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Conceptual Process Design
Chemical Process Technology

Subject

Recovery of Hydrocarbon Products from a 5 Gsm^3/a Natural Gas Stream

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CPD Bespreking

Recovery of Hydrocarbon Products from a 5 Gsm3/a Natural Gas Streams

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Algemene opmerkingen

Het verslag is wat aan de te dikke kant, dit had in het textuele gedeelte makkelijk kunnen worden gereduceerd door dingen niet te vaak te herhalen, zo worden bijvoorbeeld de proces keuzes zeker drie keer herhaald. Dit wordt mede veroorzaakt door een niet duidelijke lijn in sommige hoofdstukken. De appendices nodigen niet uit tot lezen, dit komt door de grote hoeveelheid "onnodige" informatie en de niet uniforme layout.

1. Samenvatting

In de samenvatting ontbreekt een duidelijke probleemstelling, en de relevantie van het uitvoeren van dit proces. Een sterkte zwakte analyse van het uitgevoerde proces ontbreekt, zodat niet duidelijk beoordeeld kan worden hoe betrouwbaar de gepresenteerde waarden zijn. Ook wordt geen duidelijke aanbeveling om dit project te continueren. Wel wordt vaag weergegeven waar eventueel verbeteringen kunnen plaats vinden. De samenvatting had naar mijn idee bondiger kunnen worden opgeschreven.

2. Inhoudsopgave

Duidelijk en volledig genummerd

3. Inleiding

Ook hier ontbreekt weer een duidelijke probleemstelling en de eventuele opdrachtgever. Wel worden dit keer de mogelijke redenen voor het uitvoeren van dit proces gegeven.
De producten de toepassingen en de marktsituatie worden duidelijk weergegeven evenals de beschikbaarheid van de grondstoffen. Al wordt hier uitgegaan van cijfers op basis van het jaar 1985, terwijl het niet zo moeilijk moet zijn om aan recentere data te komen, hierdoor wordt het hele verhaal lichtelijk ondermijnd.
Het bespreken van andere concurrerende plants is er summier, en het wordt niet echt duidelijk op welke technologie deze zijn gebaseerd.
In de bespreking van de "environmental issues", ontbreekt de molsieve 3A als afvalstroom. Ook had hier naar mijn mening wel wat over de vereisten met betrekking tot wetgeving kunnen worden gegeven.
In "Process selection" wordt beweerd dat het gebruikte patent voor het twee staps expander proces veranderd en verbeterd is, dan zou het noemen van deze verbeteringen wel handig zijn. Voor de NLG extractie wordt de proces keuze genoemd terwijl deze voor de andere unit operations niet wordt genoemd. De verantwoording van het proces, wordt daardoor alleen voor een unit gegeven.

4. Uitgangspunten voor het ontwerp (exogene randvoorwaarden)

De keuze voor Noorwegen als vestigingsplaats is duidelijk, door de nabije lichting van grote gasvelden, het argument voor lagere energie prijzen zou ik graag gekwantificeerd willen zien. De SHE eisen ontbreken, zoals toelaatbare emissies en de eventueel daarmee gemoeid gaande kosten. Alle andere exogene voorwaarden staan allen duidelijk weergegeven. Al lijkt mij een temperatuur van 30 °C voor het koelwater afkomstig van het oppervlakte water lichtelijk aan de hoge kant, waardoor de koelingsduty in het proces veel lager kan worden.

5. Basis van fysische en chemische procescondities (endogene voorwaarden)

De keuze voor het gebruik van de molsieves is gebaseerd op de hoge efficiency bij hoge druk en lage temperatuur. Het afvallen van de overige processen is duidelijk gemotiveerd, alleen de maximale hoeveelheid water dat mag blijven zitten in de stroom om bevriezing te voorkomen wordt niet gegeven. Daarom is het afvallen van het Selexol proces onduidelijk gemotiveerd. De selectie criteria hadden wat duidelijker kunnen worden weergegeven. Wat wordt bedoeld met free water?
Ook zou het wenselijk zijn om de maximale hoeveelheden CO_2 en H_2S die aanwezig mogen blijven te specificeren, en in het geval van H_2S wat de eventuele gevolgen zijn van teveel van deze stof.
Voor de NLG extractie, valt het gebruik van een deethaniser, depropaniser en debutaniser, maar wat zijn dan de niet bevredigende resultaten? Ook het afwijzen van het mixed refrigerant proces gebeurt op grond dat er teveel CO_2 aanwezig is, terwijl eerder is aangenomen dat CO_2 niet hoeft te worden gescheiden. De keuze tussen de andere processen is redelijk duidelijk gemotiveerd al de vergelijking op grond van de gebruikte selectie criteria per selectie criterium in een tabel duidelijker geweest. Ook een kosten schatting van de externe "vrieskoeling" koeling zou de keuze verder kunnen verduidelijken, ook een verwijzing naar andere geschatte kosten zou het duidelijker maken. Het sommeren van de totale duty's uitgedrukt in een bedrag maakt de keuze tussen de verschillende processen duidelijker nu is het heel moeilijk om te beoordelen welk proces economisch voordeliger is.
Voor de keuze van het NLG fractioneringsproces hadden de selectiecriteria beter kunnen worden weergegeven. Waarom is de gebruikte sequentie van distillatie torens gekozen ?
De keuze van de verschillende processen wordt te vaak door het verslag heen herhaald, wat veel onnodig leeswerk oplevert en waardoor het verslag aan de stevige kant is.
De keuze voor het thermodynamisch model is duidelijk, en wordt goed met cijfers en grafieken ondersteund

6. Keuze van een basis structuur van het proces

De keuze van de basisstructuur en recyclestructuur van het proces is duidelijk gemaakt, alleen de keuze van de structuur van fractioneringssectie is niet gemotiveerd. Ook is de parallelle plaatsing van de twee turboexpander niet duidelijk gemaakt. Al om moet de puzzel voor het uitvinden van de beschrijving van de basisstructuur wel uit alle delen van het verslag worden opgelost.

