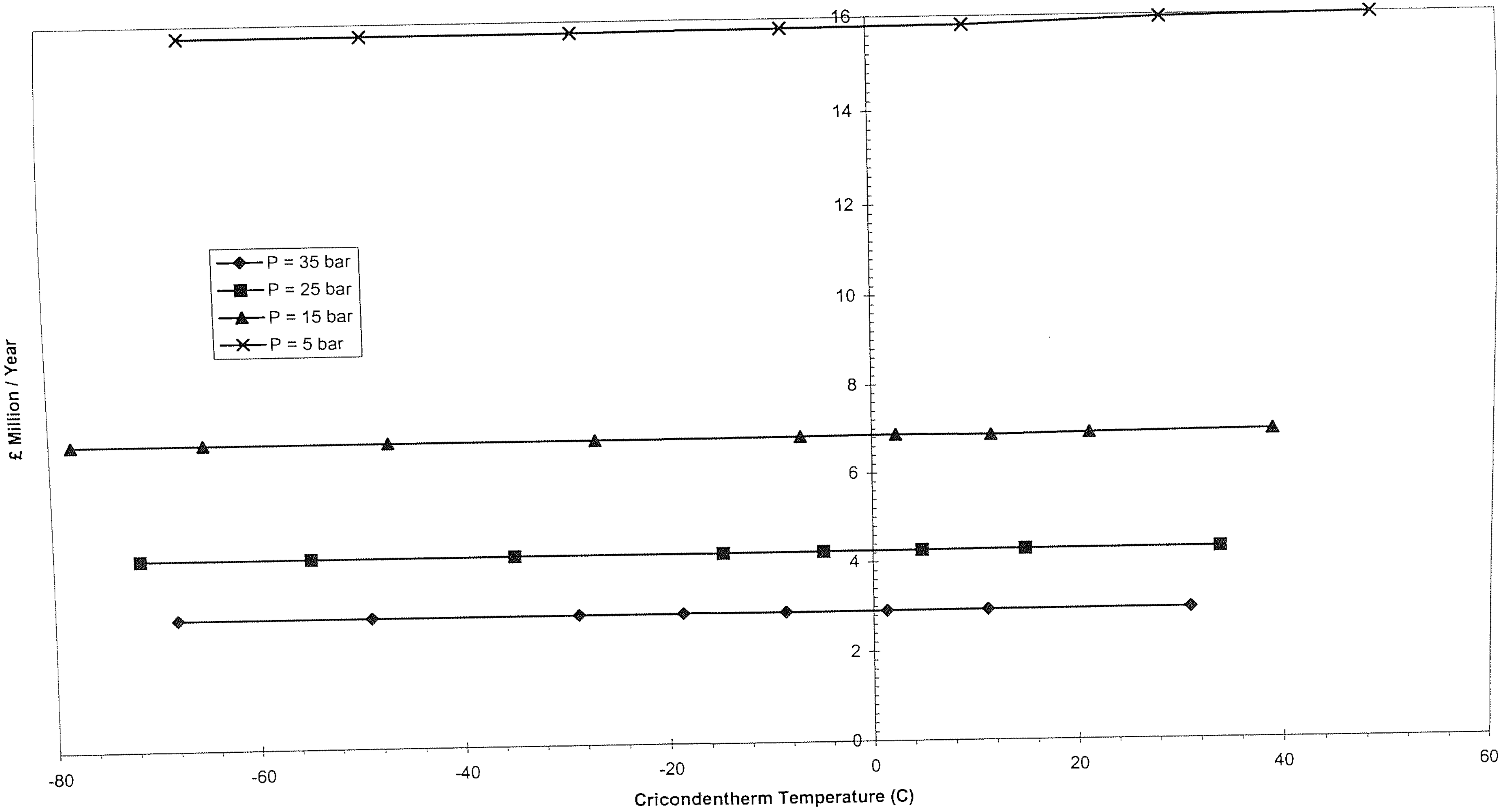


Figure 32, Case 1: Total Product value - Reprocessing cost in two phase region

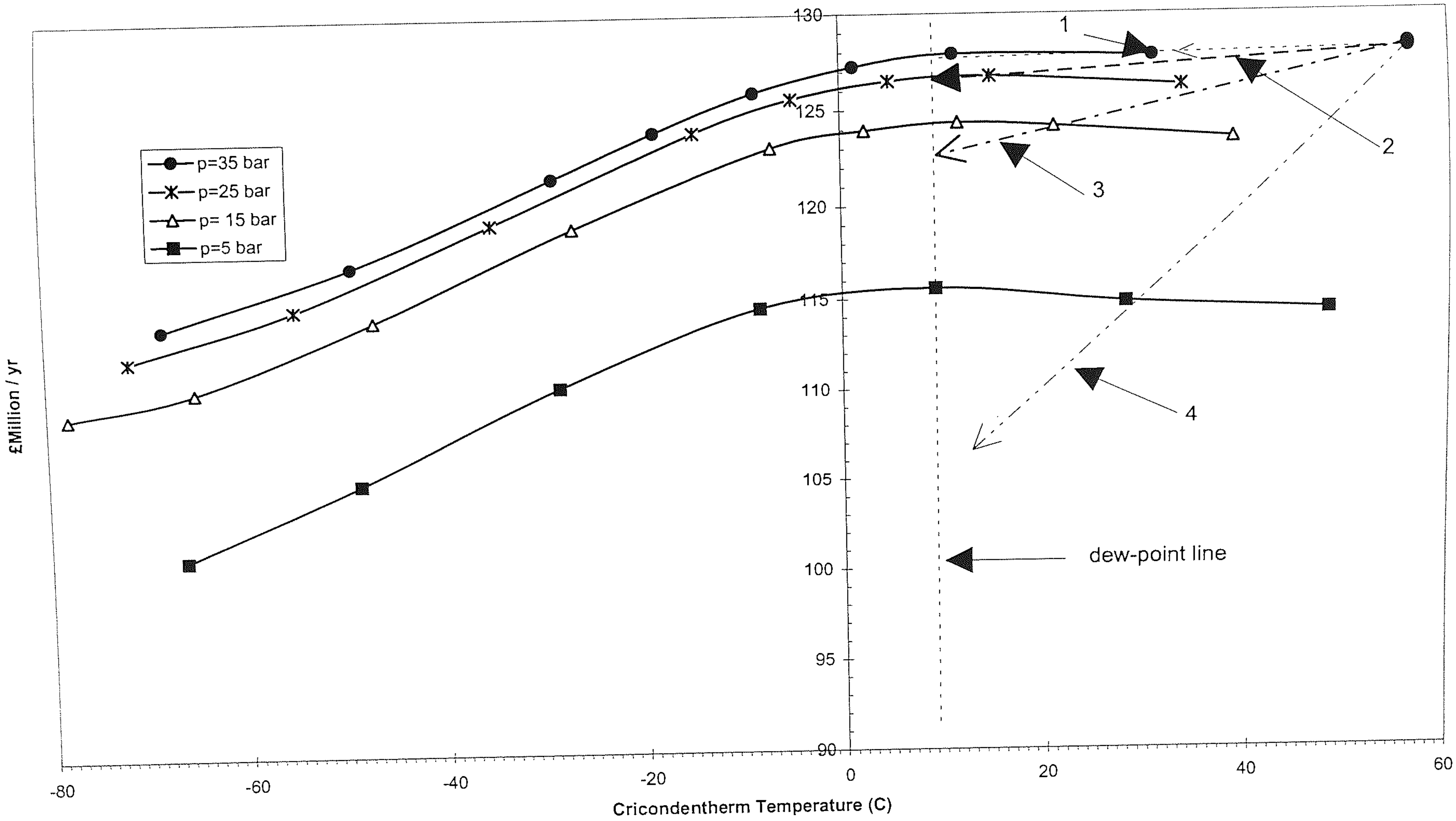


Figure 33, Case 2: Value of the gas

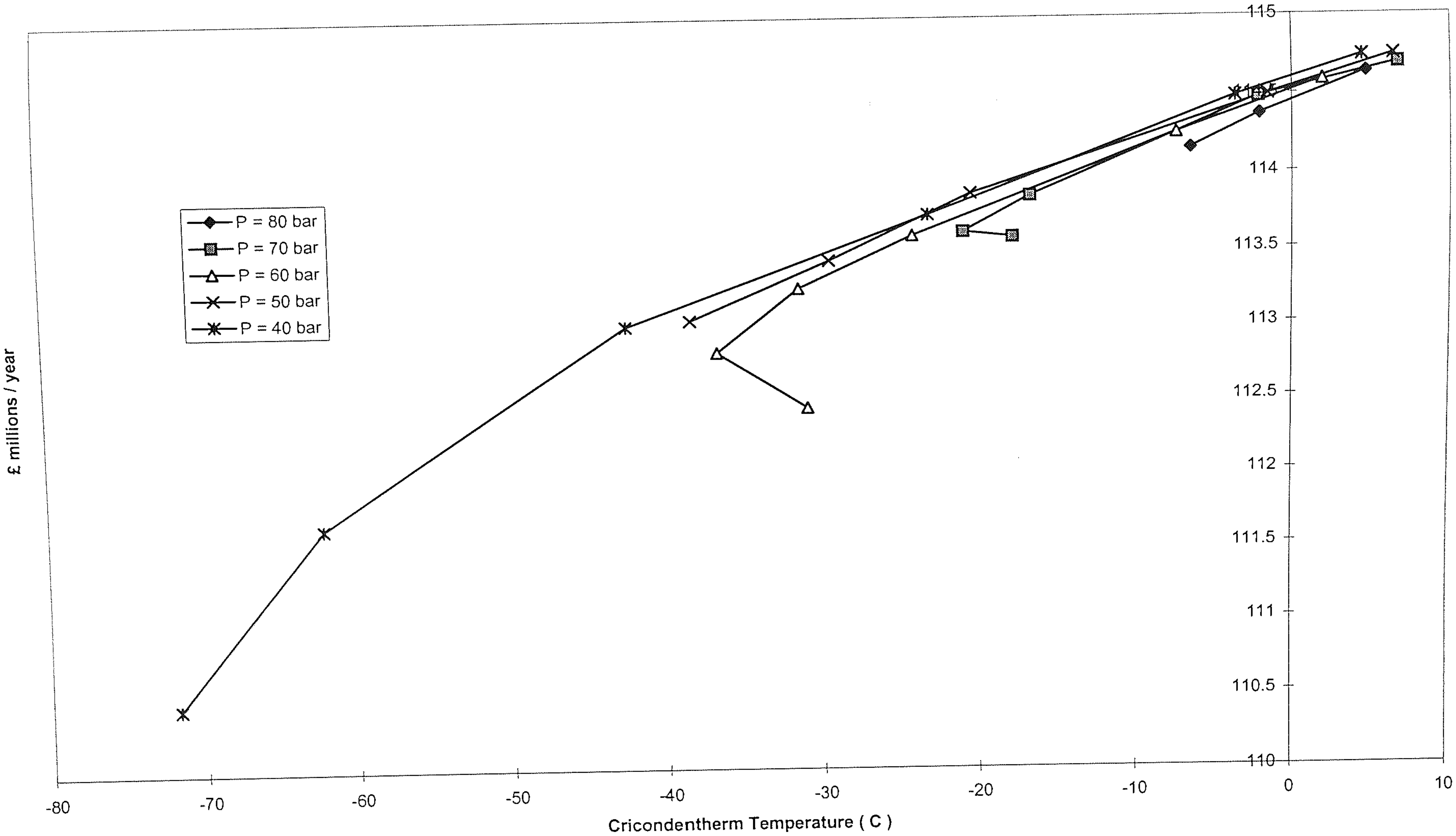


Figure 34, Case 2: Value of the liquid condensate

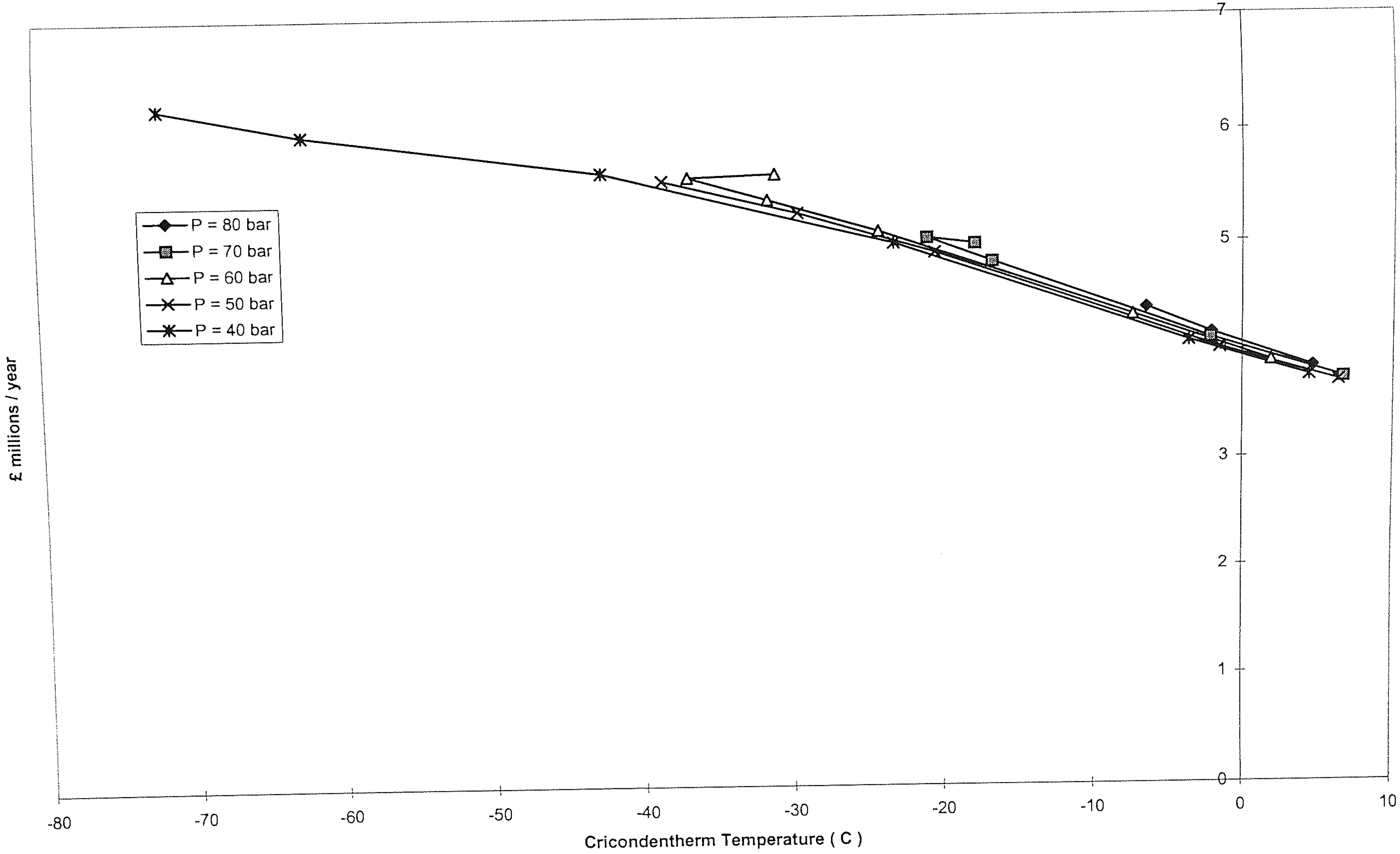


Figure 35, Case 2: Cost of gas recompression

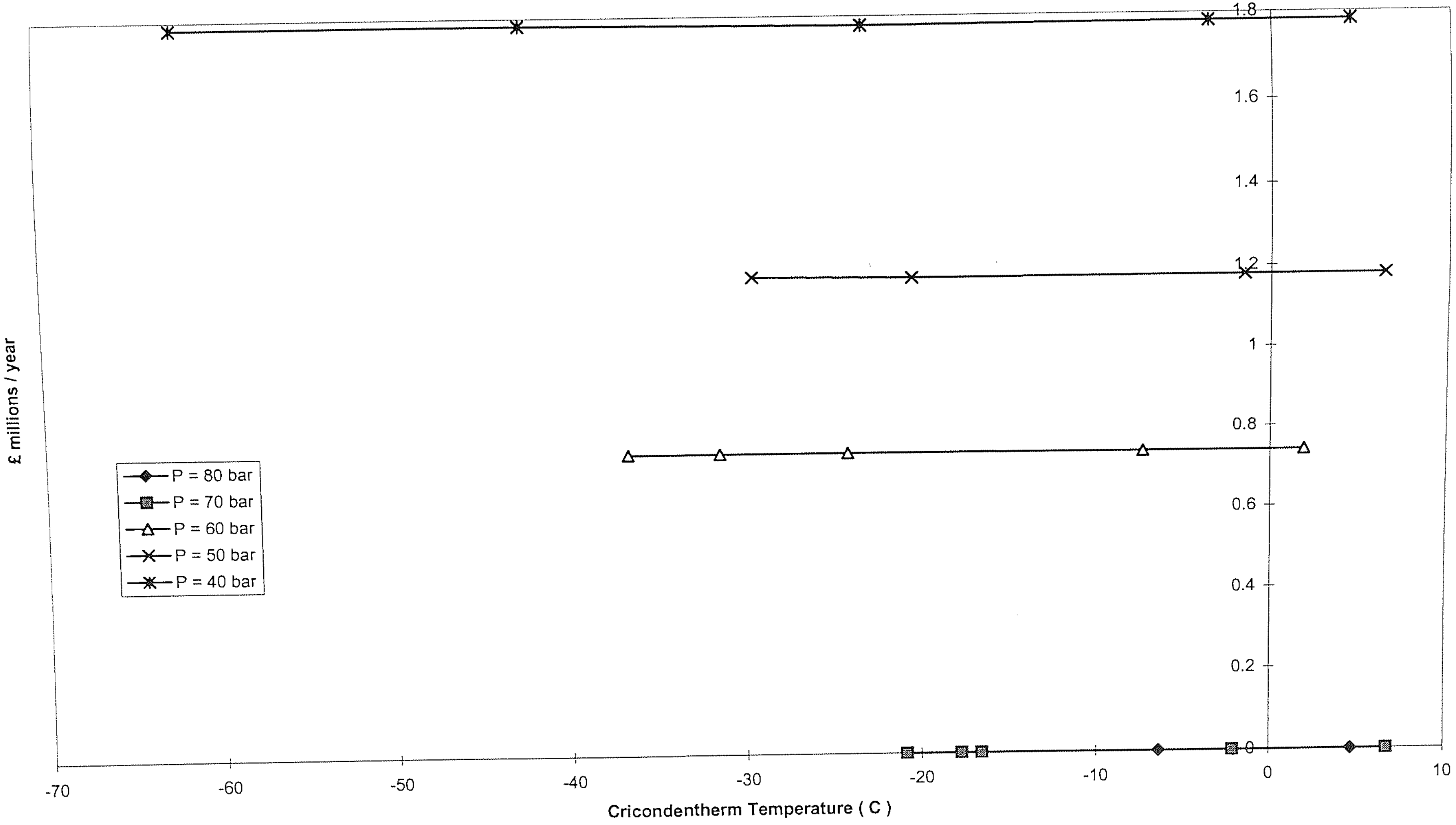


Figure 36, Case 2: Cost of liquid stabilisation

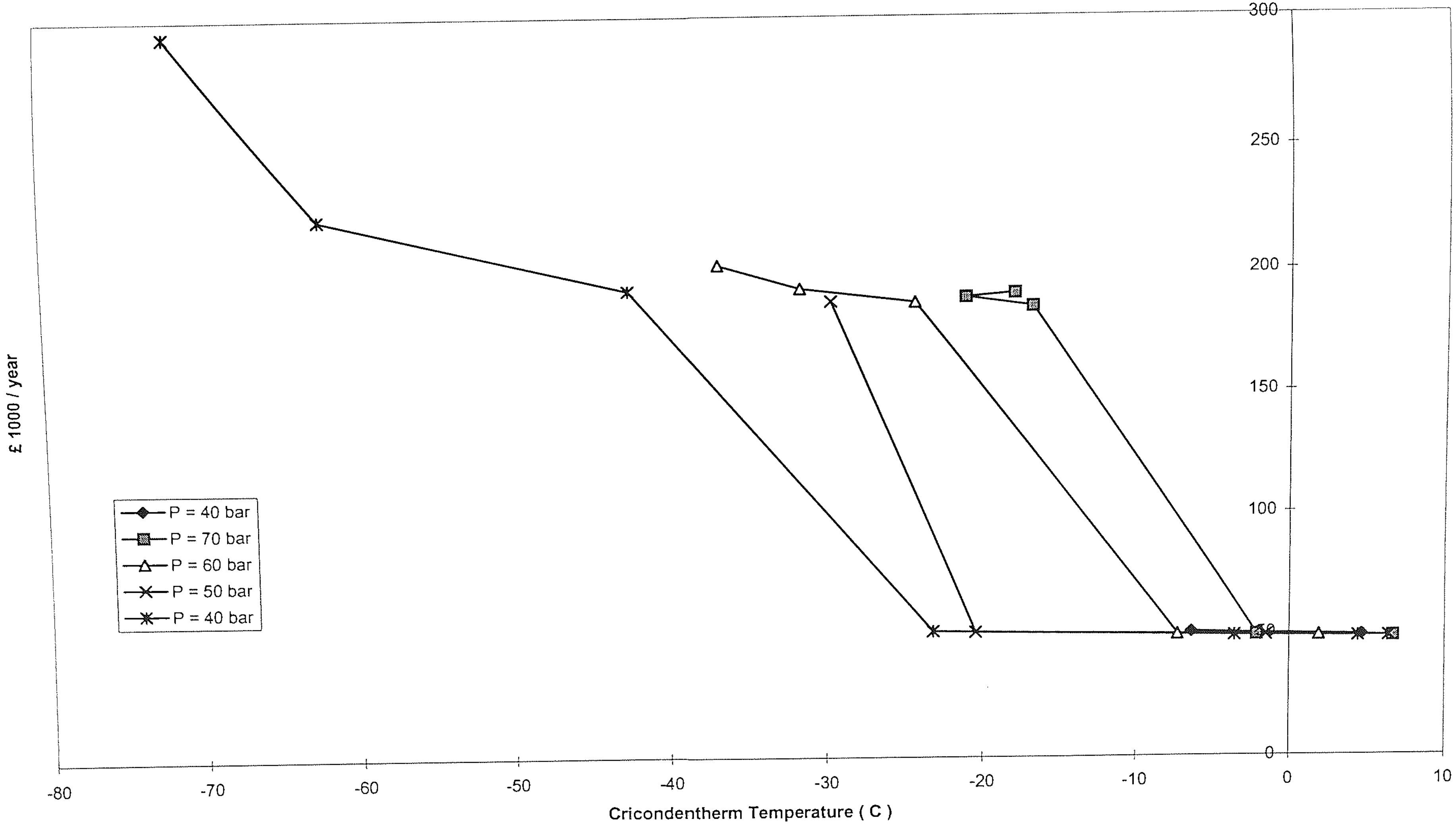


Figure 37, Case 2: Total Product value - Reprocessing cost in two phase region

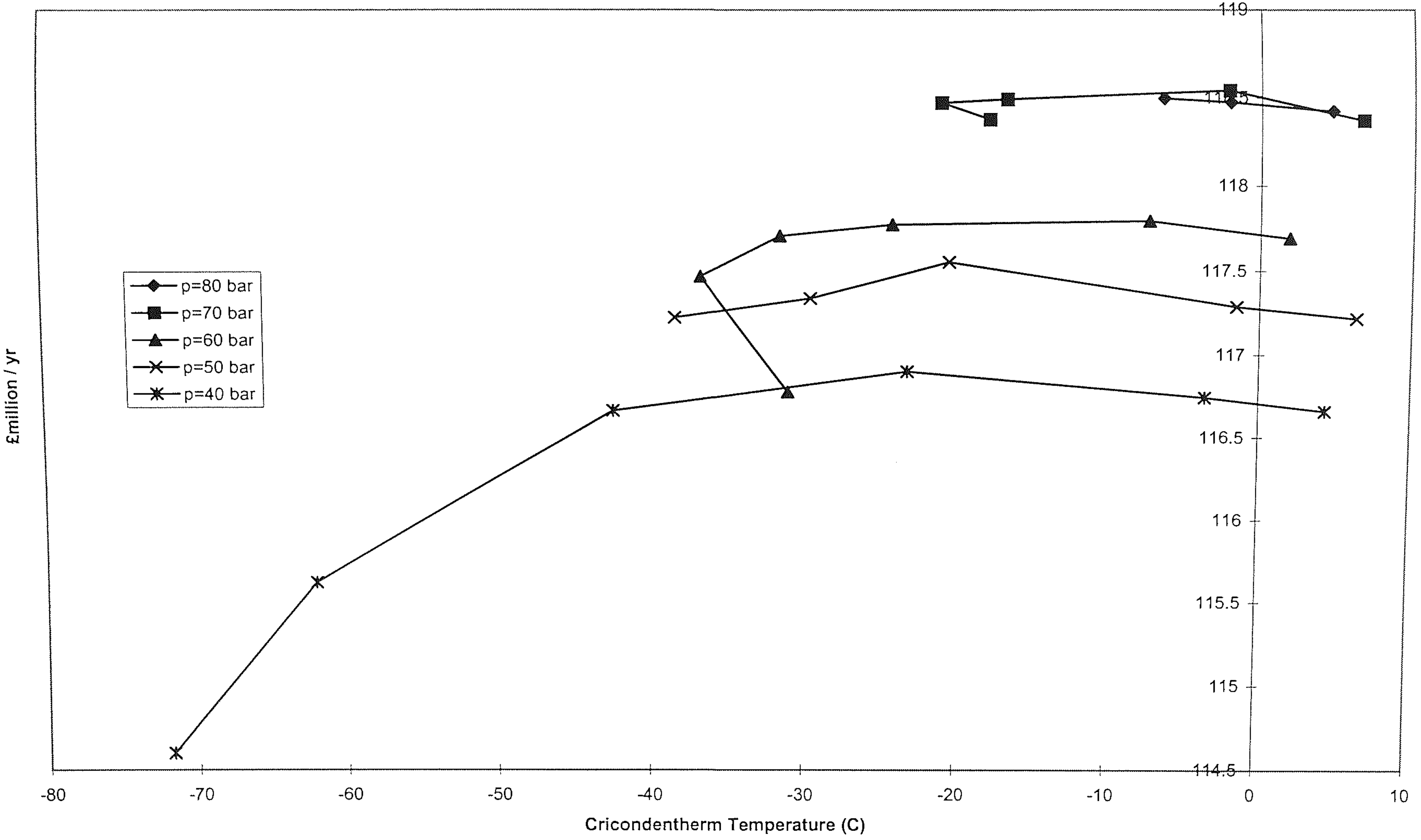


Figure 38, Case 3: Value of the gas

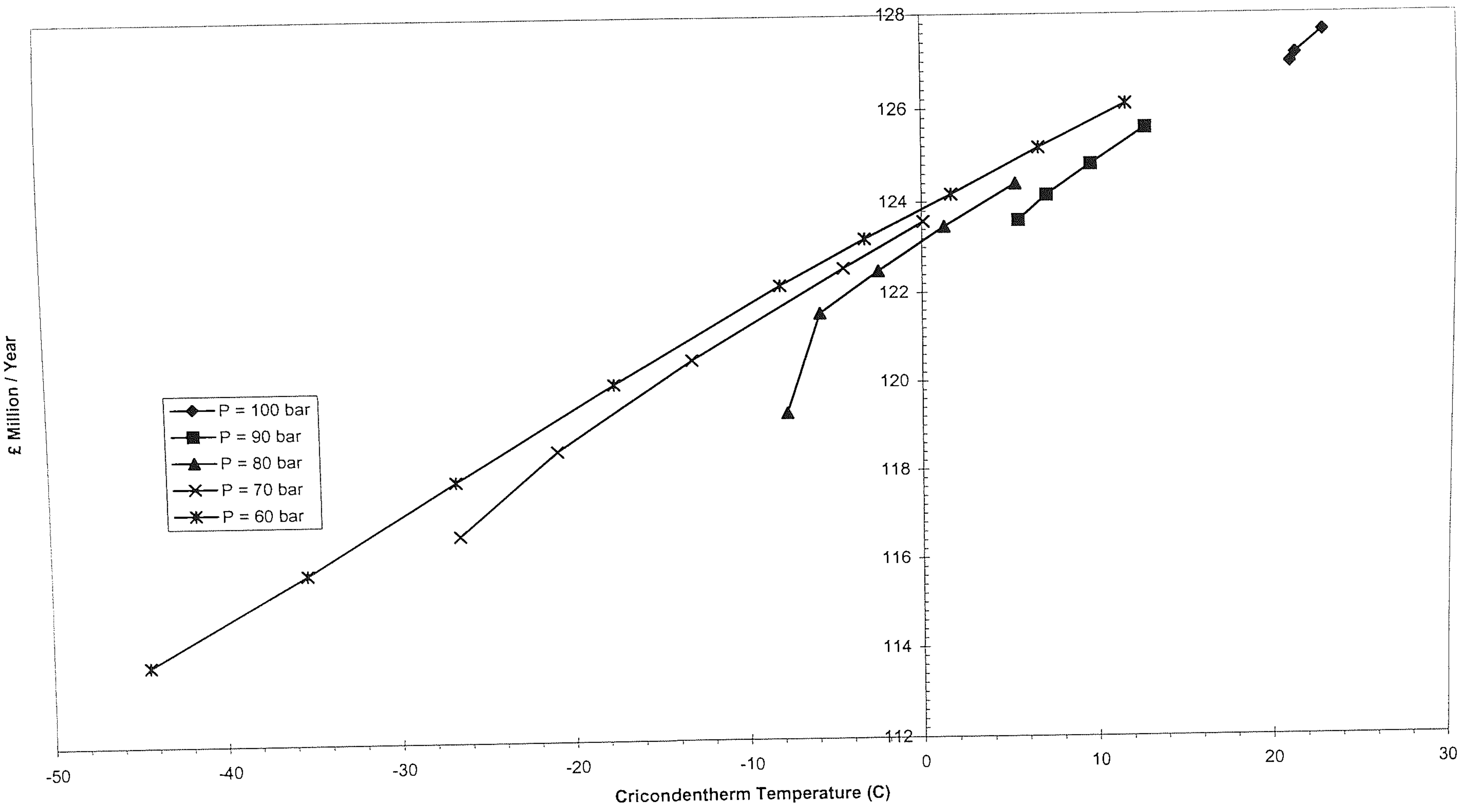


Figure 39, Case 3: Value of the liquid condensate

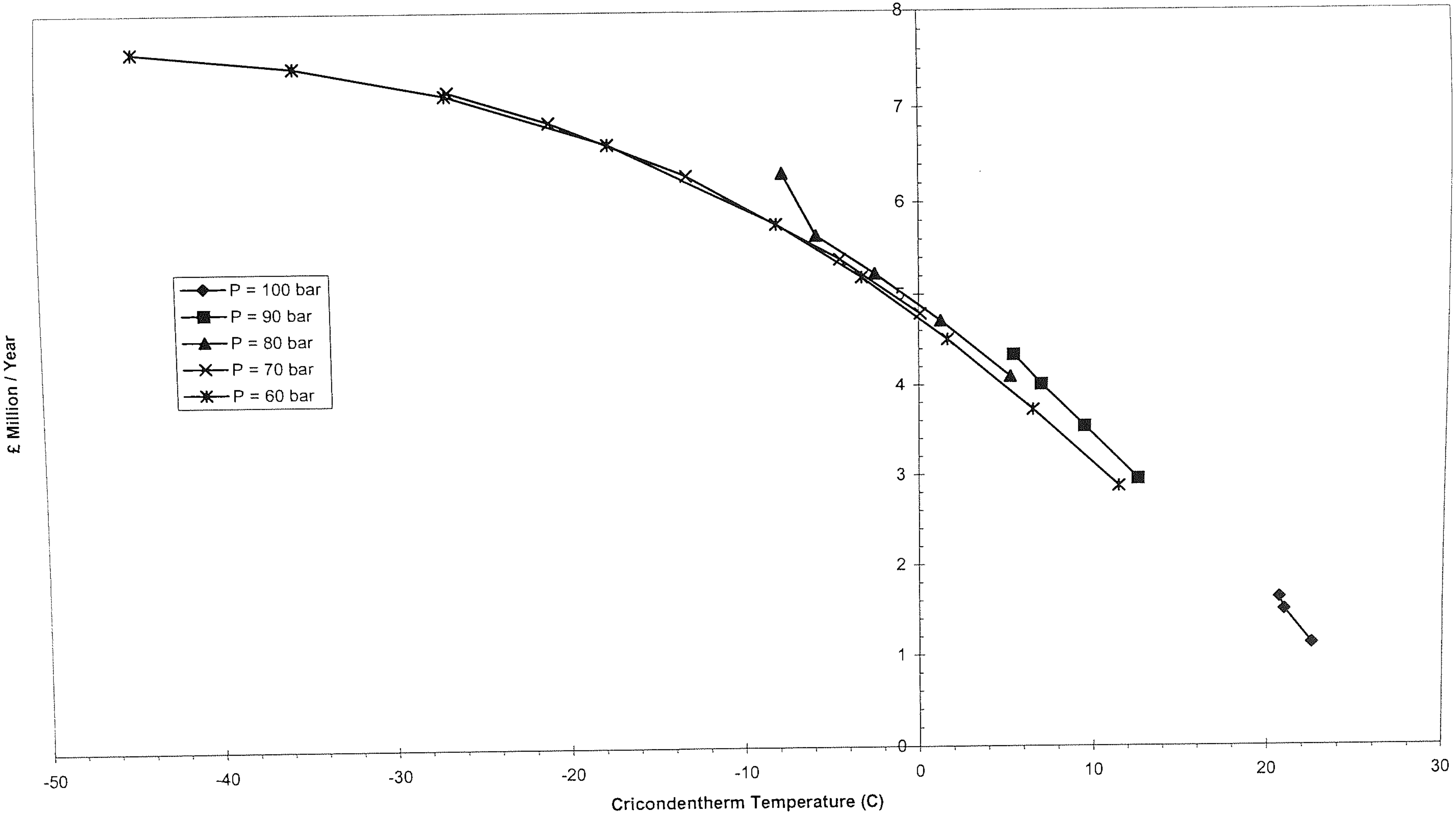


Figure 40, Case 3: Cost of gas recompression

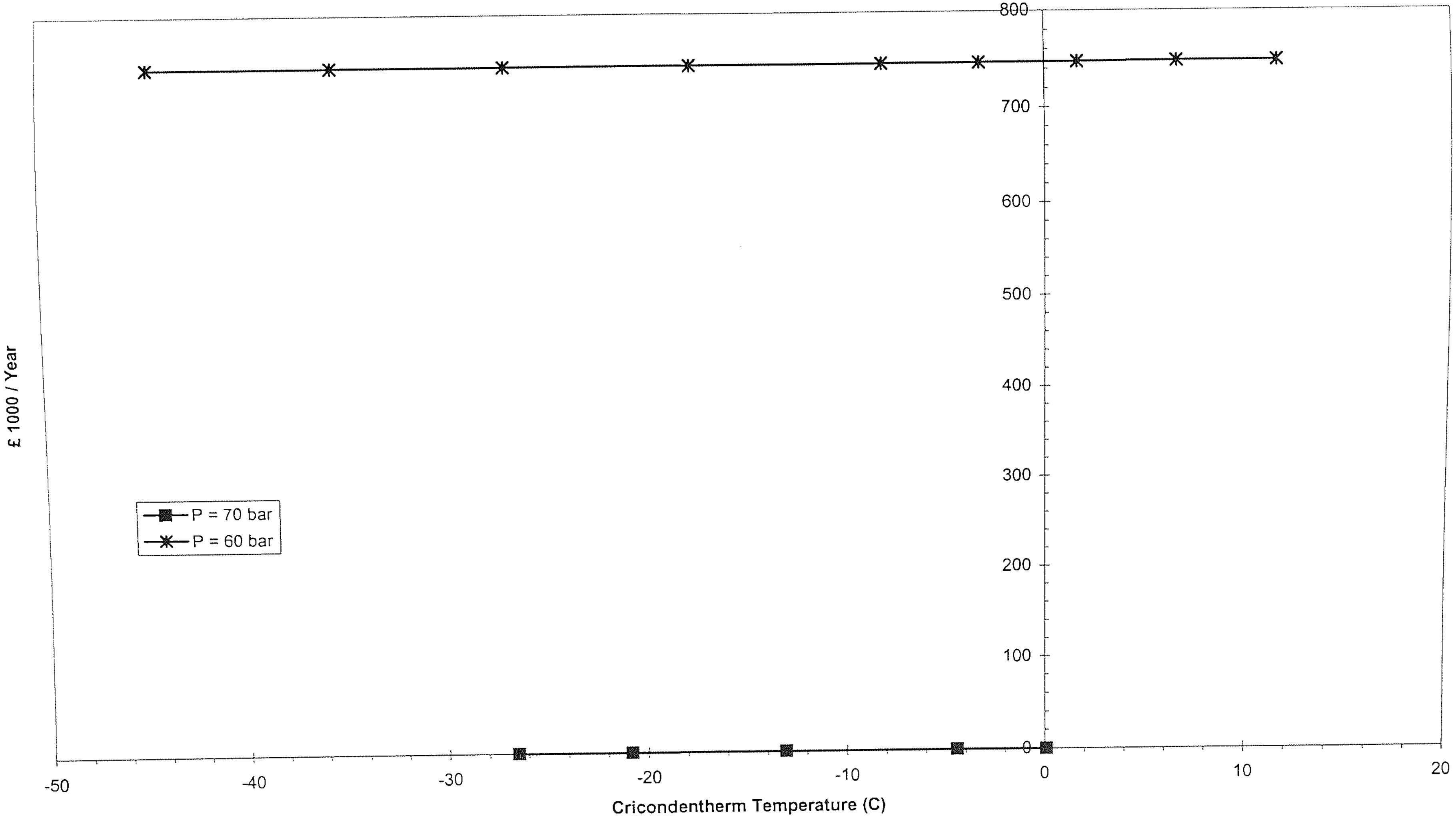


Figure 41, Case 3: Cost of liquid stabilisation

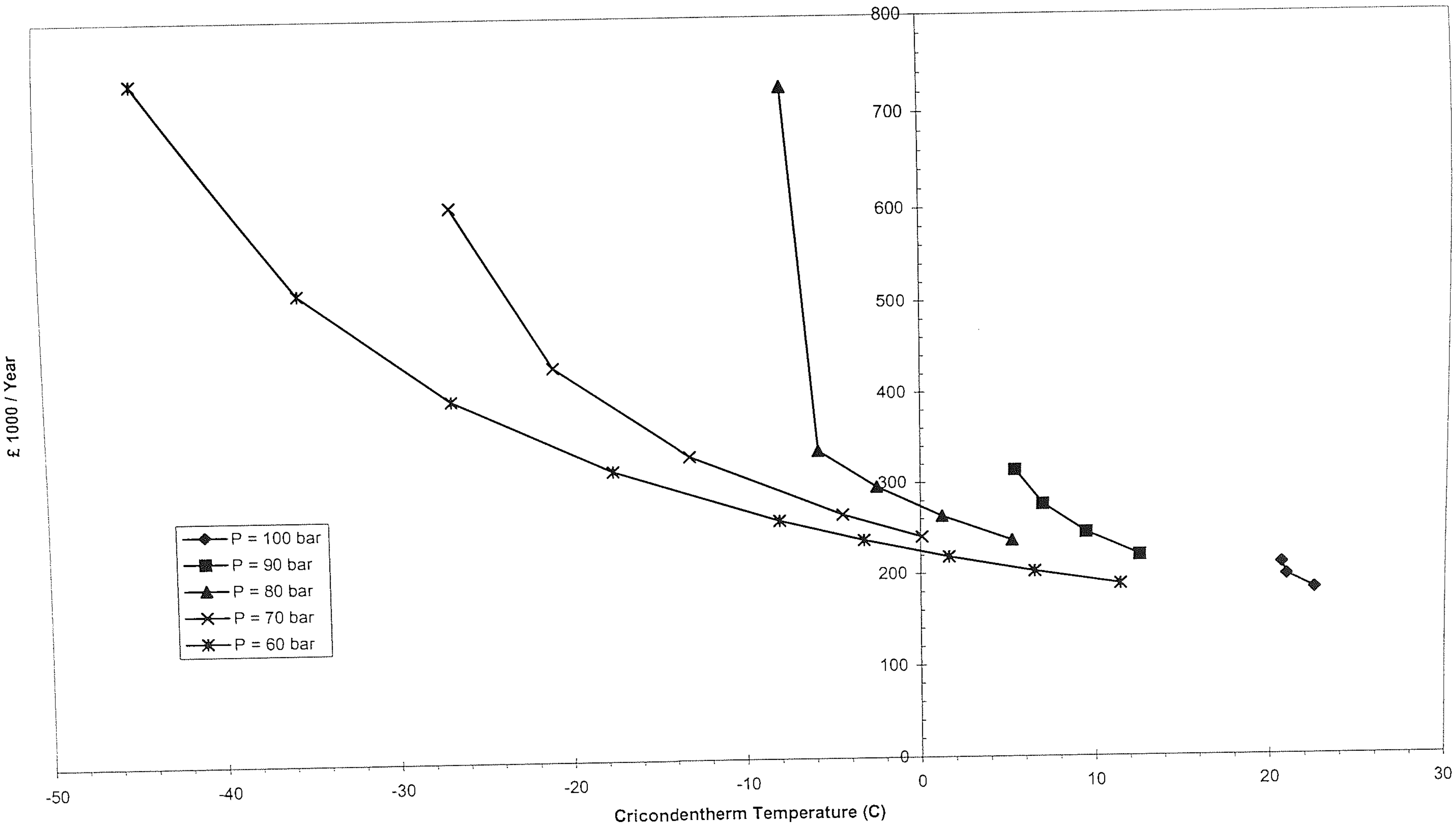
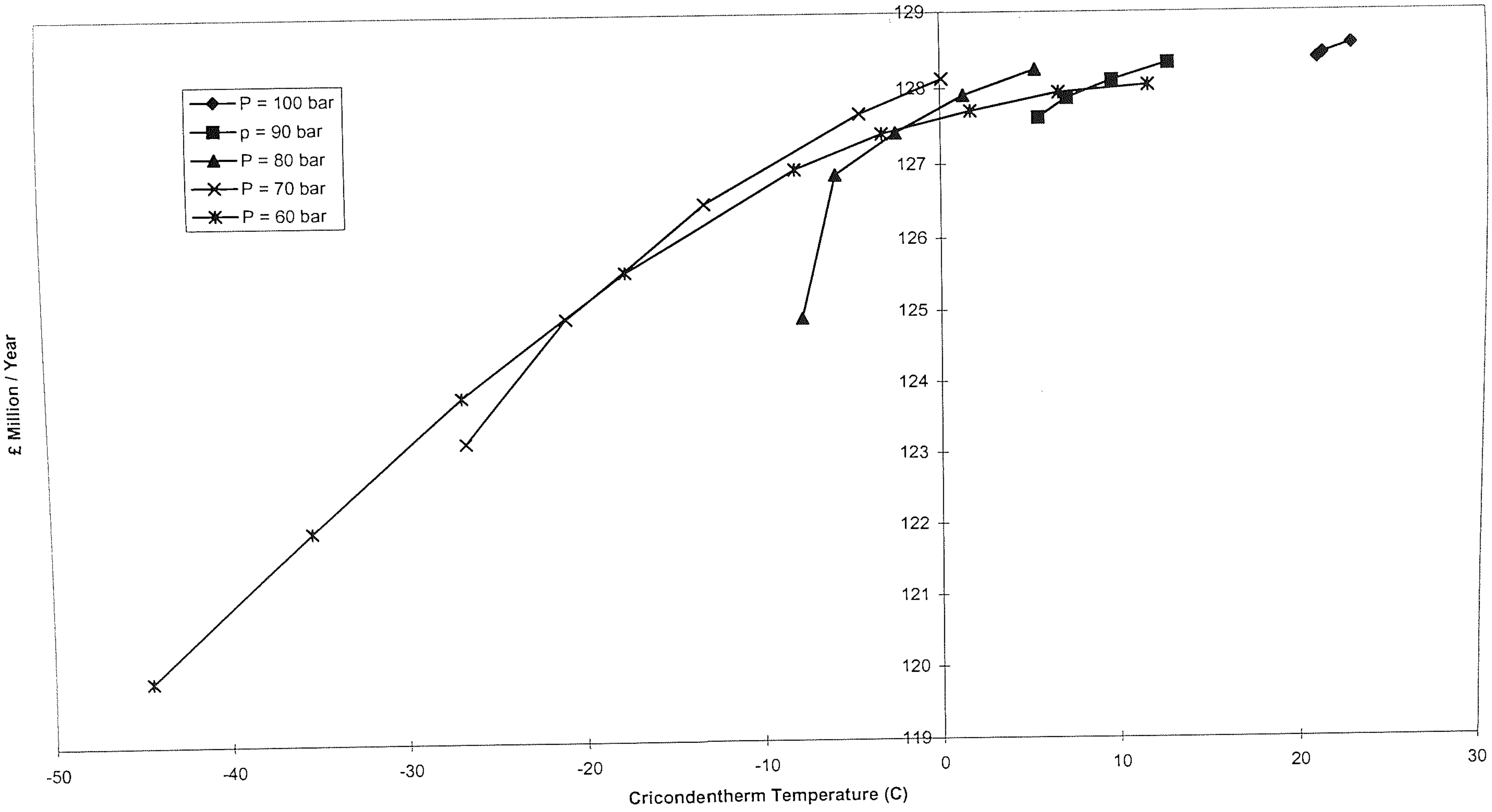


Figure 42, Case 3: Total Product value - Reprocessing cost in the two phase region



6.3 QUANTITATIVE METHOD OF PROCESS EVALUATION

This section is concerned with the equilibrium processes. Each of these processes may be represented on the cost and value diagrams by a line as shown in figures 28, 29, 30, 31, and 32. The corresponding values and costs are then used to determine the comparative costs of the alternative processes. The separator capital costs are obtained from the cost equations in chapter 5.

Case 1: On figure 32 (page 134), the constant pressure process, refrigeration is shown by the horizontal line 1, the turbo-expander by line 2, the vortex tube by line 3 and the Joule-Thomson valve by line 4, (see chapter 2, for equations defining the performance of the methods). Here the refrigeration process is taken as the base case for the comparative cost analysis. The amortised capital cost of the separators is as shown in Table 13, on page 148.

For a case such as this where the feed pressure of 35.5 bar is much lower than the specified product pressure of 84 bar, the cost of recompression can be significant. The cost increases with increasing pressure drop, hence making the pressure drop processes less attractive as shown in figure 31 on page 133. Of the equilibrium processes, turbo-expander and vortex tube both offer higher TPV than the base-case process refrigeration, as shown in Table 13. The difference in TPV between the processes is small. When the amortised reprocessing cost and the separator cost are taken into account it becomes clear that the base-case process refrigeration is the cheapest process. If the turbo-expander or vortex tube is used instead of the refrigeration process it will cost 23.5 and 99.15 percent more respectively then refrigeration. This is due to the higher cost of compression of the product gas. This shows to what extent the cost of compression dominates the overall process cost.

The Joule-Thomson valve even with a gas/gas exchanger cannot meet the transportation dew point of 10°C cricondentherm.

Case 2: Here the sales gas dew point translates into a cricondentherm of -2°C and an outlet pressure of 70 bar. Note, only 10 bar of free pressure drop is available, with the inlet feed coming in at 80 bar. For the comparative cost analysis the refrigeration process is taken as the base-case process.

The results of the analysis presented in Table 14 on page 148, show that the refrigeration process is the cheapest option, with the turbo-expander coming a close second. The vortex tube and Joule-Thomson valve are respectively 16 and 60 percent more expensive than the refrigeration process due to the high reprocessing costs. This is because the vortex tube and Joule-Thomson valve obtain cooling as a result of pressure drop and if they are to meet the specified cricondentherm temperature of -2°C the pressure falls below the required outlet pressure of 70 bar, which means gas recompression is needed.

The turbo-expander is able to compete with refrigeration with only 10 bar free pressure drop because unlike the other pressure drop processes it does not require a compressor due to its greater efficiency at cooling the gas and its ability to recover approximately two-thirds of the pressure drop in the compressor section of the expander-compressor module. As a result, the turbo-expander costs only 4 percent more than the base case method. However, the turbo-expander may be a less attractive option, if as expected the gas reservoir pressure decreases during the life of the project.

Case 3: All the pressure reducing methods can meet the required cricondentherm temperature of - 22°C. The turbo-expander process meets the cricondentherm without having to pre-cool the feed gas and the power produced in the turbine can be used either to drive a compressor or generate power. This added-value of power generation is not included in the Net Value. Even with 38 bars free pressure drop being available the Joule-Thomson valve and vortex tube processes still require small amount of recompression, in the region of 1 to 2 bar respectively.

If the turbo-expander is taken as the base-case process, the analysis shows that it costs half as much as against the Joule-Thomson valve or vortex tube process. The loss in the total product value in the case of the Joule-Thomson valve and vortex tube processes of over £300, 000, with respect to the base case is taken as a penalty cost. The reprocessing costs of the two methods are also much higher than the turbo-expander. Only the capital cost of the turbo-expander is slightly higher than the other two processes, but when all the costs are considered, the base-case process, the turbo-expander, is the best process.

For refrigeration to meet the dew point specification the feed pressure has to be reduced to below 80 bar due to the shape of the liquid fraction lines in the two-phase region. This will require the combination of two processes, it will therefore be considered in chapter 8.

In the next chapter we will look at the non-equilibrium separation processes.

Table 13: case 1-comparative cost analysis for equilibrium processes.

Process	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	TPV_{bc} - TPV	Comparative Separation Cost	% difference
Turbo-expander	131.32	4290.77	578.80	-38.22	4831.34	23.52
Refrigeration	130.94	3281.54	629.89	0.00	3911.43	0.00
Vortex tube	131.22	7530.00	287.89	-28.24	7789.65	99.15

Table 14: case 2-comparative cost analysis for equilibrium processes.

Process	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	TPV_{bc} - TPV	Comparative Separation Cost	% difference
Joule-Thomson	118.60	680.49	270.63	14.08	965.20	60.67
Turbo-expander	118.62	49.30	578.84	-2.33	625.81	4.18
Refrigeration	118.62	49.35	551.37	0.00	600.72	0.00
Vortex tube	118.63	680.69	35.56	-14.23	702.02	16.86

Table 15: case 3-comparative cost analysis for equilibrium processes.

Process	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	TPV_{bc} - TPV	Comparative Separation Cost	% difference
Joule-Thomson	125.08	872.21	270.63	324.66	1467.50	100.84
Turbo-expander	125.41	398.98	331.72	0.00	730.70	0.00
Vortex tube	125.09	751.00	264.78	318.49	1334.27	82.60

6.4 INTERIM OBSERVATIONS

1. A graphical representation of the condensate stabilisation and gas recompression has been developed for all mixtures in the two-phase region for the feed compositions being considered.

2. The methodology can be used to illustrate why one process has an advantage over another.

3. All the processes very easily can be located on the value and cost diagrams as operating lines very easily.

4. Using the selection methodology developed, it is possible to determine the optimum process operating point for the required product specifications.

5. The methodology is generally applicable for any given feed and/or product gas specification.

6. The methodology illustrates the economic differences between the separation processes.

7. The methodology takes into account the loss in total product value as a result of the efficiency of the processing method as well as the added costs.

CHAPTER 7

7.0 NON-EQUILIBRIUM PROCESSES

In the previous chapter we looked at the separation methods that follow equilibrium principles, such as the turbo-expander. In this chapter we will concentrate on the non-equilibrium processes; as noted previously these are:

Absorption
Adsorption
Membrane processes
Refluxing condensers

The same principles as used for selecting between the equilibrium processes are used here, i.e., they are based on comparative cost. The process with the lowest overall processing cost is then chosen as the best method in the conceptual design phase of the project.

A further complication in this case is that each method separates the heavy hydrocarbons from the gas stream in a different way. Each has its own complexities and technical characteristics, for example the adsorption and membrane processes are able to operate outside the two-phase region of the phase envelope. These processes therefore have to be evaluated separately.

Note that as for the equilibrium processes, further processing of the products may be required. Stabilisation of the liquid produced will be required in some cases, whereas for the product gas recompression will only be necessary if the feed pressure is below the specified product pressure, since these are constant pressure processes.

One of the advantages of the non-equilibrium processes is that they maintain the product pressure. This can be very useful and cost effective over the long term, especially when there is not much free pressure drop available.

Some of the cost savings as compared with equilibrium processes that may be expected are shown in the following table.

Table 21: Potential cost savings.

Process	Gas Recompression	Liquid Stabilisation	Higher Value Products	Capital Cost	Complexity
Membrane	No	Yes	No	low	Low
Absorption	No	Yes	No	High	High
Adsorption	No	No	Yes	low	High
R. Condenser	No	Maybe	Maybe	Medium	Medium

By inspection of the table we infer the following:

1) The membrane process only if the feed gas is relatively lean, with no great amount of heavy components to be removed and gas recompression costs are high.

2) The absorption process for heavy duties such as Natural Gas Liquid recovery, when propane and heavier components need to be removed.

3) Adsorption processes are ideal for lean gas feeds with only small quantities of heavy components needing to be removed and if we are trying to minimise the reprocessing costs.

4) Refluxing condenser - similar to (3).

7.1 CHOOSING BETWEEN THE NON-EQUILIBRIUM PROCESSES

Any non-equilibrium process can be represented as in figure 43 as a flow diagram. The process may also be illustrated on the phase diagram, figure 44.

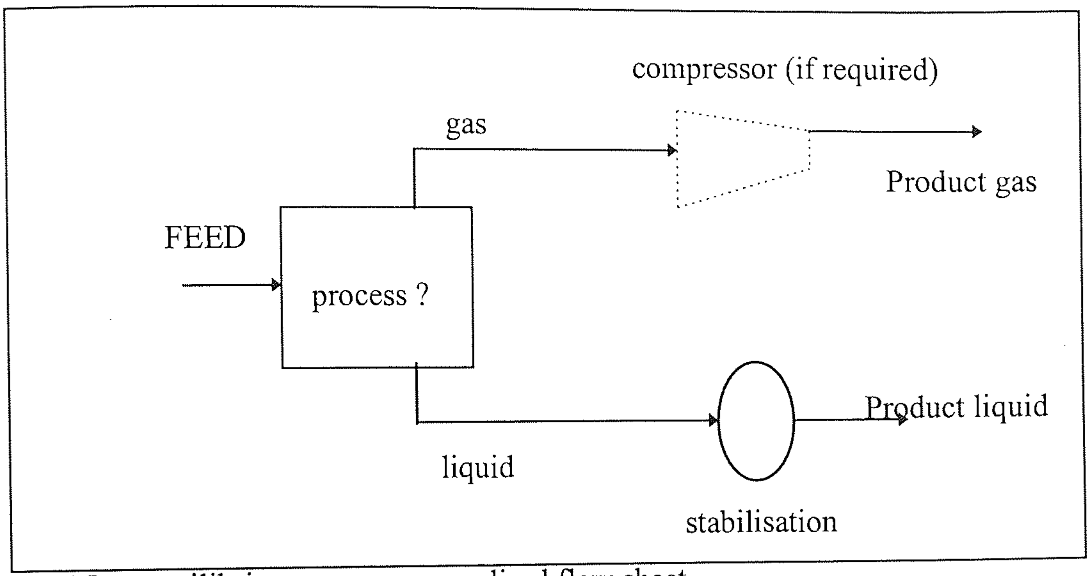


Figure 43: Non-equilibrium process generalised flow sheet.

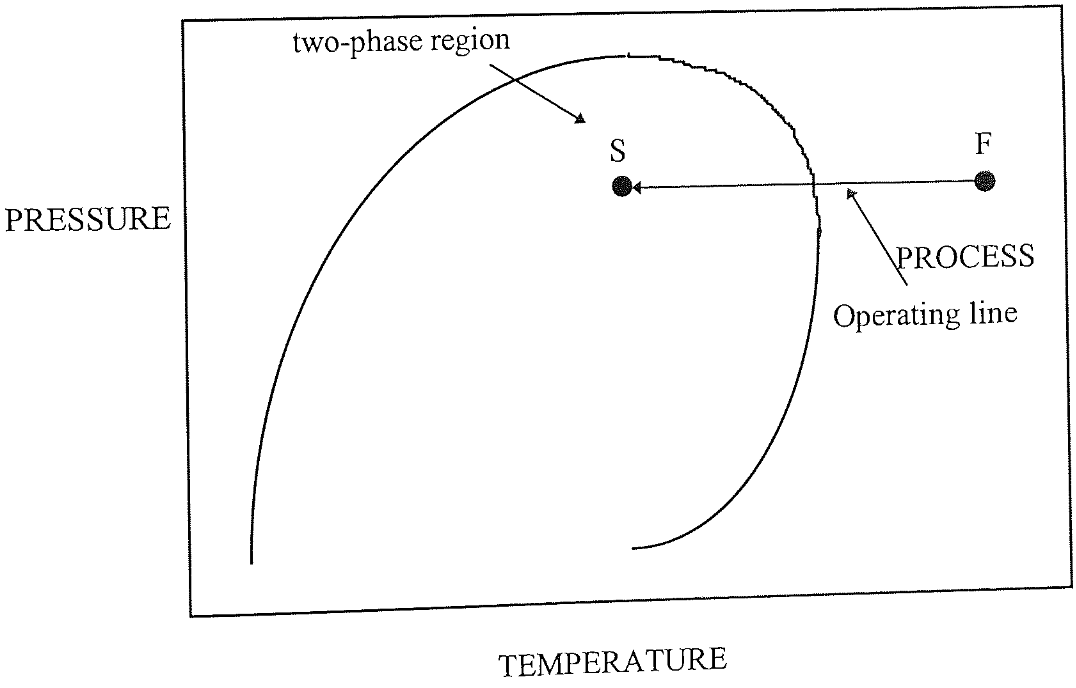


Figure 44: Typical phase diagram.

Here the feed at condition F is caused by the process to go to condition S. For the non-equilibrium processes it represents:

a) a gas stream at the feed pressure and

b) a gaseous and / or liquid stream for stabilisation

From this the cost of stabilisation (and recompression) and the value of the products can be calculated at point S. This procedure can be repeated for any given feed and product specification. For adsorption and membrane processes point S is not restricted to the two-phase region.

An example calculation for a process (membrane) is given in Appendix A1-7.

The calculated costs and values may be presented as cost and value lines. These are illustrated in figures 45-54, on pages 154-163.

The feed gas conditions and product specifications are as set out in chapter 6.

Figures 45-48, on pages 154-157 are for feed 1 and transport pipeline specification.

Figures 49-51, on pages 158-160 are for feed 2 and sales gas specification.

Figures 52-54, on pages 161163 are for feed 3 and sales gas specification.

Figures 45 to 54 were constructed by selecting a number of points and calculating the compositions, flow rates, costs and values for each point by using the BP GENESIS flowsheeting package, as illustrated in Appendix A1-7. When using this method of process selection, the calculations need only be carried out at the specified product dew-point conditions to compare them with the equilibrium processes, but not over the whole range as illustrated in the above charts.

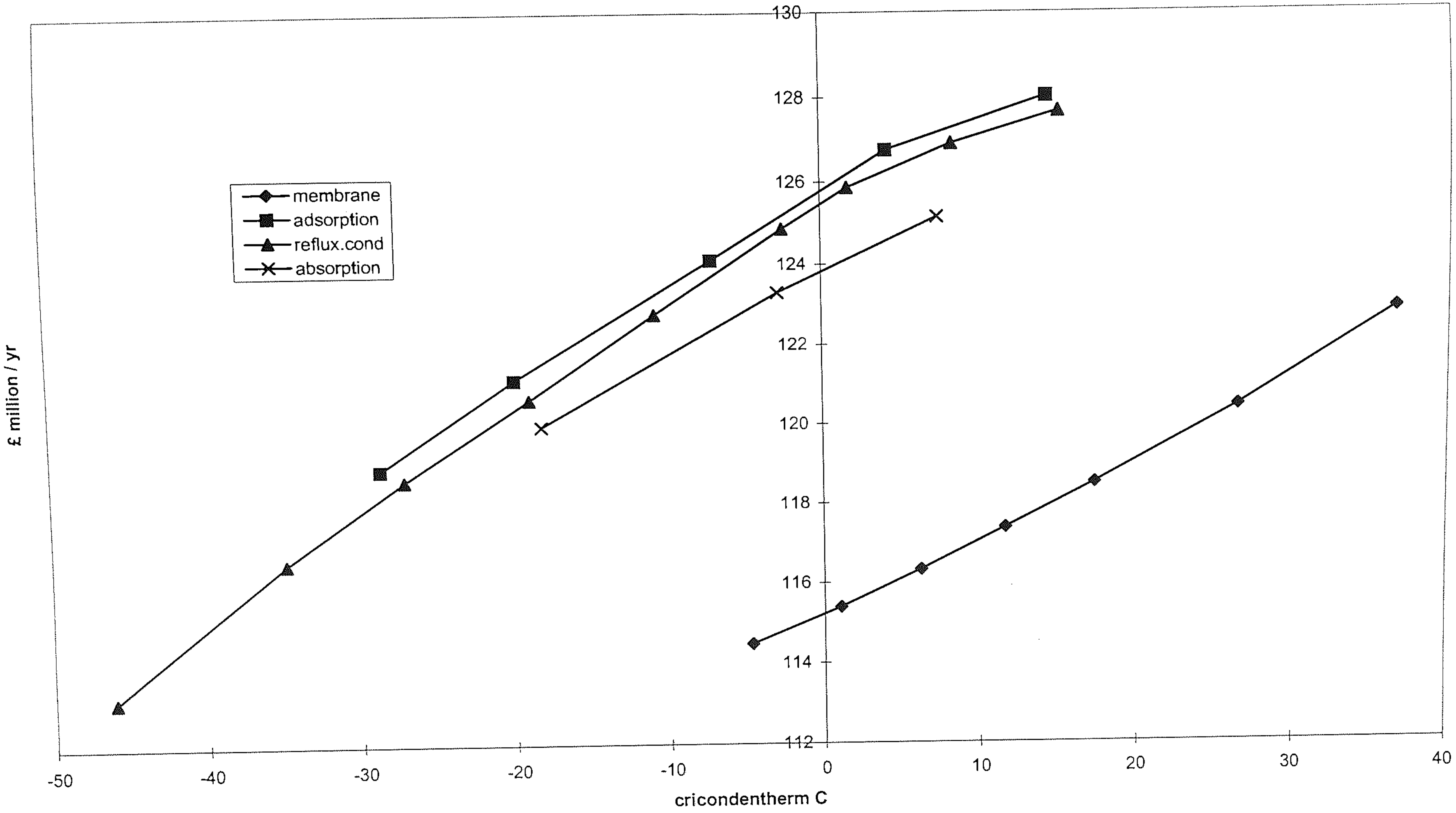


Figure 45, Case 1: Value of the gas

Figure 46, Case 1: Value of the liquid condensate

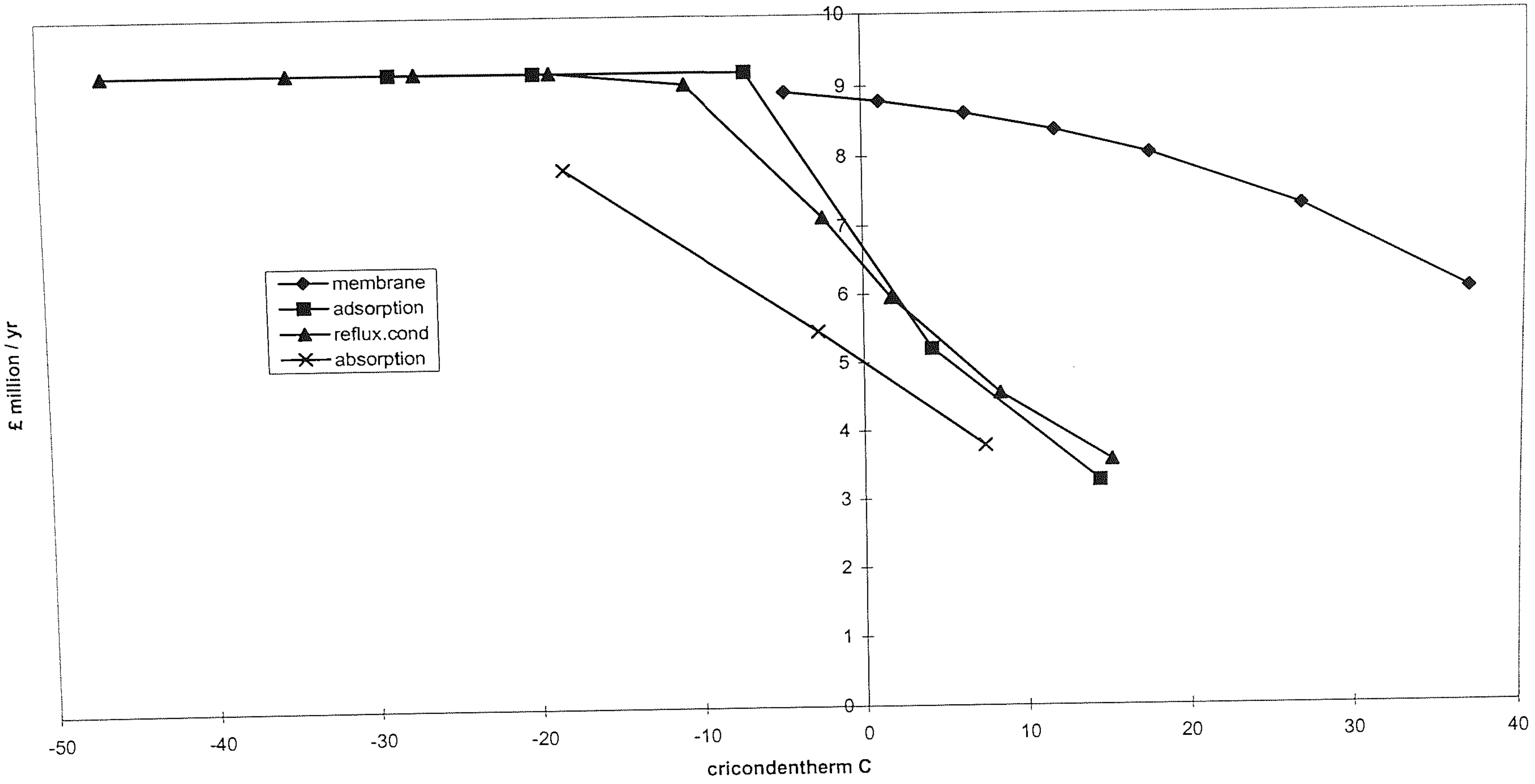


Figure 47, Case 1: Cost of stabilisation

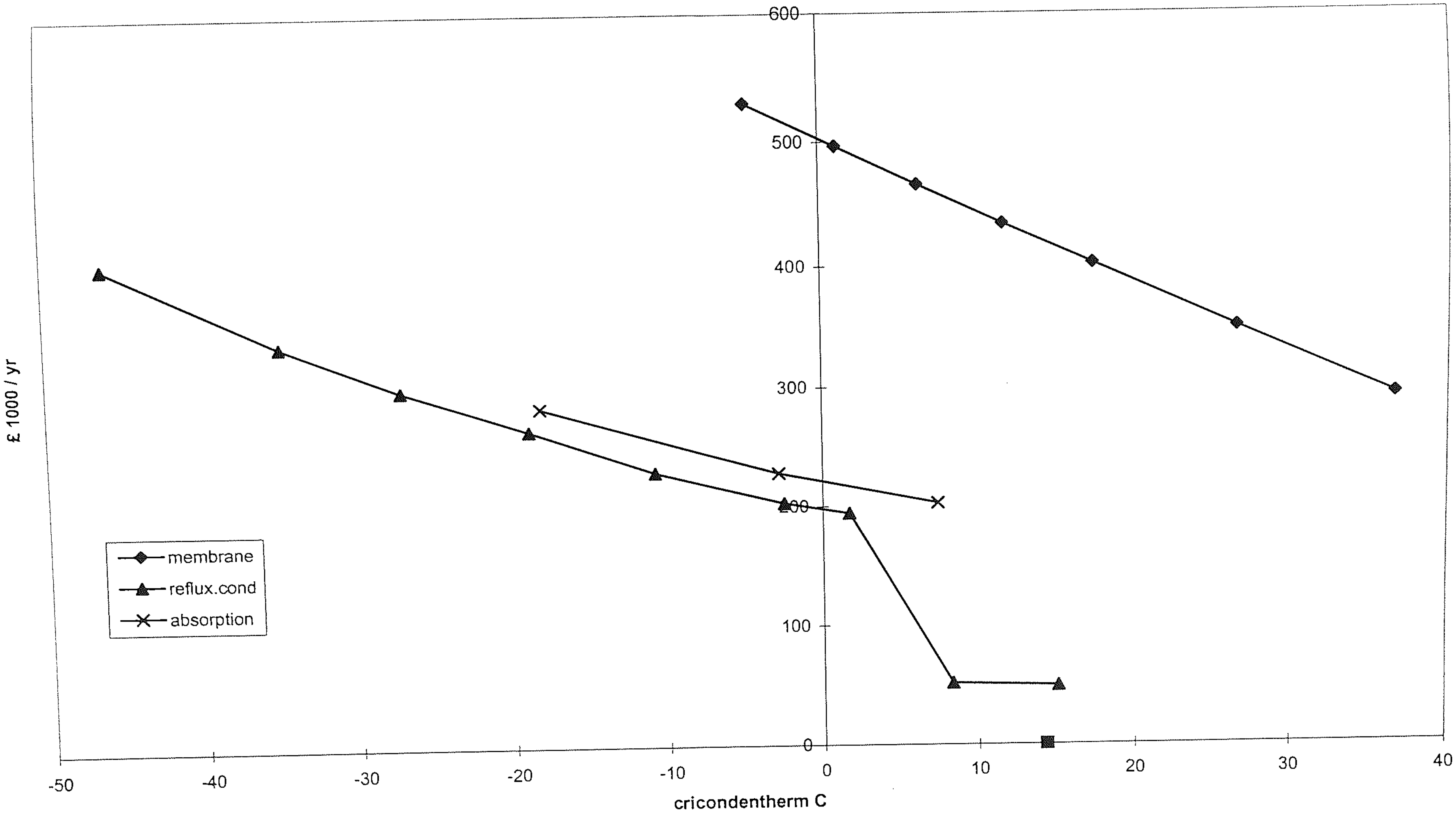


Figure 48, Case 1: Cost of recompression

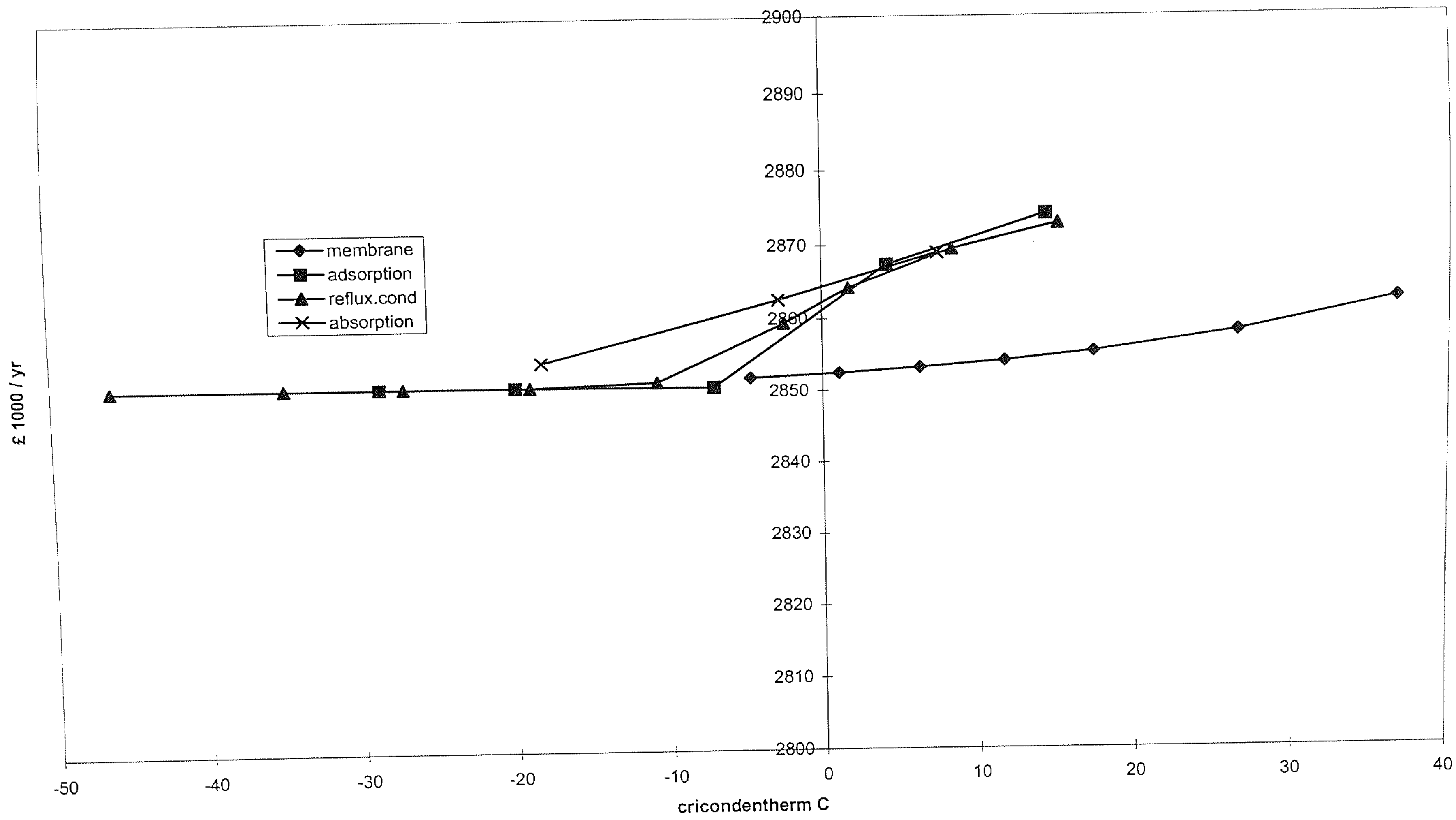


Figure 49, Case 2: Value of the gas

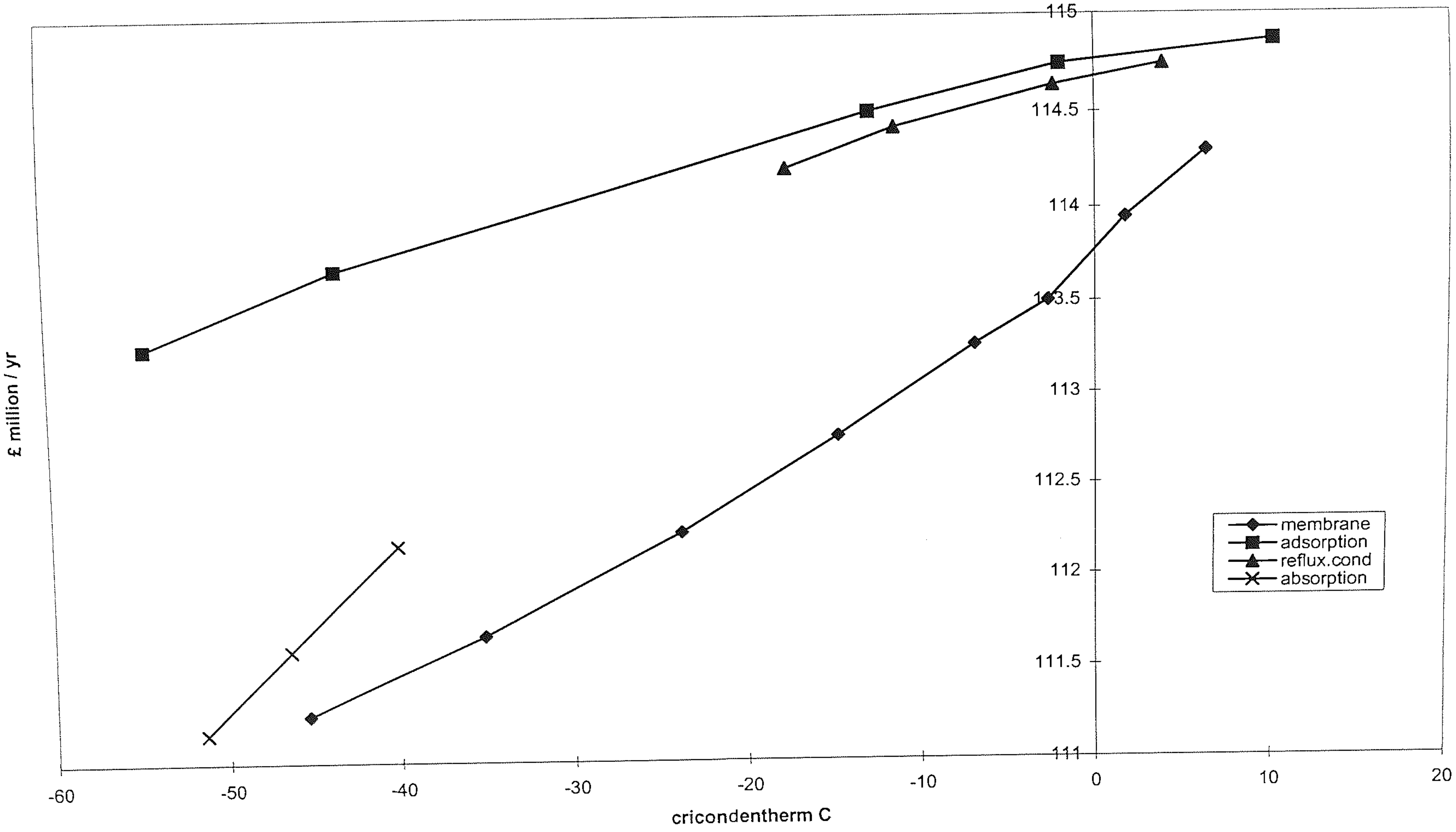


Figure 50, Case 2: Value of the liquid condensate

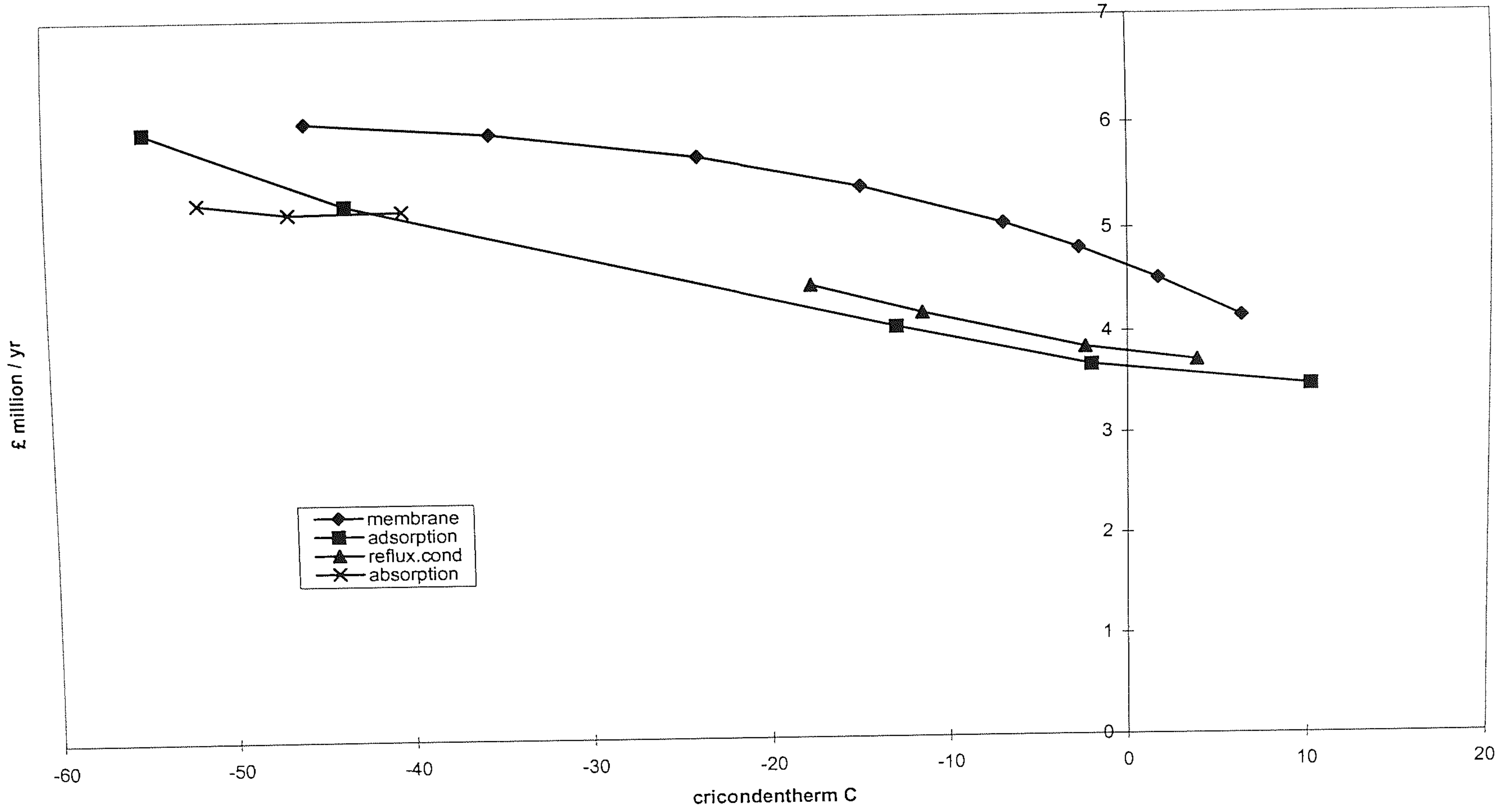


Figure 51, Case 2: Cost of condensate stabilisation

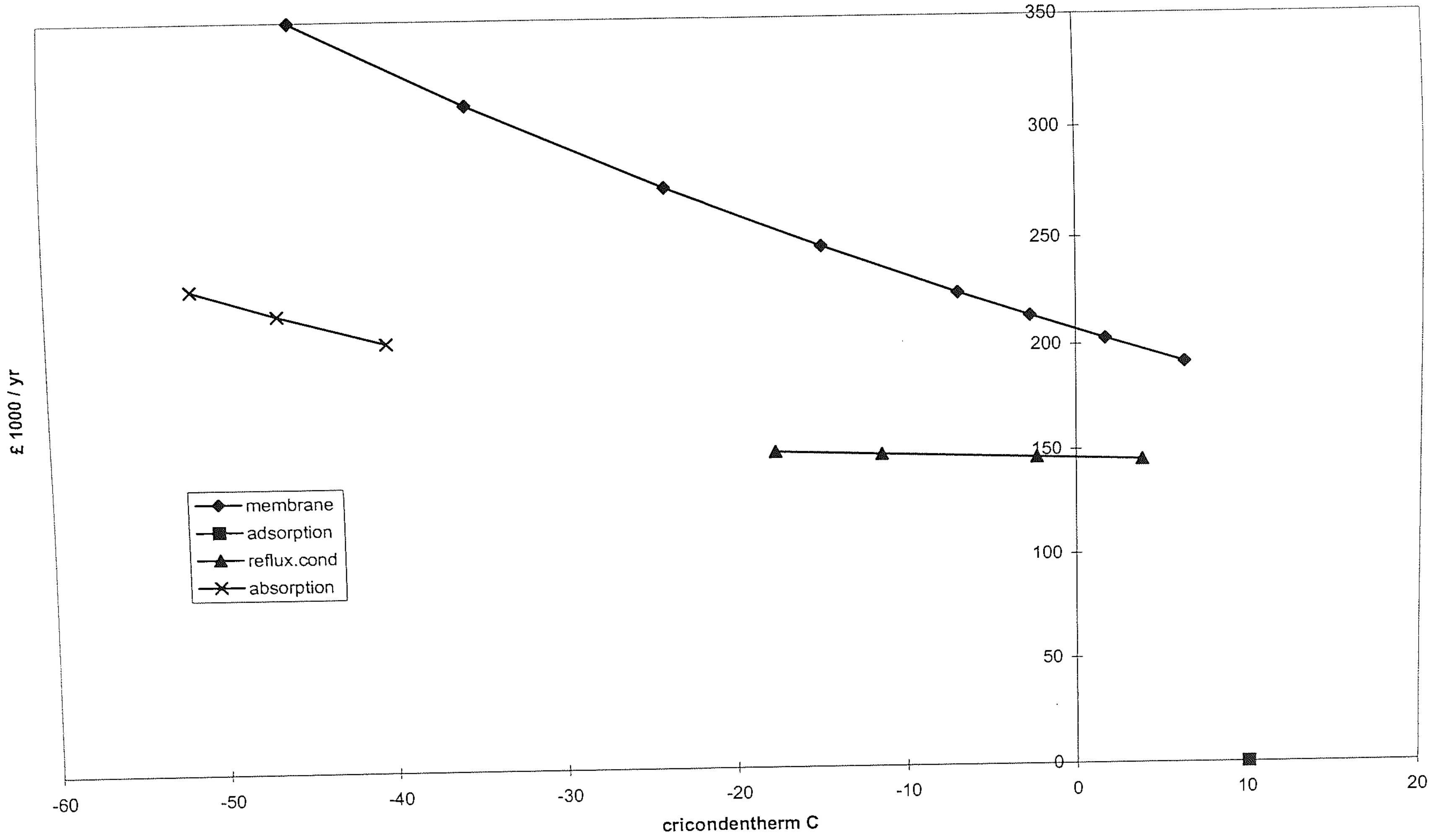


Figure 52, Case 3: Value of the gas

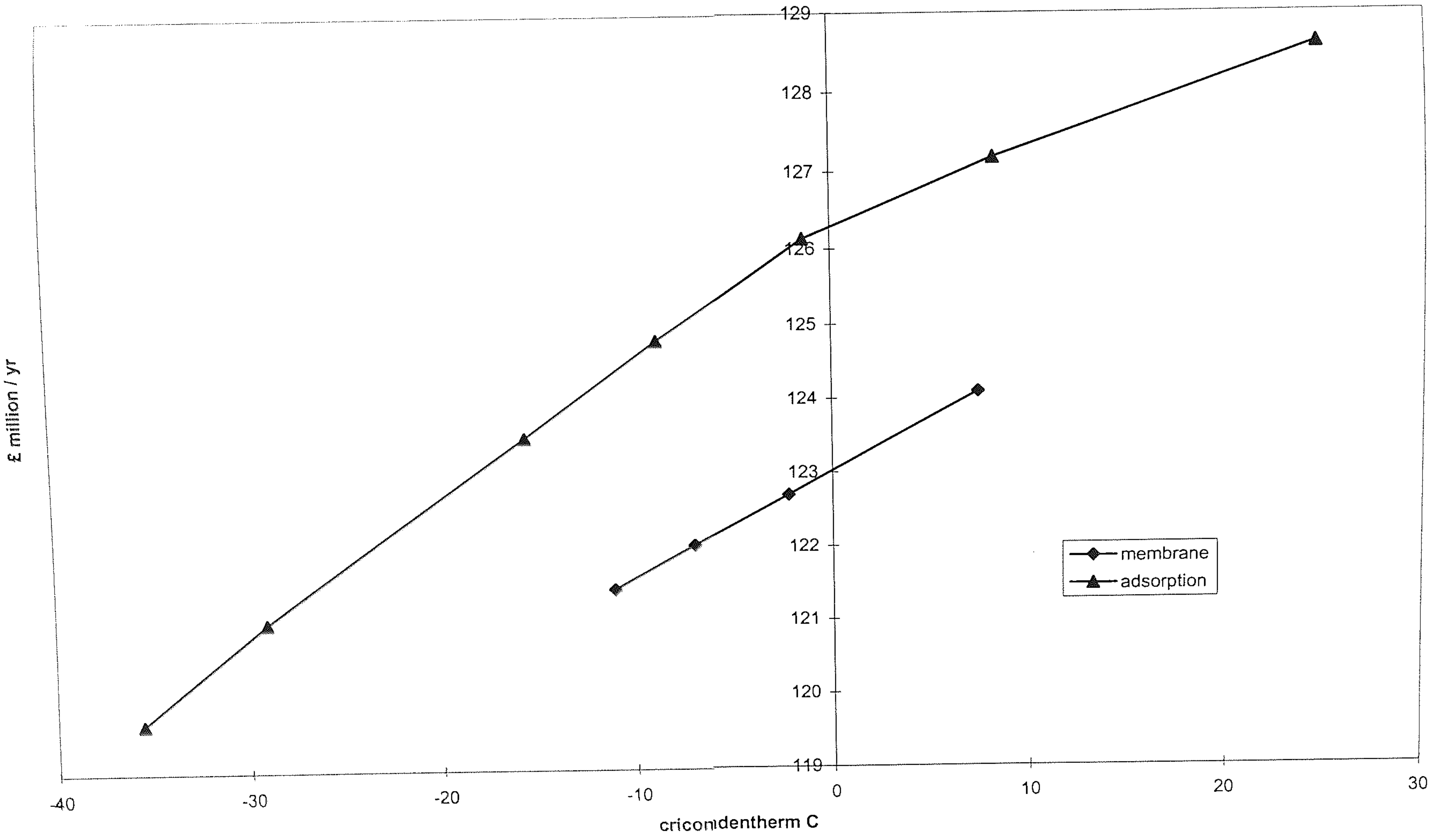


Figure 53, Case 3: Value of the liquid condensate

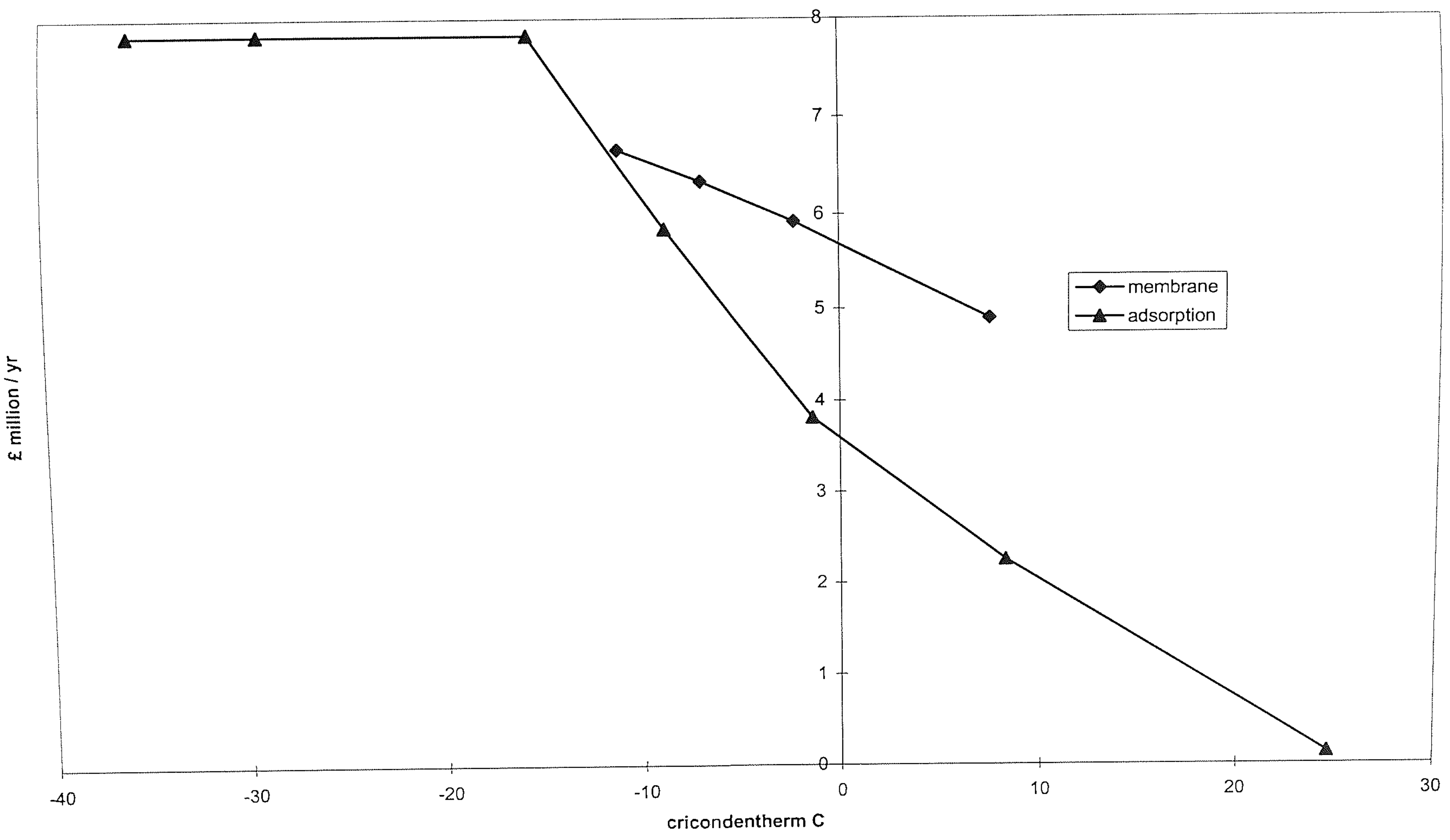
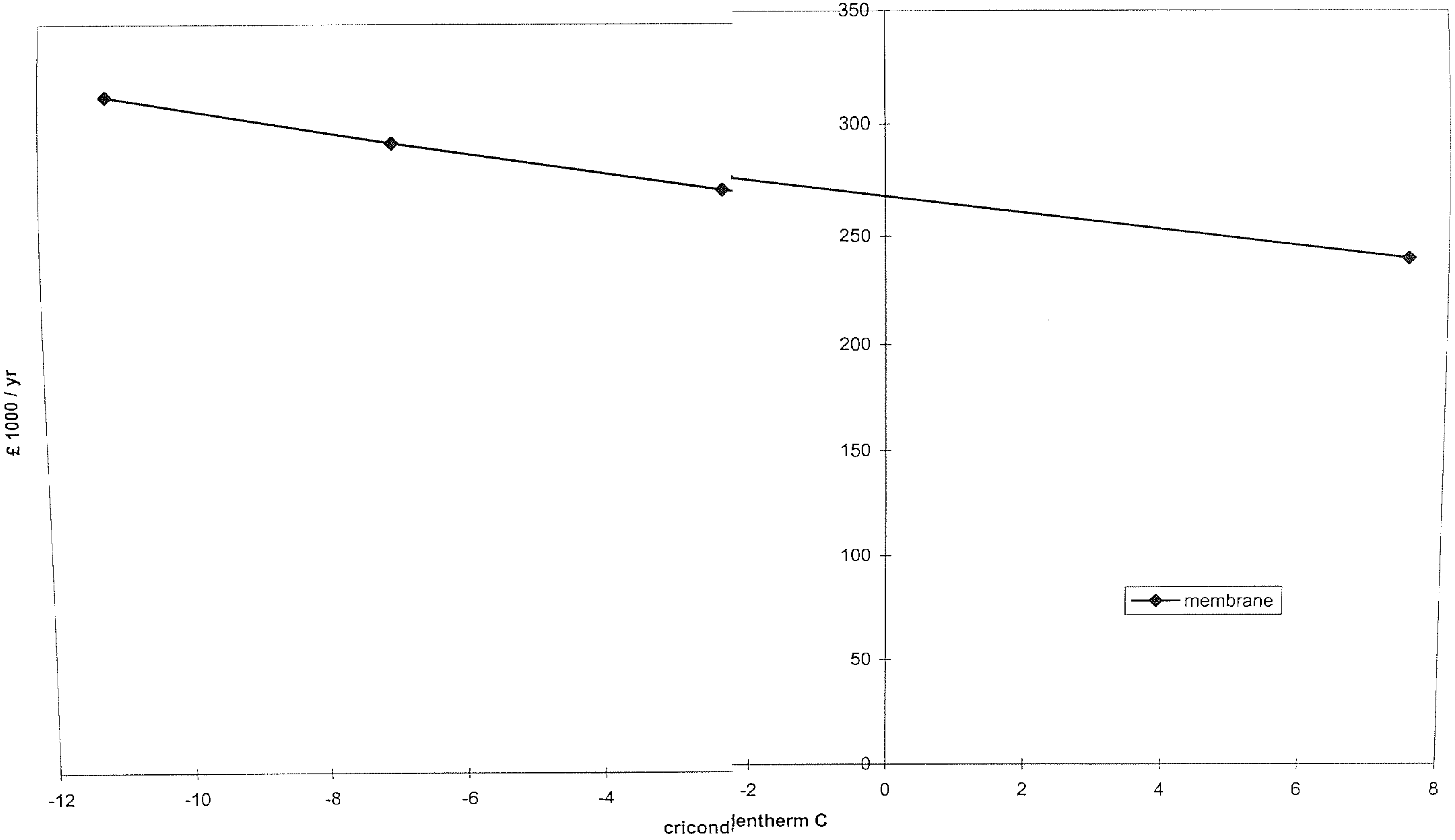


Figure 54, Case 3: Cost of condensate stabilisation



Each of the processes will now be evaluated quantitatively which will be illustrated on the charts and their performance compared with those of the equilibrium processes.

7.2 QUANTITATIVE METHOD OF PROCESS EVALUATION AND COMPARISON

All of the non-equilibrium methods may be represented on the cost and value diagrams as shown in figure 45 on page 154. The comparative cost analysis is carried out by selecting a base-case method, from the processes to be analysed. The total product value of a given process is then subtracted from the total product value, TPV, of the base-case process. If the TPV of the process is less than the TPV of the base-case it is taken as a penalty cost, whereas any gain with respect to the base-case is taken as extra value. To this are added the amortised reprocessing cost and the amortised capital cost of the separator.

The value of the products and the costs of reprocessing is obtained from the value and cost curves, figures 45 to 54. The amortised separator capital costs are calculated using cost equations as in chapter 5.

Since all the non-equilibrium processes maintain the product pressure at the feed pressure the cost of recompression does not have the same effect that it has for pressure reducing processes. The difference between the non-equilibrium processes will therefore be more dependent on the efficiency of the process separating the heavy hydrocarbons from the gas stream.

Case 1: The value of the products and the reprocessing costs are obtained from Figures 45 to 48. In this case irrespective of the process method, the produced gas will require recompression. This is due to the feed pressure at 35.5 bar being lower than the required transport pipeline pressure of 84 bar. The cost of recompression is therefore significant for all the processes as shown in Figure 48 on page 157.

In order to evaluate the non-equilibrium processes and compare them with the equilibrium processes the same base-case has to be used. Of the alternative processes available refluxing condenser is selected as the base-case for comparative cost analysis.

The analysis shows that apart from adsorption the other non-equilibrium processes have a lower total product value than the base-case, Table 16 (page 168). This is taken as a penalty cost. When all the costs are considered the most competitive non-equilibrium processes are absorption and membrane with respective to the base-case.

Note that two different membrane processes are considered here, in the first case the feed is at a pressure of 35.5 bar. In the second case the feed has been compressed to 85 bar pressure, 1 bar above the specified product gas pressure of 84. This would be done if a membrane process was selected since the membrane separation is dependent on the pressure drop and pressure ratio across the membrane. The high pressure not only reduces the cost of reprocessing, because the separation is better, but it also reduces the area of the membrane. The comparative cost of the membrane processing decreases by more than £ 2.92 million, so that it is only 14 percent more expensive than a refluxing condenser process.

The analysis shows (Table 16, on page 168) that the best equilibrium process, refrigeration is 9 percent more expensive than a refluxing condenser process. Of the two processes, that using a refluxing condenser has the higher capital cost but a much lower reprocessing cost. This is because the refluxing condenser process is not only more selective but it is also more efficient in cooling the gas than refrigeration. At this, the conceptual design phase of the study, the best process for case 1 is the refluxing condenser process.

Case 2 : For this we use figures 49 to 51. The aim here is to process the feed so that it meets the sales gas specifications of -2°C cricondentherm with an outlet pressure of 70 bar. All the non-equilibrium processes can meet the product specifications. For the comparative cost analysis the adsorption process is taken as the base-case.

The results presented in Table 17 on page 168, show that there is no reprocessing cost for adsorption because only nC_8+ components need to be removed to meet the specified cricondentherm temperature of -2°C, which means no stabilisation is needed to meet the Reid vapour pressure specification of 14.5 psi. This advantage combined with a low capital cost makes adsorption the cheapest process. The refluxing condenser process though having the highest TPV is 32 percent more expensive because its capital cost is 50 percent higher than the adsorption process.

If the membrane process is used it will be almost twice as expensive as adsorption. This is mainly due to a higher reprocessing cost and the added penalty incurred as a result of the loss of TPV with respect to the base-case method. The absorption process, which is not suited to this type of feed and product specification since only a very small percentage of the components needs to be removed, will cost more than six times as much.

In order to determine the best processes for case 2, the equilibrium and non-equilibrium processes are compared in Table 17 on page 168. The results show that the adsorption process is the best option with the equilibrium processes, refrigeration and turbo-expander, costing 15 and 22 percent more respectively than adsorption to meet the same product specification. The advantage gained by the equilibrium processes in obtaining higher TPV is lost due to their capital cost being more than 50 percent higher than the adsorption method.

Case 3 : Figures 52 to 54. Apart from adsorption, none of the processes can meet the product specification of -22°C cricondentherm temperature. This is because the feed pressure at 118 bar is much higher than the product pressure of 70 bar. Unlike adsorption and membrane the feed pressure has to be reduced to below 80 bar for absorption and refluxing condenser before processing. This will result in a hybrid combination of the two processes. They are therefore evaluated in the next chapter.

The comparative cost analysis shows (Table 18, on page 169) that adsorption with no reprocessing cost obtains the highest Total Product Value. However, it also has the highest separator capital cost due to a heavy processing requirement. The separator cost is higher than the corresponding additional value of the products compared to the equilibrium processes. Consequently, it costs more than four times as much as the turbo-expander to achieve the same product specifications. Thus the best process for case 3 is the turbo-expander.

The non-equilibrium processes tend to have an advantage over the equilibrium processes when the feed gas is lean and the required product pressure is close to or higher than the feed pressure.

The membrane process is not able to meet the product specification of -22° C cricondentherm temperature.

In the next chapter this methodology will be applied to hybrid processes.

Table 16: case 1-comparative cost analysis.

Process	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	TPV_{bc} - TPV	Comparative Separation Cost	% difference
Turbo-expander	131.32	4290.77	578.80	14.32	4883.89	34.36
Refrigeration	130.94	3281.54	629.89	52.54	3963.97	9.06
Vortex tube	131.22	7530.00	287.89	24.30	7842.19	115.75
Membrane (1)	125.91	3349.37	3167.89	554.91	7072.16	94.57
Membrane (2)	128.72	3232.73	637.04	274.76	4144.54	14.02
Adsorption	131.58	2921.81	2779.92	-11.31	5690.42	56.55
Reflux condenser	131.46	2972.45	662.36	0.00	3634.81	0.00
Absorption	128.93	3123.07	1330.38	253.71	4707.16	29.50

Table 17: case 2-comparative cost analysis.

Process	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	TPV_{bc} - TPV	Comparative Separation Cost	% difference
Joule-Thomson	118.60	680.49	270.63	-184.22	766.90	120.34
Turbo-expander	118.62	49.30	578.84	-200.63	427.51	22.83
Refrigeration	118.62	49.35	551.37	-198.30	402.42	15.62
Vortex tube	118.63	680.69	35.56	-212.53	503.72	44.72
Membrane	118.32	213.98	364.57	102.35	680.90	95.63
Adsorption	118.42	0.00	348.06	0.00	348.06	0.00
Reflux condenser	118.48	48.87	479.16	-65.67	462.36	32.84
Absorption	117.39	203.90	1366.34	1032.42	2602.66	647.76

Table 18: case 3-comparative cost analysis.

Process	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	TPV_{bc} - TPV	Comparative Separation Cost	% difference
Joule-Thomson	125.08	872.21	270.63	324.66	1467.50	100.84
Turbo-expander	125.41	398.98	331.72	0.00	730.70	0.00
Vortex tube	125.09	751.00	264.78	318.49	1334.27	82.60
Adsorption	128.66	0.00	6941.63	-3247.27	3694.36	405.59

7.3 INTERIM OBSERVATIONS

1. The methodology developed here can be used to identify the most competitive options available to the process engineer at the beginning of a project.

2. If the processing requirement is small, the more selective process methods have a clear advantage over the less-selective separators.

3. A process which is more selective has an advantage over other processes because it maintains the heating value of the gas, thereby maintaining its value, while recovering enough high quality condensate, which does not require much reprocessing.

4. Non-equilibrium processes have an advantage over the equilibrium processes because they eliminate or minimise the need for recompressing the product gas.

5. This methodology can be applied to other feed gases of varying composition, pressure, and temperature, to choose the best process.

6. The evaluation of the different processes can be seen clearly by presenting the data graphically.

7. Using this methodology it is possible to demonstrate how the cost of processing will vary if the inlet feed conditions are altered, as demonstrated for the membrane process in case 1.

8. If there are two or more processes with equal value or cost then they need to be evaluated over the project life to select the best.

CHAPTER 8

8.0 HYBRID PROCESSES - AN EVALUATION

A methodology which enables the economic evaluation of existing processes for natural gas conditioning, can this be used to suggest and evaluate novel processes? In particular, does it enable the rapid selection and evaluation of hybrid schemes inco-operating two or more processes of different types? Further, do hybrid processes offer higher value and/or do they have any additional advantage over the individual processes?

Hybrid processes as defined here are combinations of two or more different process technologies. It seems likely that combination of two or more process methods may be economical. However, to-date there is very little information published on hybrids in the literature. The aim in this chapter is to suggest ways of selecting a hybrid process to condition natural gas feeds. The processes developed may than be evaluated and compared, using the methodology developed in chapters 6 and 7, against the single process routes so far considered to determine whether they are economically viable.

It is clear that selecting a hybrid process is a highly complex problem, as there are many alternative combinations of the processes as tabulated below. As with the individual equilibrium and non-equilibrium processes, further processing of the product streams is usually required. The gas product may need to be recompressed and the liquid product may need to be stabilized to remove any high vapour pressure products.

A further complication in this case is determining how much separation should be carried out in the first separation module and how much in the second separation module. The order of the processes in a hybrid combination also needs to be determined so that the processes may be operated such that their ability to separate heavy hydrocarbons from the feed gas is maximized.

Note, any two or more of the processes presented below can be combined to form a hybrid. However, not all the hybrid combinations will yield an economically viable process. In order to select a viable process it is therefore helpful to understand the individual process characteristics. Some of the separation characteristics of the individual processes applied to hydrocarbon dew-point control are shown in Table 19, below.

Any hybrid processes may be represented in the form of a generalised flow chart as in Figure 55.

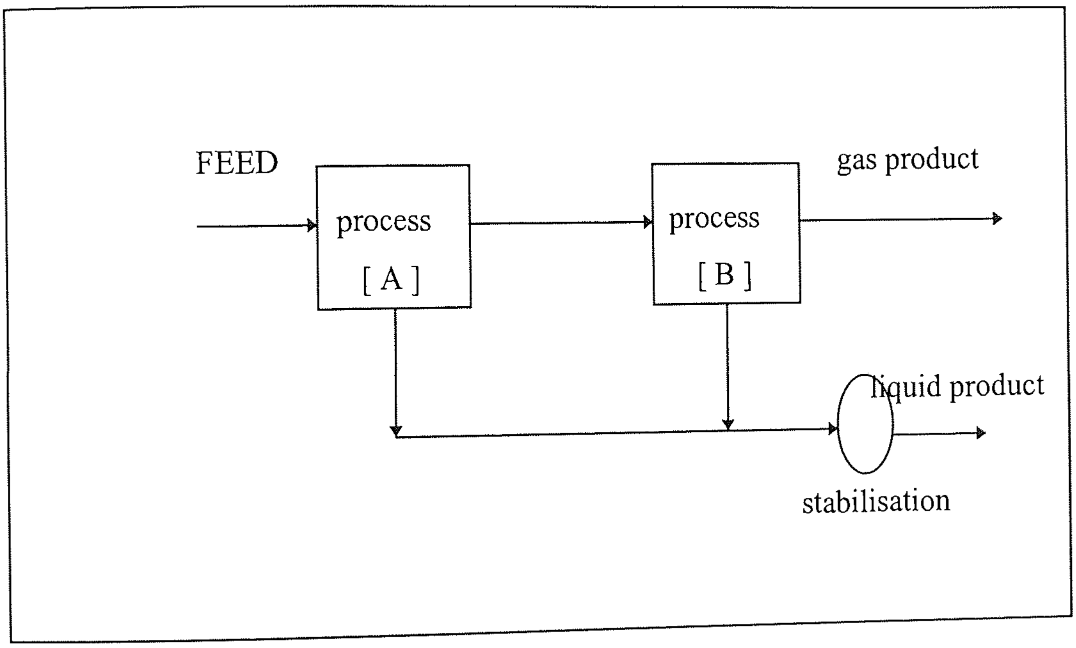


Figure 55 : Typical generalised flow chart for a hybrid process.

In developing the methodology for choosing the best possible hybrid process combinations, the separation processes may be classified into two main categories, see Table 19. These are 'pressure drop processes' and 'constant pressure processes'. They may be further classified as crude, medium, and highly selective separation. A crude separation is the one that will have a high percentage of light hydrocarbons in the condensate stream, whereas the highly selective process will have relatively low percentage of light hydrocarbon gases in the condensate stream.

The pressure drop processes produce cooling as a result of pressure drop, thereby condensing the heavy components from the feed gas. This is a simple but crude method of separating the heavier molecules from the gas stream. Note that turbo-expander is generally more efficient than both the Joule-Thomson valve and the vortex tube at separating the heavy hydrocarbons from lights. The constant pressure processes, however, use various alternative methods to achieve separation, for further details see chapter 2. Note that the constant pressure processes, adsorption and refluxing condenser are highly selective processes, whereas the membrane process are classified as a crude separation since a high percentage of light hydrocarbons is present in the condensate stream.

Table 19 : Process classification.

Process	Classification	Separation selectivity	Capital Cost	Complexity
Membrane	constant pressure	crude	low	low
Absorption	constant pressure	medium	high	high
Adsorption	constant pressure	high	low	high
Joule-Thomson	pressure drop	crude	low	low
Vortex tube	pressure drop	crude	low	low
Refluxing cond.	constant pressure	high	medium	medium
Refrigeration	constant pressure	medium	medium	medium
Turbo-expander	pressure drop	medium	medium	low

When combining processes it is important for technical and economic reasons that the chosen processes complement each other. For example, if limited 'free' pressure drop is available then there is a strong case of using a pressure drop process with a constant pressure process, rather than another pressure drop process.

Thus, the selection of the hybrid process combination may be simplified by taking into account the following guide lines:

1) Generally 'good' to combine:

a) A pressure drop followed by a constant pressure process, preferably using medium or highly selective separation process. This will minimise the reprocessing costs, by keeping the condensate stabilisation costs low.

b) A crude separation process followed by a highly selective process. This will mean the combined process is not too complex and will keep the capital cost to a minimum.

2) Avoid the combination of:

a) A pressure drop process with another pressure drop process. This could result in higher recompression and stabilization costs.

b) A crude separation method with another crude separation method. This will lead to high stabilization costs.

c) Two highly selective processes. This may unnecessarily make the process highly complex, leading to higher capital and operating costs.

8.1 POSSIBLE HYBRID PROCESS COMBINATIONS

Following the guide-lines above, some of the possible hybrid processes are considered below.

8.1.1 VORTEX TUBE - REFRIGERATION HYBRID

The flow scheme in Figure 56 on page 176 may be used to represent the hybrid process that uses a vortex tube and refrigeration. The feed gas from the plant inlet facilities enter the gas/gas exchanger, where the cold product gas is used to cool down the feed gas. The cooled gas is then expanded through the vortex tube. The expanded gas then flows to the gas separator, where the condensed heavy hydrocarbons are separated from the gas. The gas then passes to the gas chiller, where further cooling takes place. The refrigerant unit is used to obtain a low chiller temperature. The cooled gas flows to the cold gas separator, where the condensed liquids are separated from the product gas. The gas is than used to cool the incoming feed in the gas/gas exchanger.

These two processes complement each other in that, if free pressure drop is available, the vortex tube will take advantage of it. Any further cooling of the product gas may then be carried out by the refrigeration process. This will avoid recompression of the product gas.

8.1.2 JOULE-THOMSON - ADSORPTION HYBRID

The feed gas from the plant inlet facilities is expanded through the Joule-Thomson valve. The expanded gas then flows to the cold gas separator, where the condensed heavy hydrocarbons are separated from the gas. The gas then passes through the silica gel adsorber beds, where any remaining heavy hydrocarbons are removed. The produced condensate from the process units flows to the condensate stabilisation column, (Figure 57, on page 177).

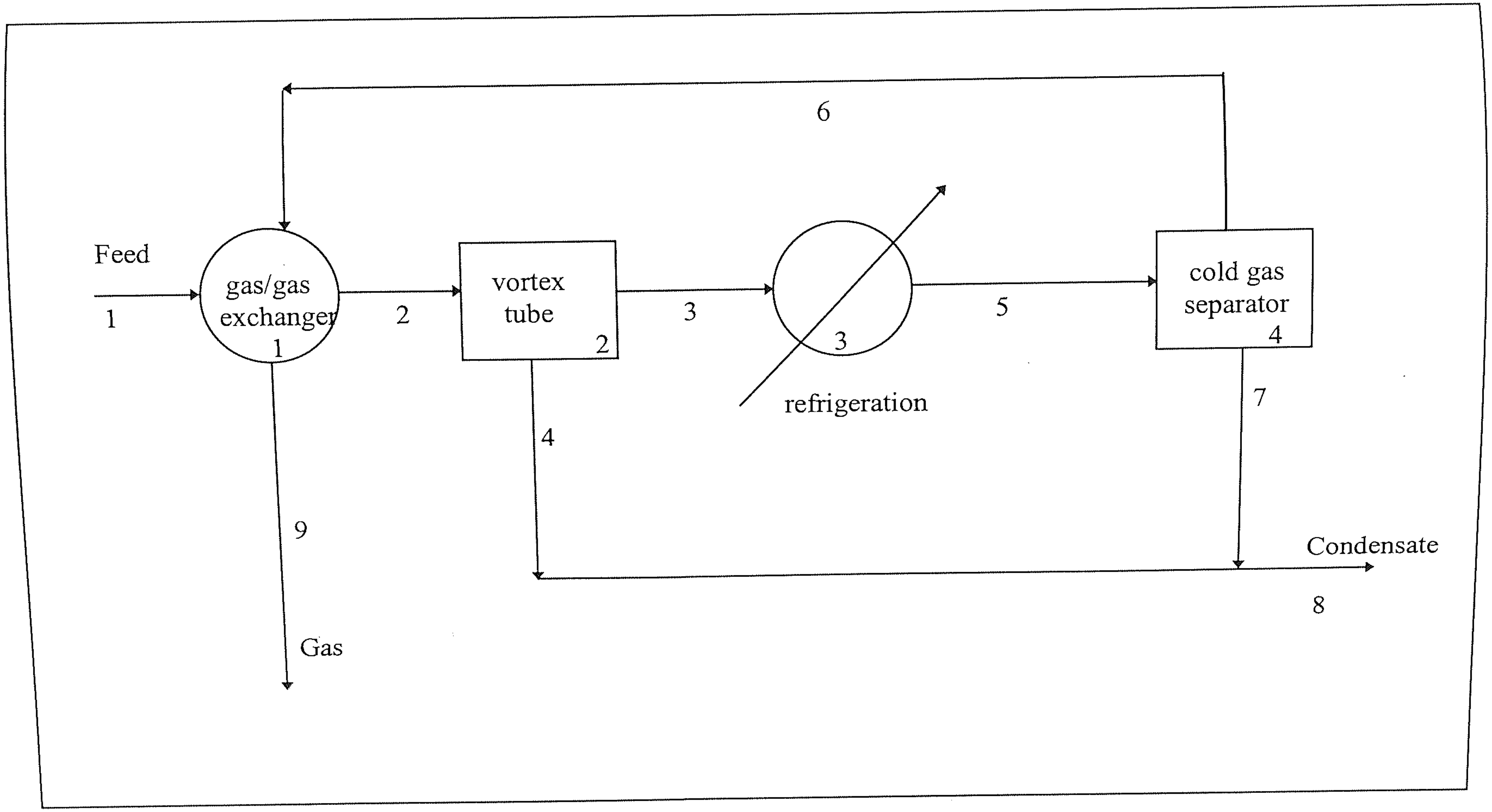


Figure 56: Vortex tube - Refrigeration hybrid.

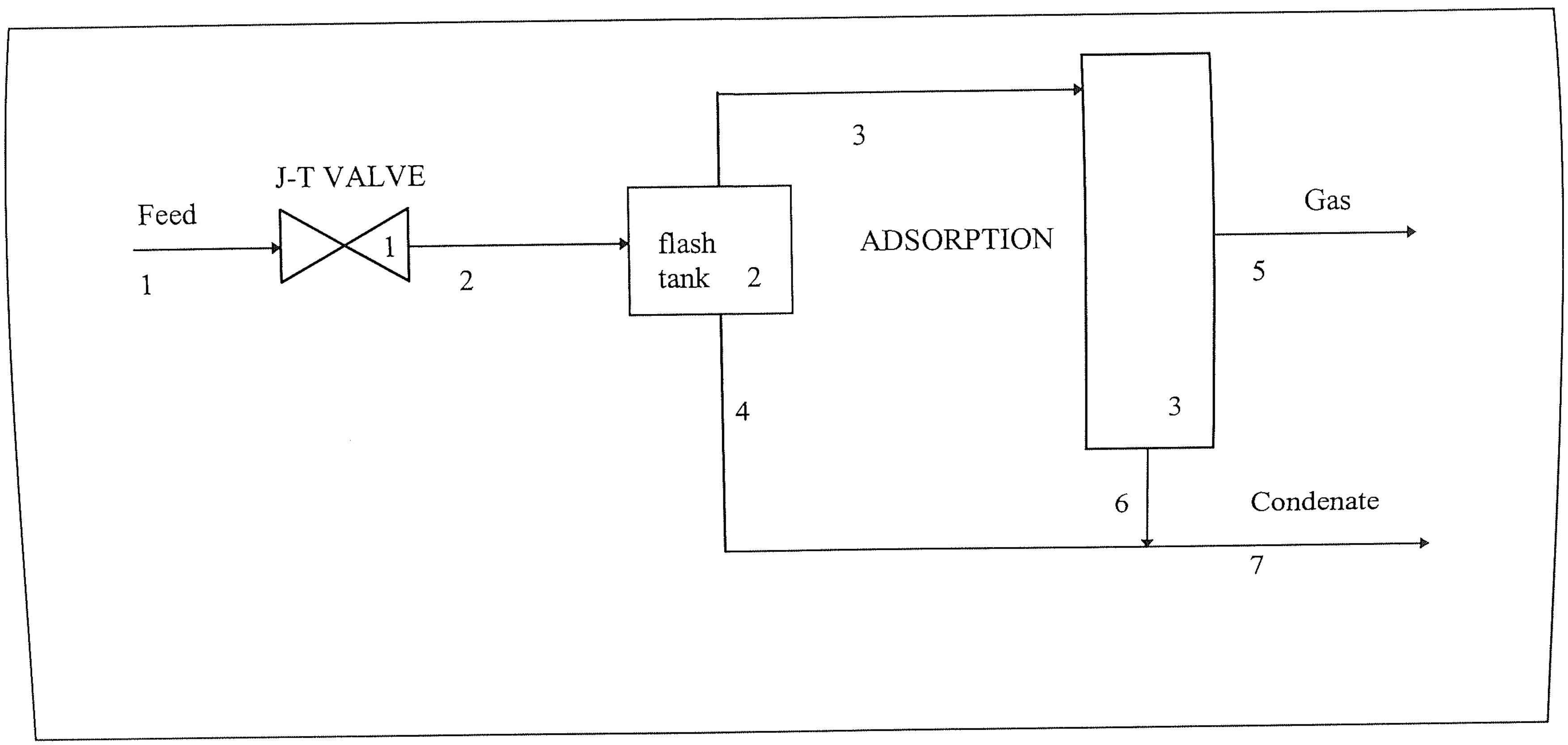


Figure 57: Flow-sheet for Joule-Thomson - Adsorption hybrid.

8.1.3 MEMBRANE - REFLUXING CONDENSER HYBRID

The feed gas flows through the filter coalescer and then through the membrane units, as shown in the flow diagram of Figure 58 on page 179. The permeate (condensate) stream, which contains a high percentage of light hydrocarbon gases, then flows to the refluxing condenser, where the heavier components are liquefied efficiently. This combination may reduce the reprocessing cost, which are high due to the crude nature of separation and therefore make the membrane process comparatively cheaper. The other option is to use adsorption to remove the heavier components of the permeate stream, see the next hybrid.

8.1.4 MEMBRANE - ADSORPTION HYBRID

Unlike the case above the product gas from the membrane module flows to the adsorption columns, where further separation of the heavy hydrocarbons from the gas stream takes place, (Figure 59, on page 180). Adsorption is probably the best combination with a membrane module since it can selectively remove heavy components from the gas stream without the need to reduce pressure or temperature. Therefore the condensate produced by the adsorber may not require any further stabilisation. As mentioned previously both membrane and adsorption processes are able to process gas outside the two-phase region of the phase envelope.

8.1.5 TURBO-EXPANDER - REFLUXING CONDENSER HYBRID

The feed gas from the plant inlet facilities enters the gas/gas exchanger, where the cold product gas is used to cool down the feed gas. The feed gas is than expanded through the turbo-expander. The expanded gas flows to the gas separator, where the condensed liquids are separated. The produced gas then flows to the refluxing condenser unit to be cooled further. The cooled gas flows to the gas separator, where the condensed liquids are again separated from the product gas. This gas is then used to cool the incoming

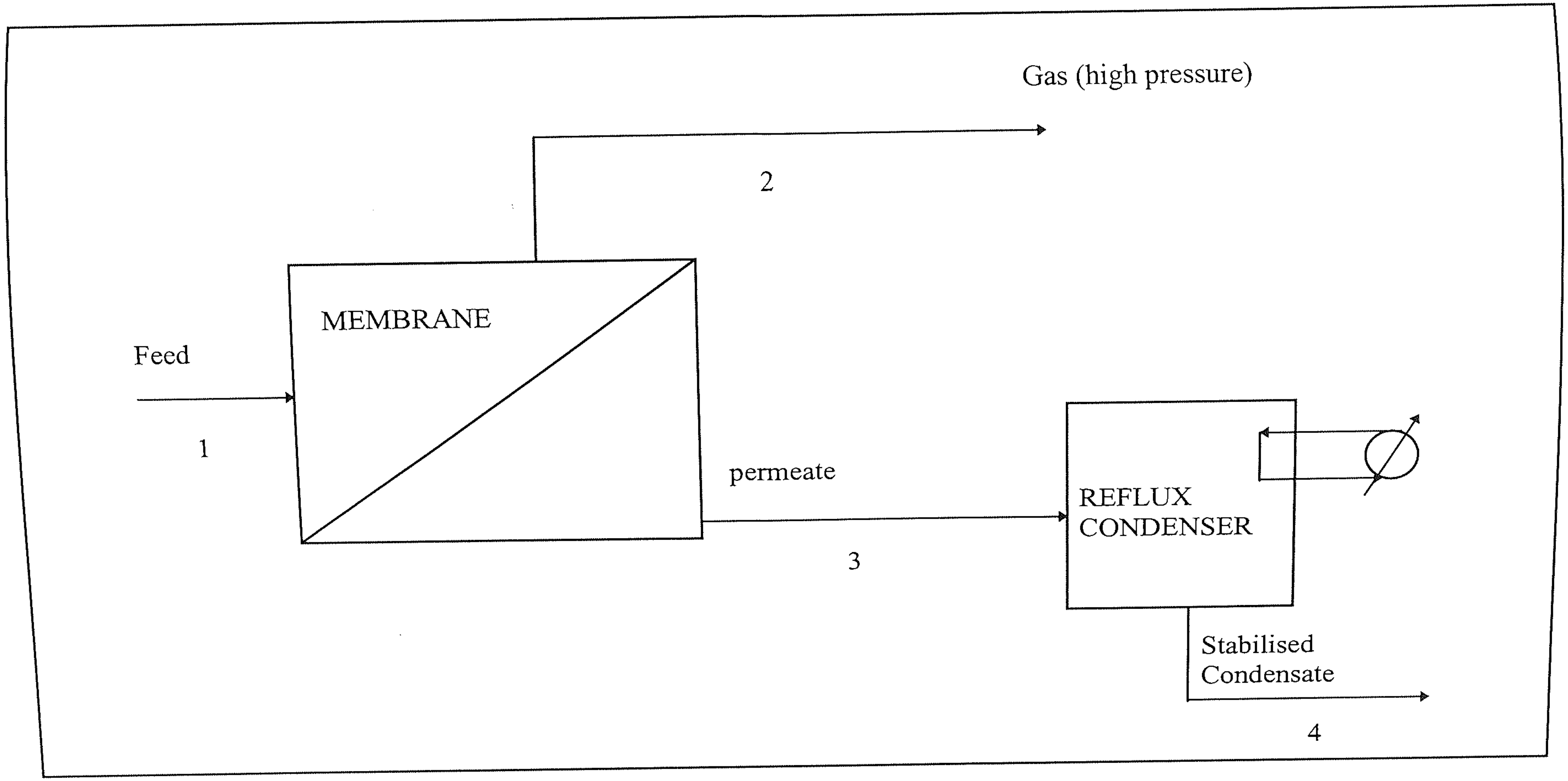


Figure 58: Flow-sheet for Membrane - Reflux condenser hybrid.

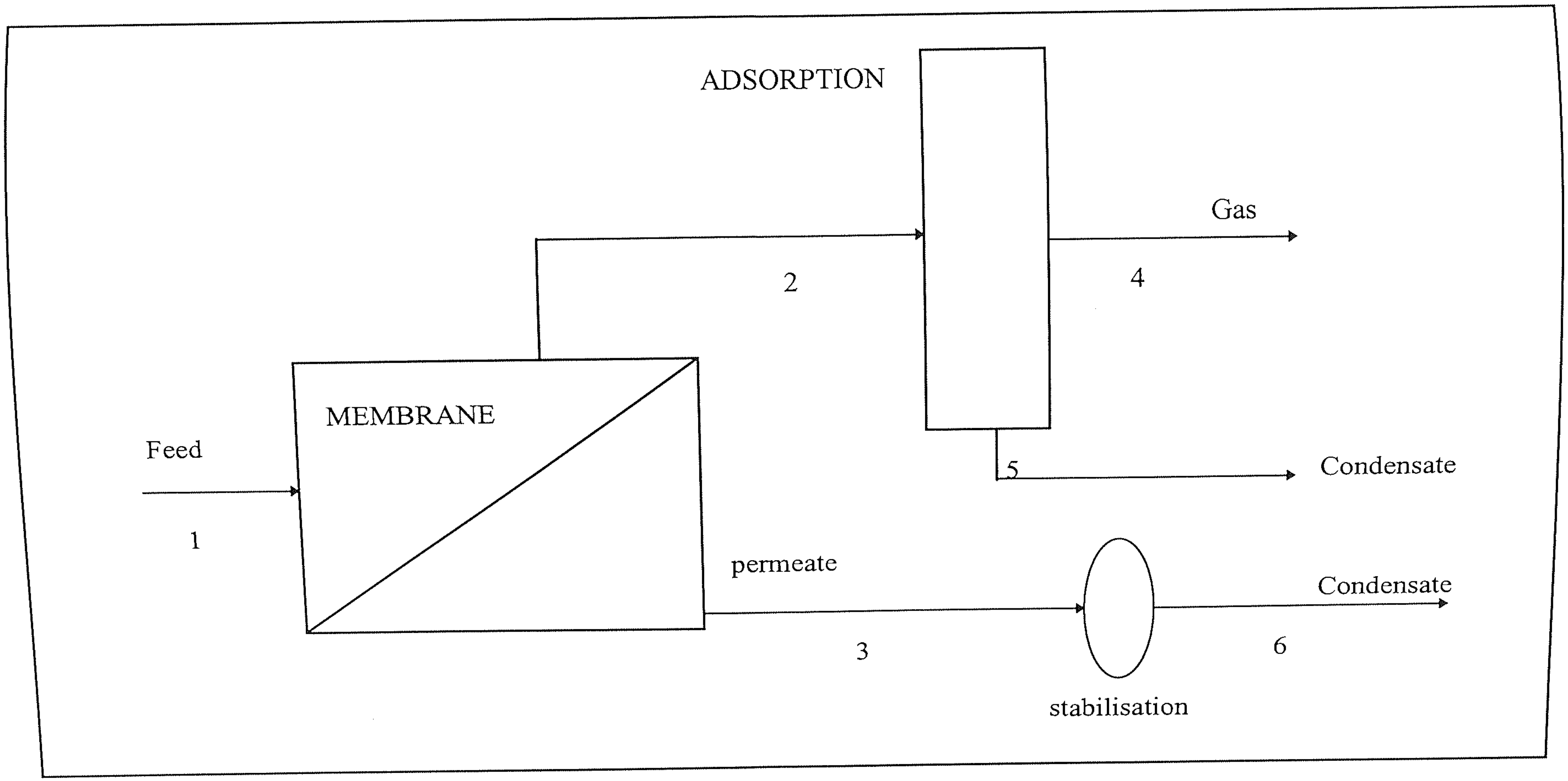


Figure 59: Flow-sheet for Membrane - Adsorption hybrid.

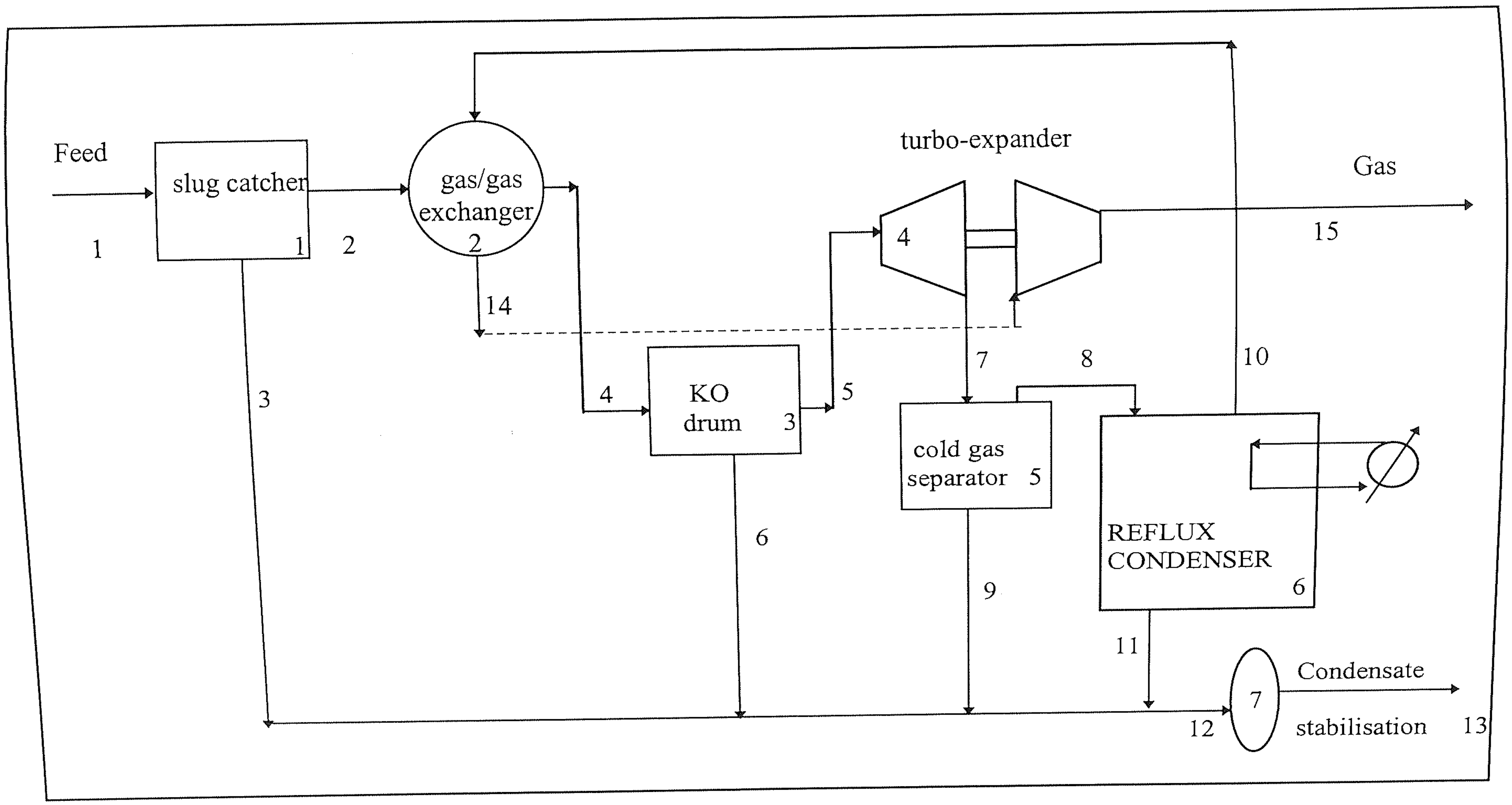


Figure 60 : Flow-sheet for Turbo-expander - Refluxing condenser hybrid.

feed in the gas/gas exchanger. The product gas then flows through the compressor section of the turbo-expander, to be compressed, (Figure 60, on page 181).

This combination may be necessary, if heavy hydrocarbon dew point duty is specified.

8.2 WHEN TO CONSIDER THE USE OF A HYBRID PROCESS

There are no published rules or guide lines as to when the hybrid processes can or cannot be considered for hydrocarbon dew pointing. Hybrid processes may be evaluated for any feed or product specification. As to whether they will offer any additional advantage cannot be judged in advance. The following points may be used as a guide, when considering whether to evaluate the use of hybrid processes or not. They are by no means exhaustive.

1) If only a limited 'free' pressure is available, but not enough for the pressure drop processes to meet the specified dew point, without having to recompress the product gas.

2) The capital cost of the individual process is high (it may need to include a compressor), and the overall comparative cost of the hybrid is likely to be lower.

3) The hybrid process may add additional value to the product by maintaining higher calorific value of the product gas or by recovering a higher percentage of heavy hydrocarbons in the condensate.

4) When as a result of the shape of the phase envelope, the quality lines and the inlet feed conditions, the pressure has to be dropped, before the constant pressure processes such as refrigeration can process the gas to meet the specified cricondentherm.

8.3 SELECTING THE OPTIMIUM DEGREE OF SEPARATION

The next step is to determine how much separation is to be carried out in the first separation module and how much in the second separation module of the hybrid.

This is because different separation combinations will produce the same dew-point specification but they are likely to produce hybrids with different total costs. It is therefore necessary to determine the optimum separation split, which will give a hybrid process with the lowest cost.

The individual processing units of the hybrid are evaluated separately, using the methodology illustrated in chapters 6 and 7. The results of the evaluation may then be plotted on cost charts, as in Figures 61 and 62, below. Note that the degree of separation corresponds to the cricondentherm temperature obtained by the process.

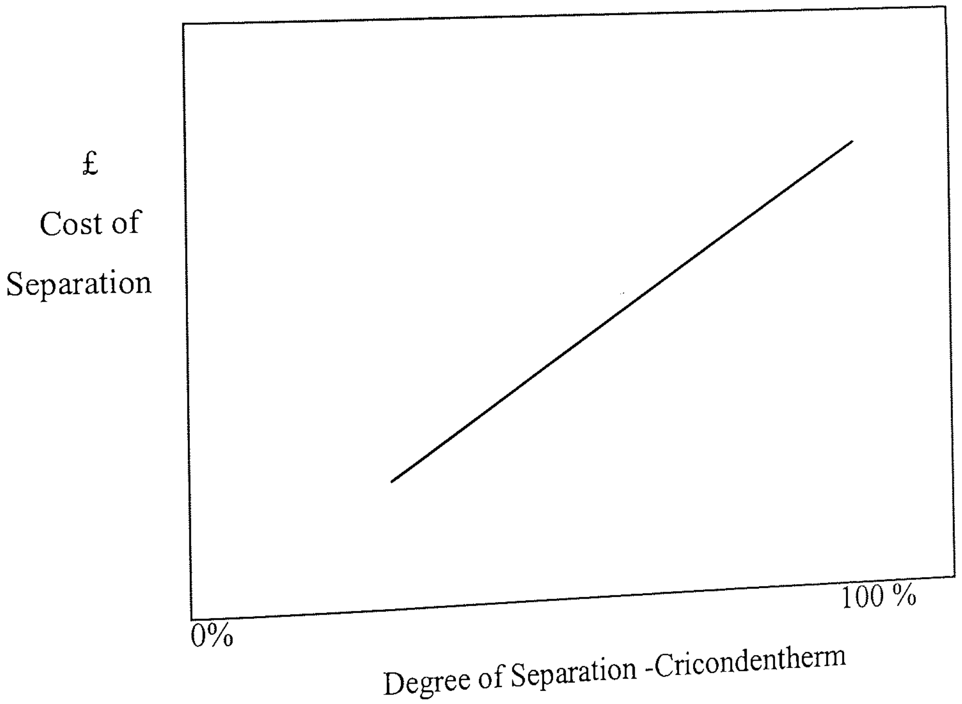


Figure 61 : cost of using process 'a'

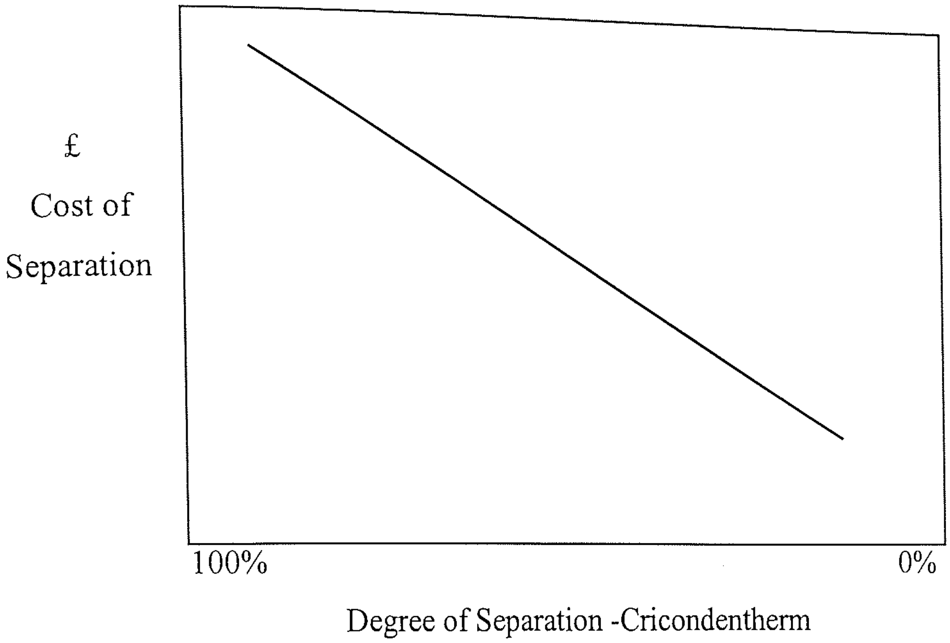


Figure 62 : cost of using process 'b'

When the cost of the process 'a' and 'b' are combined to give a cumulative cost of the hybrid process, the following 'u' shaped cost curve may be obtained. The minimum point represent the optimum separation split between the two separation processes. However, if there is no minimum, then the suggested hybrid combination is not economical.

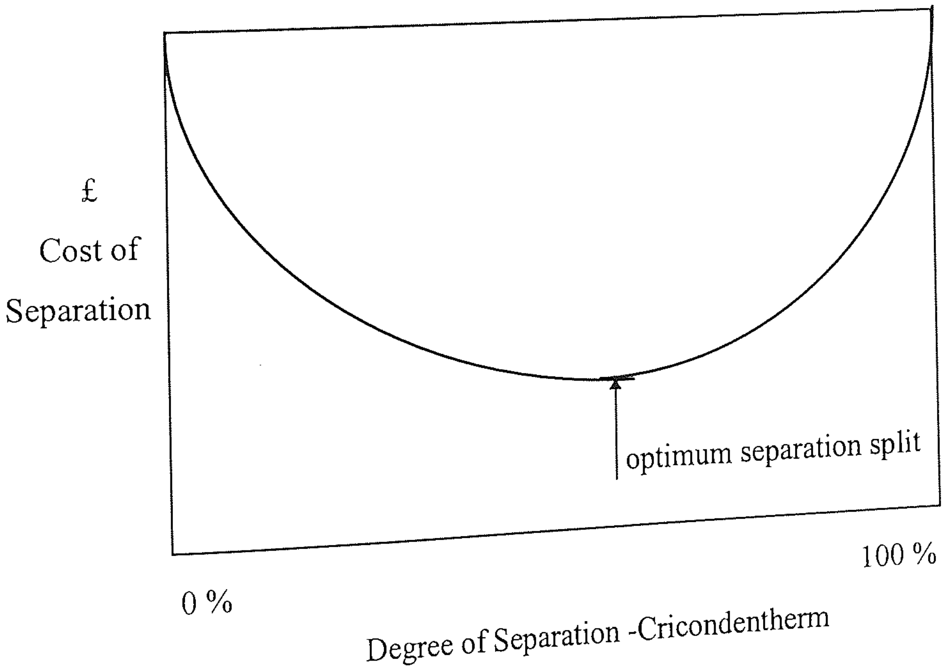


Figure 63 : evaluation of the hybrid process

Note that the x-axis represents the cricondentherm temperature of the product gas leaving the vortex tube section.

The same procedure is used, if three or more processes are combined to form an hybrid.

The advantages of this approach will now be explained by a quantitative evaluation of the hybrid process considered.

8.4 QUANTITATIVE METHOD OF PROCESS EVALUATION

For the purpose of demonstration the same feed gas composition and product specifications as for case 2, in both chapters 6 and 7, will be used here. The vortex tube-adsorption hybrid combination to be evaluated is as shown in Figure 64.

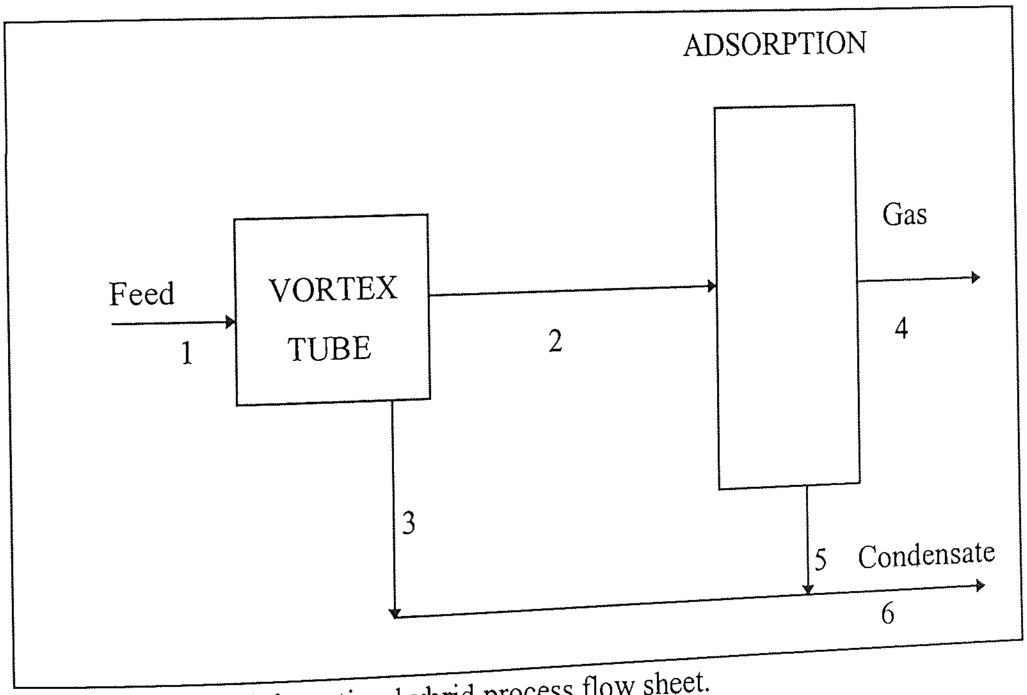
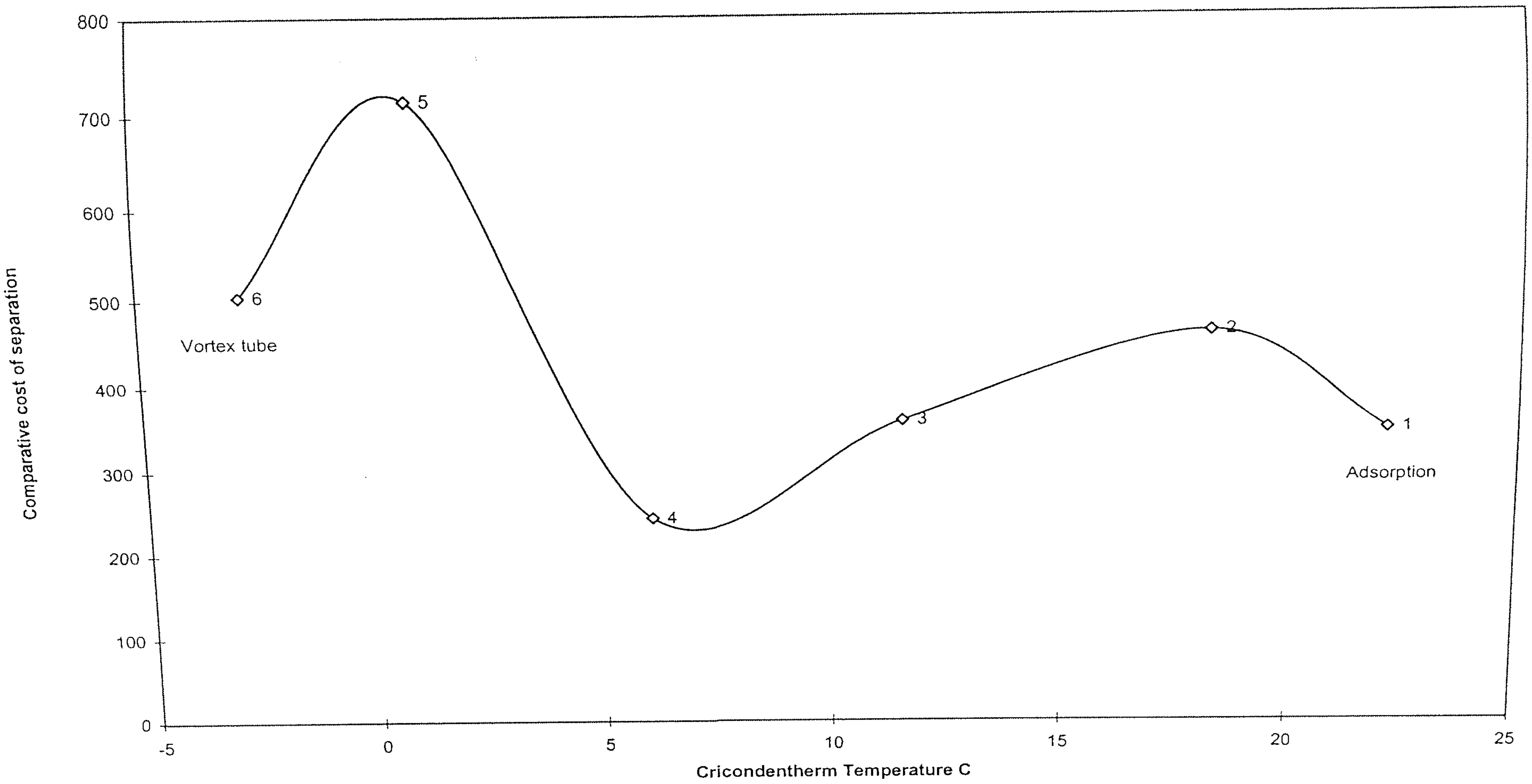


Figure 64 : Vortex tube - Adsorption hybrid process flow sheet.

The results of the analysis presented in Table 20 on page 188 and illustrated in Figure 65 on page 186, show that the hybrid process reduces the cost of processing, when a product gas cricondentherm temperature of 6.1°C is obtained after the first section of

Figure 65: Evaluation of Vortex tube-Adsorption hybrid



the hybrid process (vortex tube) before the gas enters the adsorption section for further processing to meet the specified cricondentherm of -2°C.

On Figure 65, point 1 represents the feed gas cricondentherm, here 100 % of the duty is performed by the adsorption process. The cost is the cost of the adsorption process, conversely point 6 represents 100 % of the duty being performed by the vortex tube process.

For the comparative cost analysis, the cheapest process for case 2, adsorption, is used as the base-case process (as in chapter 7). The new hybrid process is more than 29% cheaper, see Table 20 on page 188. This is mainly due to the lower capital cost of the hybrid process. This is because the vortex process reduces the amount of the heavy hydrocarbons to be removed by the adsorber. This in turn reduces the size of the adsorption process and therefore the capital cost.

The added advantage of using a vortex tube is that it produces two separate product streams, a liquid condensate and a product gas stream. This means that a high pressure flash tank is not required. By limiting the pressure drop to 10 bar in the vortex tube also eliminates the need to recompress the product gas. The highly selective adsorption process only needs to remove NC_8+ components to meet the sales gas cricondentherm temperature of - 2°C. This therefore minimises the condensate stabilisation cost.

Thus the best (cheapest) process for processing feed composition 2 to the product specifications is the vortex tube-adsorption hybrid.

Table 20: Comparative cost analysis of Vortex tube-Adsorption hybrid.

% Separation	Cricondentherm achieved in Vortex tube	Total Product Value £million/y	Amortised Reprocessing cost £1,000	Amortised Capital cost £1,000	Comparative Separation Cost	% difference
100 % Vortex tube	-2.75	118.63	680.69	35.56	503.72	44.72
	0.98	118.55	577.80	263.86	714.78	105.36
	6.1	118.52	48.62	293.46	245.37	-29.50
	11.61	118.44	48.30	332.8	359.53	3.30
	18.28	118.38	48.10	374.31	462.42	32.86
100 % Adsorption		118.42	0	348.06	348.06	0.00

8.5 INTERIM OBSERVATIONS

1. The guide lines for hybrid process selection may be applied to processes other than those considered here.

2. Using this methodology it can be determined whether the selected hybrid will cost more or less than the individual component processes.

3. The optimum separation split needs to be determined to obtain the best (cheapest) hybrid process.

4. The same procedure could be applied to evaluate a hybrid process with three or more individual processing units.

5. These rules can be used to target the combination of two processes that are most likely to be commercially viable. This will save time and effort at the conceptual design phase.

6. The hybrid process can offer higher value than individual processes, as shown above for feed composition 2.

7. The vortex tube-adsorption hybrid process is comparatively 29 %, cheaper than all the processes, including vortex tube and adsorption.

CHAPTER 9

9.0 DISCUSSION

In this work the aim has been to develop a methodology for process selection, which can be used at the conceptual design phase by a process engineer or a researcher to evaluate and select the best/cheapest process method for gas conditioning. The methodology has to be able to cope with variations in any given feed condition such as composition, temperature, pressure and output product specifications and to be applicable to any traditional or novel process method that exist or may be invented in future. Therefore, to ensure coverage of the majority of the processing options available, the following set of technologies have been considered :

1. Joule-Thomson expansion,
2. Turbo-expansion,
3. Vortex tube,
4. Refrigeration,
5. Membranes,
6. Absorption,
7. Adsorption, and
8. Refluxing condenser.

In this list are included some traditional processes that are commonly used, such as refrigeration and turbo-expanders and some novel technologies such as refluxing condensers and vortex tubes, whose application to gas conditioning have only recently been considered. To date, there is only limited information available, if any, on the economic benefits of these technologies for gas conditioning.

The problem was complicated further, by the number of different process technologies available and because each process technology uses a different separation technique to process natural gas. In addition, some of these processes are restricted to within the two phase region of the phase envelope, whereas others are not. For evaluation purposes, all these processes have to be modelled and costed realistically so that actual processes are correctly represented.

In addition, we need to identify innovative schemes which combine these separate technologies and quantify the potential benefits of using them as hybrids.

To illustrate the general applicability of the evaluation methodology three different cases have been considered in detail. These cases were suggested by BP Exploration. Each case is different in terms of feed composition, pressure, temperature and the required product specifications.

The process models used to evaluate the three cases have been presented in chapter 4. Only the major equipment components of the respective processes were included in the flowsheets. The process parameters and the resultant product outputs are what one would expect from the corresponding real process. These complex processes were successfully modelled using a design simulation package with an acceptable degree of accuracy.

The Joule-Thomson valve, turbo-expander, refrigeration and absorption processes have been around for a long time and as such the respective simulation models are commercially available. Whereas adsorption, membrane, refluxing condenser and vortex tube are relatively new and novel technologies and as such, no commercial models of these processes exist. To overcome this, the BP's in-house proprietary designs were adopted and modelled using GENESIS.

Each process must be evaluated on an equal footing. This is a problem because each process has its own complexities and technical characteristics. This was solved by transforming the product specifications, such as dew point, calorific value and Wobbe number into a single cricondentherm temperature. The same cricondentherm temperature is obtained with many different combinations of the product gas compositions. Thus the competing process technologies, each with its own complexities and technical characteristics, can be compared.

Data for economic evaluation were obtained from vendors for the corresponding model process flow schemes. The inquiries were based on the three example cases and were made under the auspices of BP Exploration Facilities Engineering. The cost data received were correlated with the model design or sizing parameter or both for the corresponding processes. The resultant cost equations can be used to estimate the capital cost of any process, only if cost parameters are obtained by using the above mentioned process models. Additional limitations on these cost equations are that they should be used for costing hydrocarbon dew-point plants only. For example, in the case of the turbo-expander plant the cost is based on feed flow rate and moderate pressure drop. For a NGL recovery plant with the same feed flow rate, the pressure drop would be higher, this would escalate the cost of plant. Furthermore, if the correlations are used outside the specified ranges then doin so will lead to significant errors.

The selection methodology developed is based on comparative separation cost. It has been defined as :

$$\text{Comparative Separation Cost} = [\,TPV_{bc} - TPV\,] + AR + AC$$

The comparative separation cost calculation takes into account the total value of the products (TPV), separator capital cost (AC), and any reprocessing cost (AR), due to the recompression of the product gas and/or condensate stabilisation.

By including the value of the products, reprocessing cost as well as the capital cost of the process for comparing, means that any advantage one process has over another due to the way it separates the heavy hydrocarbons from the feed gas is taken into account and vice versa. An added advantage of this methodology, is that it recognises that different processes will produce products that meet the same product specification, but will have different values. For example, in Case One all the processes meet the specified product cricondentherm temperature of 10°C. However, data tabulated in Table 16 on page 168, shows that there are clear differences between the total value of the products. Here, the value varies from £131.58 million for adsorption to £125.91 for membrane. Similar differences exist when considering the reprocessing and capital costs. The effect of reprocessing cost varies depending on the type of process. For the equilibrium processes they can be significant, if the product gas requires recompression to meet the pressure specification, see Tables 16, 17, and 18 on pages 168-169.

The equilibrium processes (methods 1 to 4, in the list above) are limited to the two phase region of the phase envelope. Any point in the two phase region represents the composition and quantities of the gas and liquid in equilibrium. Therefore, for any given quantity of feed and product specifications, the value of the products, the cost of recompression and condensate stabilisation can be calculated. These are then presented as cost and value lines. By using these all the equilibrium processes are compared simultaneously. This is achieved by drawing the operating lines for each process method on the diagram, see figure 32 on page 134. This simplifies and speeds up the evaluation procedure.

The non-equilibrium processes (methods 5 to 8), on the other hand are evaluated independently for the given quantity of feed and product specifications, in each case. The cost and value calculations are carried out at the specified product cricondentherm temperature.

The non-equilibrium processes plus refrigeration have an advantage over the other equilibrium processes because they maintain product pressure and thus avoid compression. This is illustrated in Case Two, Table 17 (on page 168), where Joule-Thomson valve and vortex tube, though with high product values and low capital costs are two of the most expensive processes due to the cost of compression. The non-equilibrium process also have an advantage, when the feed gas composition is relatively lean and little or no free pressure drop is available, as in Case Two. The more selective a process is at separating the heavier hydrocarbons from the gas stream the more competitive it is likely to be. This is because, it is able to produce a high value gas, by maintaining its heating value, which dominates the profit/cost calculation. Of all the processes evaluated, adsorption is the most selective and consequently, is the best process for Case Two, with the lowest overall comparative cost. The methodology also shows that if free pressure drop is available as in Case Three, then the other process methods cannot compete with the pressure drop processes.

Furthermore, the methodology shows that the non-equilibrium processes have an advantage over the equilibrium processes because they minimise the reprocessing cost. For example, Case One, is a rich gas feed at low pressure and high temperature. The comparative cost analysis of the equilibrium processes shows the refrigeration process to be the best. However, when it is compared with the non-equilibrium processes, it is second best to the refluxing condenser. The cost difference between the two process methods is due to the lower reprocessing cost of the refluxing condenser and higher value of its products. This is due to the efficiency of the separation performed in the refluxing exchanger, the greater efficiencies reducing the condensate stabilisation load, and so decreasing the reprocessing cost.

The analysis of the three case studies clearly demonstrates that the value of the products as well as the reprocessing and separator capital cost must all be taken into consideration to select the best process. Moreover, it illustrates why one process is better than another, since we may ask, Does it produce products with higher value? Does it have lower reprocessing costs or Does it have a low separator cost?

It is important to remember that this methodology has been developed to be used at the conceptual design phase of a project, to screen processes. By screening processes, all the possible options are examined, to select the best process method(s) to be considered in more detail in the next phase of the project. The chosen process method(s) can be said to be the best/cheapest with a high degree of confidence, because the methodology evaluates all the processes on an equal footing and clearly illustrates why the best method is the best or the cheapest.

This selection procedure is generic in essence and thus applicable to traditional and any novel process technology. If adopted, it will reduce the time taken to screen processes. However, the results are dependent on the quality of the modelling and the economic data used. Hence, if more accurate models and cost data are available, they can be used instead of those presented here.

This selection methodology can also be extended to the evaluation and comparison of the hybrid processes. As an illustration using the guide lines developed, a hybrid process was evaluated for example Case Two. The hybrid is a combination of vortex tube-adsorption processes. The comparative separation cost analysis (Table 20, on page 188), shows that this process is more than 29% cheaper than any equilibrium and non-equilibrium process evaluated for Case Two. The methodology shows that this hybrid process is cheaper because it is able to utilize the small amount of free pressure drop available in the vortex tube and the high selectivity offered by the adsorption process.

The identification and evaluation of this innovative process combination clearly shows that there are advantages in using hybrid processes. They can offer higher value and lower processing cost over the individual processes. Moreover, the same procedure can be used to evaluate any number of process combinations, for any feed/production conditions. The application of hybrid processes will increase, since they have the potential of reducing the total capital cost as well as increasing separation efficiency.

The limitations of this thesis arise as expected from the modelling and cost estimation sections. When modelling processes, assumption were made to take into account the limitations of the process models. For example, the membrane model was limited to only ten components in the feed. Since there were more than 20 components in each of the cases, some of the heavier components present in minute quantities were lumped together. This could and will have affected the shape of the phase envelope and therefore the cricondentherm temperature.

Also greater accuracy could be achieved for the refluxing condenser, if reliable cost data were available. However, since no vendor was willing to provide any cost data due to its proprietary nature, the cost is based on the plate-fin heat exchanger and adjusted for the reflux exchanger.

The selection method can be improved by computerising it. This will reduce the time taken at the evaluation stage.

CHAPTER 10

10.0 CONCLUSIONS

10.1 MAJOR CONCLUSIONS

1. The methodology presented here can be used to identify the most competitive equilibrium and non-equilibrium process options available to the process engineer at the beginning of the project.

2. The methodology can be used to illustrate why one process has an advantage over another.

3. The selection methodology is not biased toward any process, it can be used to evaluate all the traditional and novel processes on an equal footing for any given feed gas and/or product specifications.

4. Using this methodology it can be determined whether the selected hybrid will cost more or less than the individual component processes.

5. The guide lines can be used to target the combination of two or more processes (hybrid) that are most likely to be commercially viable. This will save time and effort at the conceptual design phase of the project.

6. The hybrid processes do reduce the cost of processing and/or increase the value of the products as shown in example case two, chapter 8.

7. An innovative process, vortex tube-adsorption hybrid is 29% cheaper than all other processes, including vortex tube and adsorption.

8. All the processes can be pin pointed on the value and cost diagrams as operating lines very easily.

9. Using the selection methodology developed, it is possible to determine the optimum process operating point for the required product specifications.

10. The methodology takes into account the loss in total product value as a result of the efficiency of the processing method as well as the added costs.

11. Using this methodology it is possible to demonstrate, how the cost of processing will behave if the inlet feed conditions are altered, as in Case One, for the membrane process.

12. If the processing requirement is small, the more selective processing methods have a clear advantage over the coarse separators.

13. A process which is more selective has an advantage over other processes because it maintains the heating value of gas, thereby maintaining its value, whilst recovering enough high quality condensate (which may not require substantial reprocessing).

14. Non-equilibrium processes have an advantage over the equilibrium processes because they either eliminate or minimise the need for recompressing the product gas.

15. If there are two or more processes with equal value or cost then they need to be evaluated over the project life to select the best.

10.2 MINOR CONCLUSIONS

1. The process models presented in chapter 4 can from a process specification, produce estimates of equipment size and power requirements.

2. These models provide a suitable basis for correlation of the cost data obtained from the sources available.

3. From the cost information supplied by the vendors, cost correlations have been obtained in terms of cost parameters. The parameters used to describe the systems are:

Process	Parameters
Joule-Thomson	pressure and feed flowrate
Turbo-expander	pressure and feed flowrate
Refrigeration	pressure and heat duty
Membrane	pressure, area and power
Adsorption	pressure, feed flowrate and condensate recovery
Vortex tube	pressure and feed flowrate
Absorption	pressure, feed flowrate, heat duty and liquid flowrate.

4. The economic analysis carried out allows comparisons between different process systems to be made in terms of the overall process.

5. The correlations obtained can be used to estimate the capital costs of process technologies for a range of specifications. However, any extrapolation of the cost data beyond the specified range can introduce significant error into the cost estimates.

CHAPTER 11

11.0 SUGGESTIONS FOR FUTURE WORK

1. Further work is required to computerise the selection method. This will reduce the time taken at the evaluation stage.

2. Developments of more accurate models for refluxing condenser, membrane, vortex tube and adsorption processes.

3. Obtaining more accurate vendor cost data for refluxing condenser and adsorption for hydrocarbon dew pointing.

4. Further work needs to be done on the potential for technological stretch when using these technologies singly or in hybrid combination

REFERENCES

1. De Kraa, J.D., Thompson, S.M., "Advances in natural gas processing technology : part 2", 13th world petrol congr (Buenos Aires, Argent), 20-25/10/91),3 , 409-415, (20-25/10/91).

2. HORIZON, BP Exploration magazine, 10, (10/1993).

3. Lemin, N., "The chemical engineer", The great gas dash, (22/1996).

4. Tissot, B., "Natural Gas : growing technical and scientific requirements", Institute Francais du Petrole, 25-26, (1992).

5. International Standard ISO 6976, "Natural Gas - Calculation of calorific value, density and relative density".

6. Wrobel,W., Wright, P., "Calorific values and relative densities", The Institution of gas engineers, communication 1080.

7. BP Engineering, Chemical engineering course, 1, 12-14, (1991).

8. Campbell, M.J., Gas Conditioning and Processing, Campbell Petroleum Series, 1, 4, 79-91 (1984).

9. Nishida, N., Stephanopoulos, G., Weterberg, A.W., "A review of Process Synthesis", AIChE Journal, 27, 3, 321-348, (1981).

10. Barnicki, S.D., Fair, J.R., "Separation System synthesis: A knowledge-based approach. 1. Liquid mixture separations", Ind. Eng. Chem. Res., 29, 3, 421-432, (1990).

11. Barnicki, S.D., Fair, J.R., "Separation System synthesis: A knowledge-based approach. 2. Gas / Vapour mixture separations", Ind. Eng. Chem. Res., 31, 7, 1679-1694, (1992).

12. Rudd, D.F., "The synthesis of System Designs: I. Elementry Decomposition Theory", AIChE Journal, 14, 2, 343-351, (1968).

13. Masso, A.H., Rudd, D.F., "The synthesis of System Designs: II. Heuristic Structuring", AIChE Journal, 15, 1, 10-17, (1969).

14. Pagnard, A., Paradowski, H., "NGL recovery: Select the Right Process", Ist oil and gas Australia petrol. technol. Australia conf., (Perth, Western Australia), (26-28/11/1985).

15. Crum, F.S., "Use J-T plants for LPG recovery", Hydrocarbon process, 60, 5, 113-117, (5/81).

16. Crum, F.S., "There is a place for J-T plants in Lpg recovery", Oil and Gas Journal, 79, 32, 132 - 138, (8/1981).

17. Millar, J. H., "Production enhancement from topside changes at an offshore gas cycling project", 67th annual SPE tech conf., Washington, DC, 397 - 405, (10/1992).

18. Gas Processors Association, Engineering data book, 1, 13.40-13.46 and 14.1-14.2, (1987).

19. Barnwell, J., Wong, W., "Expanders do payouts offshore north sea", 5th SPE et al Offshore south east asia conf., Singapore, 3-66 - 3-76, (2/1984).

20. Moruzzi, L., Righi, E., "Recover energy in natural gas by turbo-expansion", Pipeline industry, 71, 4, 28-31 (10/89).

21. Erbar, J.H., Maddox, R.N., "Solving gas-processing problems. PT. 8. Expansion processes, turboexpander efficiency vital for predicting liquid-recovery levels", Oil and Gas, 79, 27, 130-134 (7/81).

22. Technology report by P&GJ Staff, "The recovered energy advantage", Pipeline and Gas Journal, 34 - 36, (3/1989).

23. Morgan, M.J., Cousino, M.R., "Major Caribbean gas processing plant starts up", Oil and Gas Journal, 89, 43, 42 - 46, (10/1991).

24. Jimenez-Gomez, H., "Enchanced prpane recovery at the eastern Venezuela cryogenic complex", 67th Annual gas processors ass conv., Dallas, 89 - 92, (3/1988).

25. Agahi, R, Ershaghi, B., "Turbo-expander redesign concepts and economics", 71st Annual GPA conv., California, 29 - 33, (3/1992).

26. Campbell, M.J., Gas Conditioning and Processing, Campbell Petroleum Series, 2, 17, 281-289 and 273-277 (1984).

27. BP Engineering, Chemical engineering course, 3, 57-58, (1991).

28. Fekete, K.A., "Preventing vaporization of the liquid in a centrifugal gas-liquid separator", United States Patent, 4, 458, 494, (10/7/1984).

29. Fekete, K.A., "Vortex tube separator may solve weight/space limitation", World oil, 203, 1, 40-44, (7/1986).

30. United States Patent, 3, 775, 988.

31. United States Patent, 3, 546, 891.

32. Campbell, M.J., Gas Conditioning and Processing, Campbell Petroleum Series, 2, 16, 239 - 252, (1984).

33. Campbell, M.J., Gas Conditioning and Processing, Campbell Petroleum Series, 2, 17, 305 - 315, (1984).

34. Berini, D.L., Widjojo, T., "Enhanced NGL recovery in the Arun field expander plant using a low-pressure absober for maximizing propane recovery", 65th Annual SPE tech conf., New Orleans, 329 - 334, (9/1990).

35. Cooley, T.E., Dewhirst, M., "New developements in treating natural gas with membranes", Presented at Gas Processors Association meeting, European chapter, (1/1993).

36. Stern, S.A., Walawender, Jr., W.P., "Analysis of membrane separation parameters", Separation Science, 4, 2, 129 - 159, (4/1969).

37. Saltonstall, C.W., "Calculation of the membrane area required for gas separations", Journal of Membrane science, 32, 185 - 193, (1987).

38. Heinemann, S.B., "Beharra springs - fast track development of a small gas field utilising membrane separation technology", Australian Petrol. Explor. Ass. Conf. (Gold Coast, Australia), APEA, 33, 1, 352-362 (1993).

39. Cheriyan, G.K., "Membrane Technology for Natural Gas Dehydration", 4th EC Symposium, Berlin, 121-133 (1992).

40. Laverty, B.W., O'Hair, J.H., "Applications of the membrane technology in the gas industry", Fourth BOC priestley Conference, 291 - 310.

41. Koros, W.J., "Membranes: learning a lesson from nature", Chemical Engineering Progress, 68 - 81, (10/1995).

42. Campbell, M.J., Gas Conditioning and Processing, Campbell Petroleum Series, 2, 19, 341-370 (1984).

43. Keller II, G.E., "Asorption: Building upon a solid foundation", Chemical Engineering Progress, 56 - 67, (10/1995).

44. Parsons, J.P., Templeman, J.J., "Model's performance leads to adsorption-unit modifications", Oil and Gas, 88, 26, 40-44 (6/1990).

45. Coker, A.K., "Program sizes solid desicant dryer for natural gas", Oil and Gas Journal, 74 - 78, (2/1994).

46. Limb, D.I., Czarnecki, B.A., "The petroflux process for NGL recovery experience to date and new developments", 66th GPA Annual convention, 105 - 114.

47. Chiu, C.H., "Advances in gas separation", Hydrocarbon processing, 69, 1, 69 - 72, (1/1990).

48. Finn, A.J., "Enhance gas processing with reflux heat exchangers", Chemical Engineering, 142 - 147, (5/1994).

49. Tomlison, T., Finn, A.J., Limb, D., "Energy analysis-a case study", The Chemical Engineer, 39 - 42, (10/1990).

50. Halford, 1., "The LPG and condensate recovery plant supplied to Hartogen energy kincora terminal", Oil and Gas Australia, 35 - 37, (6/1986).

51. Sawyer, P., "Simulation-it's the real thing!", The Chemical Engineer, 22 - 25, (14/1991).

52. Hudson, M., "Dynamic simulation as a design tool...", The Chemical Engineer, 22 - 31, (14/1991).

53. Liu, Y.A., McGee Jr, H.A., Epperly, W.R., Recent developments in chemical process and plant design, 7, 261-287, (1987).

54. Husain, A., Chemical Process Simulation, Wiley Eastern Limited, India, 2, 4 - 5, (1986).

55. Koch, D.H., "The future: Benefitting from new tools, techniques, and teaching", Chemical Engineering Progress, 66 - 72, (1/1997).

56. BP GENESIS user guide, section 6, flash facilities, release 4.0.

57. BP GENESIS user guide, Appendix A, Module Index, release 4.0.

58. Cheriyan, G.K., "Vortex tube separators-Modelling and suitability for offshore processing", (internal BP report), (4/93).

59. Campbell, M.J., Gas Conditioning and Processing, Campbell Petroleum Series, 2, 19, 385 - 386, (1984).

APPENDIX ONE

A1-1 FEED GAS COMPOSITIONS FOR CASE STUDY EVALUATIONS

Feed gas COMPOSTION 1 for CASE 1

Stream Conditions

Feed rate	200	MMSCFD
Feed Pressure	35.5	bara
Temperature	62	° C

Composition

Component	Mole Fractions
NITROGEN	0.0074
CO2	0.0524
METHANE	0.7607
NC2	0.1037
NC3	0.0423
IC4	0.0093
NC4	0.0126
IC5	0.0034
NC5	0.0026
BENZENE	0.0001
TOLUENE	0.0003
NC6	0.0023
PXYLENE	0.0002
C7-10	0.0024
C11-14	0.0002
C15-20	0.0000
C21-29	0.0000

Feed gas COMPOSTION 2 for CASE 2

Stream Conditions

Feed rate	200	MMSCFD
Feed Pressure	80	bara
Temperature	4	° C

Composition

Component	Mole Fractions
NITROGEN	0.0176
CO2	0.0040
METHANE	0.9200
NC2	0.0399
NC3	0.0086
IC4	0.0015
NC4	0.0020
IC5	0.0006
NC5	0.0006
BENZENE	0.0013
TOLUENE	0.0003
NC6	0.0012
NC7	0.0007
NC8	0.0004
NC9	0.0005
NC10	0.0005
NC12	0.0002
NC15	0.0001
NC20	0.0001

Feed gas COMPOSTION 3 for CASE 3

Stream Conditions

Feed rate	200	MMSCFD
Feed Pressure	118	bara
Temperature	0.49	° C

Composition

Component	Mole Fractions
NITROGEN	0.0082
CO2	0.0236
METHANE	0.8175
NC2	0.0890
NC3	0.0353
IC4	0.0056
NC4	0.0105
IC5	0.0030
NC5	0.0030
BP C6	0.0020
BP C7A	0.0012
BPC7B	0.0001
BP C7C	0.0006
BP C8A	0.0004
BP C8B	0.0000
BP C8C	0.0001
BP C9A	0.0001
BP C9B	0.0000
BP C9C	0.0000
BP 190A	0.0000
BP 190B	0.0000
BP 190C	0.0000

A1-2 EXAMPLE OF A TYPICAL STREAM RESULT FROM A SIMULATIOM RUN

```
SUMMARY PRINT-OUT FROM  PRT01
******************************                                  PAGE  1
                                                                *******
STREAM NO.                  1          2          3          4          5


 MOLECULAR              17.852     17.523     72.405     17.523     17.515
   WEIGHT
 DENSITY AT                                    639.2
20C (KG/CU M)
TEMPERATURE               4.00       4.00       4.00      -2.00      -2.00
  (DEG C)
  PRESSURE            80.00000   80.00000   80.00000   79.50000   79.50000
 (BAR ABS)
 HEAT FLOW           -5308.104  -4891.006   -417.098  -5763.022  -5751.996
    (KW)
LIQ DENSITY              9.731      0.000      9.731     10.530      0.000
(KMOL/CU M)
VAP DENSITY            4.47249    4.47249    0.00000    4.65015    4.65015
(KMOL/CU M)
 MOL FRACT              0.9940     1.0000     0.0000     0.9998     1.0000
 VAPORISED

TOTAL FLOW             9960.16    9900.41      59.75    9900.41    9898.73
 (KMOL/H)
COMPONENT MOLAR FLOWRATES  (KMOL/H)

     NITROGEN           175.00     174.86       0.14     174.86     174.85
     CO2                 39.84      39.62       0.22      39.62      39.62
     METHANE           9163.56    9143.66      19.90    9143.66    9143.09
     NC2                397.10     393.37       3.73     393.37     393.26
     NC3                 85.76      83.45       2.31      83.45      83.38
     IC4                 15.04      14.28       0.76      14.28      14.26
     NC4                 20.02      18.64       1.38      18.64      18.59
     IC5                  6.08       5.29       0.79       5.29       5.26
     NC5                  6.08       5.11       0.96       5.11       5.08
     BENZENE             12.55       7.61       4.94       7.61       7.45
     TOLUENE              2.99       1.11       1.88       1.11       1.05
     NC6                 11.65       8.03       3.62       8.03       7.91
     NC7                  6.67       3.16       3.52       3.16       3.04
     NC8                  4.28       1.19       3.10       1.19       1.09
     NC9                  4.88       0.67       4.21       0.67       0.56
     NC10                 4.88       0.34       4.54       0.34       0.23
     NC12                 1.79       0.03       1.77       0.03       0.01
     NC15                 1.39       0.00       1.39       0.00       0.00
     NC20                 0.60       0.00       0.60       0.00       0.00
```

A1-3 PHASE DIAGRAM CALCULATION OUTPUT

```
OUTPUT FROM MODULE NO.    4                       EFV01 EFV DATA TABULATION

 PHASE ENVELOPE                                      18 FEB 88      PAGE  3
                                                                    -------
0 STREAM NUMBER    12                       FLOWRATE    9886.72  KMOL/H
  -----------------                                   172637.94  KG/H
0DATA GENERATORS   - K-VALUE  P-R           THERMAL  P-R
0BUBBLE POINTS
 -------------
0 TEMPERATURE  PRESSURE        ENTHALPY            DENSITY    COMPRESSIBILITY
     DEG C      BAR ABS       KW        KJ/KG      KG/CU M       FACTOR (Z)

   -125.86      10.000   -33103.051   -690.3      432.733       0.3295E-1
   -117.49      14.098   -31685.566   -660.7      410.514       0.4633E-1
   -105.82      21.596   -29506.869   -615.3      375.083       0.7227E-1
    -96.66      29.101   -27573.061   -575.0      342.541       0.1011
    -89.60      35.883   -25901.076   -540.1      313.802       0.1308
    -83.55      42.325   -24299.891   -506.7      286.002       0.1639
    -79.50      46.896   -23116.592   -482.0      265.461       0.1916
    -76.79      50.022   -22268.795   -464.4      250.852       0.2133
    -74.86      52.252   -21638.211   -451.2      240.102       0.2305
    -73.23      54.118   -21092.957   -439.8      230.931       0.2462
    -71.96      55.564   -20657.141   -430.8      223.704       0.2593
    -70.76      56.909   -20247.248   -422.2      217.030       0.2721
    -69.66      58.144   -19868.316   -414.3      210.989       0.2844

    -68.80      59.088       **CRIT**

 DEW POINTS
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    -68.20      59.761   -19384.664   -404.2      203.541       0.3009
    -67.44      60.611   -19147.086   -399.3      200.048       0.3093
    -66.96      61.163   -19003.545   -396.3      198.020       0.3146
    -66.88      61.258        MAX.P
    -66.88      61.258        MAX.T
    -66.95      61.171   -18999.854   -396.2      197.961       0.3147
    -67.71      60.156   -19195.822   -400.3      200.408       0.3069
    -69.69      57.263   -19665.922   -410.1      205.922       0.2870
    -73.24      51.344   -20456.188   -426.6      214.587       0.2514
    -73.48      51.024        MIN.P
    -73.50      51.063        MIN.T
    -64.81      64.716   -18661.279   -389.1      196.101       0.3327
    -59.69      71.181   -17560.498   -366.2      184.837       0.3789
    -52.65      78.054   -16026.257   -334.2      169.216       0.4393
    -41.00      84.224   -13431.892   -280.1      143.002       0.5328
    -35.98      84.857        MAX.P
    -22.98      80.180    -9179.101   -191.4      100.477       0.6699
    -15.27      72.307    -7192.663   -150.0       80.609       0.7305
    -10.53      64.746    -5862.578   -122.3       67.132       0.7713
     -7.11      57.222    -4802.255   -100.1       56.137       0.8047
     -4.64      49.707    -3917.464    -81.7       46.621       0.8339
     -3.01      42.195    -3170.686    -66.1       38.120       0.8606
     -2.24      34.686    -2549.077    -53.2       30.367       0.8855
     -2.19      32.450        MAX.T
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   -2.48       27.178    -2058.269    -42.9     23.190      0.9093
   -4.11       19.669    -1729.245    -36.1     16.463      0.9327
   -8.13       12.169    -1653.514    -34.5     10.086      0.9561
  -15.73        5.922    -2016.382    -42.0      4.947      0.9766
  -10.08       10.000    -1712.052    -35.7      8.290      0.9631
   -5.42       16.487    -1656.631    -34.5     13.722      0.9426
   -2.97       23.987    -1895.495    -39.5     20.283      0.9193
   -2.20       31.480    -2322.453    -48.4     27.240      0.8958
   -2.19       32.462        MAX.T
   -2.57       38.973    -2888.764    -60.2     34.713      0.8714
   -3.84       46.466    -3579.580    -74.6     42.848      0.8457
   -5.93       53.957    -4399.396    -91.7     51.857      0.8178
   -8.89       61.445    -5372.058   -112.0     62.085      0.7866
  -12.95       68.926    -6555.139   -136.7     74.179      0.7500
  -18.72       76.387    -8099.147   -168.9     89.696      0.7030
  -29.44       83.684   -10747.485   -224.1    116.119      0.6210
  -36.11       84.856        MAX.P
  -52.16       78.437   -15920.705   -332.0    168.144      0.4433
  -59.86       70.995   -17595.572   -366.9    185.195      0.3775
  -65.58       63.638   -18825.771   -392.6    197.789      0.3255
  -70.35       56.320   -19842.680   -413.8    208.255      0.2801
  -73.23       51.359   -20455.717   -426.6    214.595      0.2514
  -73.59       50.713        MIN.T
  -73.40       51.055   -20492.592   -427.3    214.986      0.2497
  -71.98       53.542   -20180.549   -420.8    211.628      0.2641
  -69.91       56.918   -19716.521   -411.1    206.493      0.2848
  -68.15       59.547   -19303.666   -402.5    201.714      0.3024
  -67.08       61.008   -19034.018   -396.9    198.401      0.3134
  -66.90       61.238   -18985.877   -395.9    197.781      0.3153
  -67.26       60.815   -19092.852   -398.1    199.273      0.3113
  -67.95       60.041   -19306.926   -402.6    202.391      0.3036



 BUBBLE POINTS
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  -69.34       58.492   -19761.426   -412.1    209.310      0.2880
  -70.43       57.286   -20131.205   -419.8    215.164      0.2758
  -71.60       55.971   -20533.352   -428.2    221.674      0.2631
  -72.85       54.554   -20961.748   -437.1    228.741      0.2501
  -74.17       53.040   -21409.131   -446.4    236.230      0.2370
  -75.57       51.433   -21871.607   -456.1    244.064      0.2240
  -77.34       49.387   -22445.006   -468.0    253.878      0.2086
  -79.72       46.645   -23182.232   -483.4    266.595      0.1900
  -82.66       43.314   -24047.904   -501.5    281.620      0.1696
  -86.40       39.225   -25075.719   -522.9    299.488      0.1473
  -91.30       34.170   -26321.418   -548.9    321.066      0.1229
  -98.00       27.907   -27872.322   -581.2    347.632      0.9626E-1
 -107.45       20.418   -29826.855   -622.0    380.376      0.6804E-1
 -117.59       14.039   -31704.520   -661.1    410.815      0.4614E-1
 -132.04        7.585   -34084.691   -710.8    447.705      0.2522E-1
1GENESIS * Asap *    Release 9.01    (31 Mar 93)        BP International
                                                                  Page  12
 ***************                                                  *******
                   ****** NON-FATAL ERRORS HAVE BEEN INDICATED
                                                   EFV01 EFV DATA TABULATION
  OUTPUT FROM MODULE NO.   4                       ***********************
 **************************
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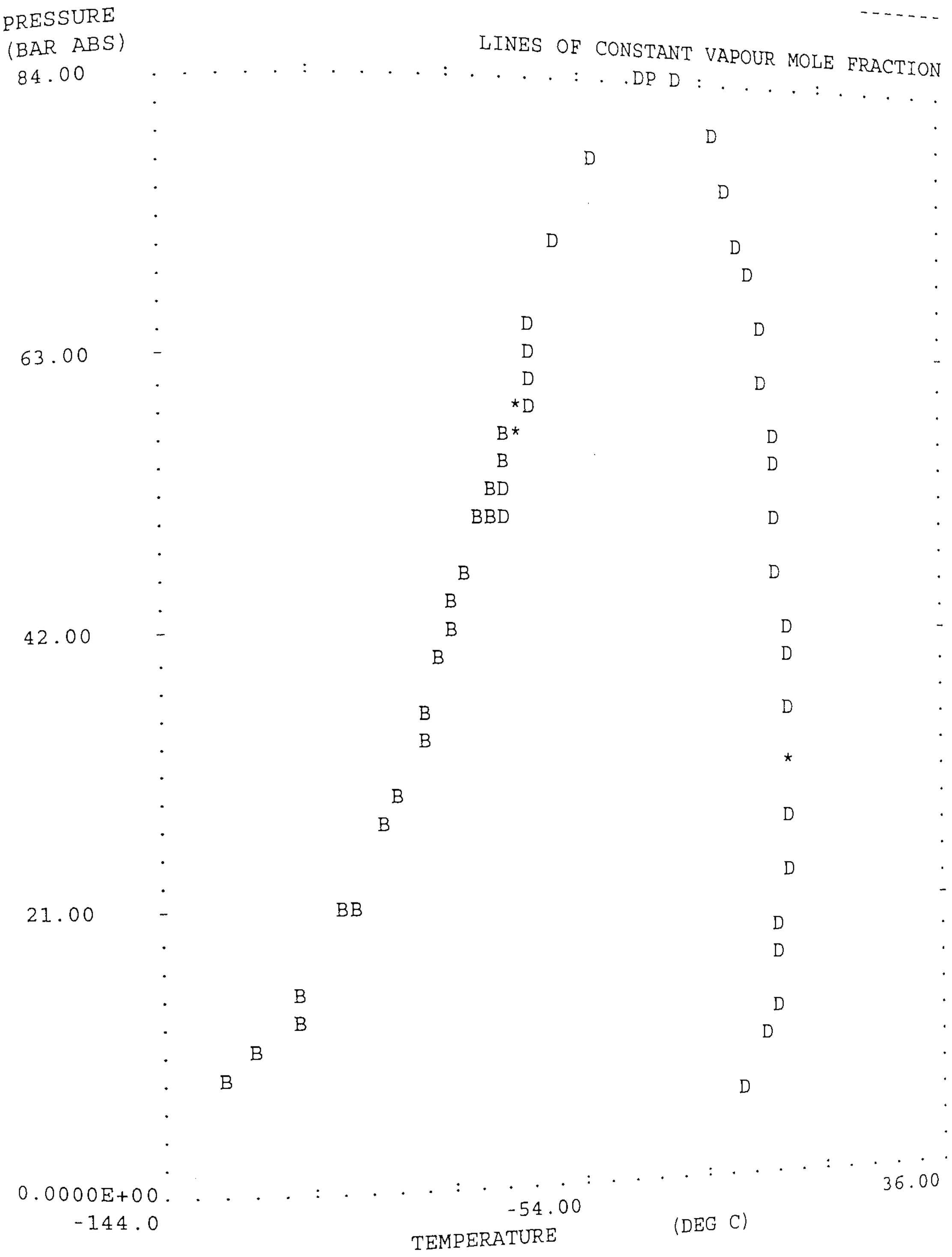
PHASE ENVELOPE   18 FEB 88   PAGE 4



KEY - LINES OF CONSTANT VAPOUR MOLE FRACTION

B = 0.0000E+00
D = 1.000
B = 0.0000E+00
C = CRITICAL POINT
T = MAX/MIN TEMPERATURE
P = MAX/MIN PRESSURE

A1-4 EXAMPLE OUTPUT OF GAS/GAS EXCHANGER CALCULATION

```
OUTPUT FROM MODULE NO. 1                          HEX05 HEAT EXCHANGER


 GAS-GAS EXCHANGE                                        03 JAN 80

     STREAM NO.              4
     TEMP.   DEG C      -2.000
     PRESSURE BARA      79.500
     WT. VAP. FRAC     0.99937 ---<---+
     HEAT      KW -5.7630E+03         |
                                      |
                                      |
                                      |
  STREAM NO.               8     +--+---------+      STREAM NO.      10
  TEMP.    DEG C      -9.883     |            |      TEMP.DEG C -3.953
  PRESSURE BARA       63.500 --->---+---/\/\/\---+--->---PRESSURE BARA  63.0
  WT. VAP. FRAC     1.00000     |            |      WT. VAP. FRAC 1.000
  HEAT      KW -5.6722E+03     +---------+--+      HEAT KW -4.8002E+03
  FLOW KG/H     172637.953                |
                                          |
                                          |      STREAM NO.            2
                                          +---<--- TEMP.   DEG C     4.000
                                                 PRESSURE BARA    80.000
                                                 WT. VAP. FRAC   1.00000
                                                 HEAT     KW -4.8910E+03
                                                 FLOW KG/H     173486.109


 0HEAT DUTY                               KW  8.72016E+02
  CORRECTED LMTD                       DEG C          7.10
  LMTD CORRECTION FACTOR                            0.8969

            CONFIGURATION
            -------------

  COUNTER-CURRENT FLOW

  TEMA SHELL   SHELLS   SHELL    TUBE
     CODE               PASSES   PASSES

     E            1        1        2
```

A1-5 PRESSURE ADJUSTMENT FACTORS

The cost data in the economic evaluation chapter is based on a equipment design pressure of 100 bar. The following factors or equation can be used to adjust the cost for alternative inlet pressures.

Inlet pressure [bar]	pressure factor
75	0.90
100	1.00
150	1.20

The following pressure factor (PF), equation is obtained by curve fitting the data.

$$PF = 0.004\ Pressure + 0.6$$

A1-6 SPECIMEN CALCULATION FOR EQUILIBRIUM PROCESSES IN THE TWO PHASE REGION OF THE PHASE ENVELOPE

pressure bars	temperature C	cricondentherm C	calorific value btu/ft3	wobbe number btu/ft3	condensate m3/h	condensate kmol/h	C5 + kg/h	C3 & C4 kg/h	cost of stabilisation £(1000)	gas to be recompressed kmol/h
25	30	33.76	1179.89	1357.48	3.11	21.06	1668.06	195.50	134.37	9939.10
25	10	14.81	1162.28	1347.14	10.86	101.98	4921.92	1544.27	567.55	9909.56
25	0	4.81	1148.53	1338.96	17.60	178.20	6544.03	3321.51	630.34	9889.38
25	-10	-4.83	1130.81	1328.29	27.48	299.68	7943.59	6489.78	722.56	9871.18
25	-20	-14.65	1108.94	1314.98	40.82	477.74	8895.50	11062.31	846.97	9858.52
25	-40	-34.84	1061.10	1287.21	76.36	1022.55	9668.06	21301.49	1178.48	9848.11
25	-60	-54.66	1019.36	1270.45	126.97	1950.44	9833.64	28143.44	1650.48	9845.85
25	-80	-71.46	996.82	1283.64	220.74	3907.58	9863.42	30702.36	2525.07	9845.44

cal. power KW	25% above cal. power kW	compressor cost £ (1000)	Amount of fuel gas mmscf/d	cost of fuel gas/year £ x1000	value of total condensate (£ X1000)	C5+ kmol/h	C3 & C4 kmol/h	gas to be valued kmol/h	value of gas £(1000)
11397.10	14246.38	5025.32	4.34	2612.38	1558.64	13.70	3.74	9946.46	129.02
11363.23	14204.04	5007.97	4.32	2604.62	4599.04	50.60	29.47	9909.56	126.62
11340.09	14175.11	4996.14	4.32	2599.31	6114.74	70.78	63.45	9889.38	124.87
11319.22	14149.02	4985.48	4.31	2594.53	7422.49	88.98	124.63	9871.18	122.72
11304.70	14130.88	4978.08	4.30	2591.20	8311.96	101.64	214.65	9858.52	120.19
11292.77	14115.96	4972.00	4.30	2588.47	9033.84	112.05	424.67	9848.11	114.88
11290.17	14112.72	4970.68	4.30	2587.87	9188.55	114.31	571.11	9845.85	110.34
11289.70	14112.13	4970.44	4.30	2587.76	9216.38	114.72	626.97	9845.44	107.89

A1-7 SPECIMEN CALCULATION FOR A NON-EQUILIBRIUM PROCESS (MEMBRANE)

area of membrane [m2]	cricondentherm temperature [C]	condensate [kmol/h]	volume of condensate [m3/h]	C5+ [kmol/h]	C5+ [kg/h]	calorific value [btu/scf]	wobbe number [btu/scf]	cost of stabilisation [£1000]
30	6.28	169.91	11.62	40.12	4434.19	1048.03	1349.38	574.57
50	1.72	244.68	15.24	44.64	4830.49	1045.31	1347.74	608.35
70	-2.64	320.15	18.84	51.32	5151.64	1041.92	1345.71	641.93
90	-6.81	396.15	22.42	51.34	5412.65	1039.76	1344.26	675.33
130	-14.67	548.89	29.53	55.82	5798.94	1035.66	1341.69	741.62
180	-23.69	740.39	38.32	59.46	6107.33	1031.22	1338.81	823.62
250	-35.10	1007.97	50.48	62.35	6348.34	1026.36	1335.55	937.02
320	-45.33	1274.09	62.48	63.83	6470.12	1022.55	1332.87	1049.01

gas to be valued [kmol/h]	value of condensate [£1000]	value of gas [£million]
9920.04	4143.31	114.30
9915.52	4513.61	113.95
9908.84	4813.69	113.50
9908.82	5057.58	113.27
9904.34	5418.53	112.77
9900.70	5706.69	112.24
9897.81	5931.89	111.68
9896.33	6045.68	111.25