7. Integratie van de processtructuur

Hier onstaat het idee dat de integratie niet ten volle is benut, zeker in de fractionneringssectie lijken nog verbeteringen mogelijk. Waar zijn nog verbeterings mogelijkheden? Wat zijn de gevolgen van het aannemen van een lagere koelwater temperatuur. Welke warmte wisselaars kunnen worden samengevoegd in een unit. De afstemming van procesdrukken is goed gebeurd, grote drukverschillen worden benut voor compressie van het sales gas. Het koelwater systeem is niet geïntegreerd (gefaseerd), wat is de verwachte reductie in operatiekosten.

8. Ontwerp van procesapparaten (inwendig & randen)

De design criteria worden gegeven, maar over het algemeen wordt niet beredeneerd waarom deze zo zijn gebruikt. Ook worden vrijheidsgraden over het algemeen niet gegeven. Tevens mist de beschrijving van de gebruikte methode (alleen literatuur verwijzing), vooral wat betreft wat de basisaannames die wordt gebruikt, ook mist een schatting van de nauwkeurigheid van de berekende waarden. De keuze van de druk en temperatuur is gebaseerd op een optimalisatie, die wel volledig wordt genoemd maar toch niet zo mag worden getypeerd, hier is mij niet geheel duidelijk geworden welke variabelen nu werden gevarieerd, en deze optimalisatie is niet opgehangen aan kosten zodat de resultaten niet direct kunnen worden beoordeeld, presentatie in grafieken zou dit verduidelijken.
Het gebruik van de single stage compressor voor de sales gas compressor lijkt enigszins dubieus, maar dat moet ik nog even na rekenen

9. Proces flowsheet (PFD) met massa- en energie balansen

In PFD ontbreekt de unitnummering. En de PFD had wel iets ruimer kunnen worden opgezet. Lettertype van nummering mag ook wel iets groter. Massabalansen zijn sluitend. Er mist een lucht stroom naar de incinerator.

10. Procesdynamica en beheersing

Er zijn niet echt control kritische apparaten aanwezig, de belangrijkste control elementen zijn aanwezig. De gebruikte regelparameters en controls zijn duidelijk weergegeven.

11. Analyse van procesveiligheid en milieu (SHE)

Wat zijn de emissie eisen ?
Waardoor verliest het 3A zeolite zijn werking? Kan het nog als landfill worden gebruikt?
Hoeveel N_2 zit er lucht, of wordt het off-gas met pure zuurstof verbrandt -> NO_z emissie?

12. Kosten en economische evaluatie

Met welke methoden zijn de F.O.B. kosten en geïnstalleerde kosten van de apparaten berekend?
Wat is de verwachte nauwkeurigheid van de fixed capital investment, en implicaties voor de rendabelheid van het proces (verwachting +/-40 %)?
Met inachtneming van de berekende winstgevendheid en de onzekerheid hierin, moet het project worden uitgevoerd?

13 Conclusies en aanbevelingen

In de conclusie kan in tegenstelling tot de rest van het verslag wel iets meer informatie kunnen worden gegeven. Zoals moet het proces verder door worden ontwikkeld, wat zijn de gevolgen voor het milieu en getallen voor de winstgevendheid. En de verbeteringen ten opzichte van bestaande processen. De aanbevelingen zijn over het algemeen bevredigend.

14+16

Literatuur verwijzing, symbolen lijst zijn aanwezig. De bijlagen bevatten een "overload" aan informatie, en nodigen niet uit om te bekijken, dit wordt voor de hoeveelheid en de inconsequente opmaak hiervan.

Opmerkingen

(Alle punten die ik als onduidelijk heb omschreven, kan als vraag worden opgevat)

Summary

The process presented here recovers the valuable natural gas liquids (NGL, C_3, C_4 and C_5^+) from a 5 Gsm^3/a natural gas stream, with a given on-steam factor of 0.96.
This is done in three steps:

- Feed pre-treatment by molecular sieves to remove water from the feed.
- NGL extraction by expansion, cooling and C_2^- distillation.
- NGL fractionation by distillation.

Molecular sieves are chosen to dehydrate the feed, because of their efficiency and simplicity.
For NGL extraction three options were evaluated:

- The reflux heat exchanger process, which was not chosen due to the need of external refrigeration at low temperatures and unstable and uncertain simulation. An advantage of the process is the possibility to operate at high pressure.
- The Cryomax process, which was not chosen due to high compressor duties. These resulted from operating the process at low pressures, which was done to eliminate the need for external refrigeration. It does have a high recovery and low reboiler duties.
- The Turbo-expander process was divided into two options:
  - The option using one expansion step resulted in higher compression costs and was therefore not chosen.
  - The option using two expansion steps possessed the overall lowest compression costs, did not need external refrigeration, had a 97% C_3-recovery and met all required product specifications.

The Cryomax process and both of the Turbo-expander processes were fully optimised and compared. Based on this comparison, the Turbo-expander process with two expansion steps was chosen.
For NGL fractionation two distillation columns have been used, because other options (e.g. Petlyuk columns) were not possible and distillation columns are standard separation devices.
Using (a) turbo-expander(s) is not a new concept. There are twelve processes relying solely on turbo-expanders and hundreds of plants using similar processes to extract NGL from natural gas streams.
The plant's annual production is 0.342 Mton/a C_3-product, 0.212 Mton/a C_4-product and 0.174 Mton/a C_5-product. Comparing the C_3 production to the production levels of 1985, it is about 3 % of the world's annual production. In comparison to a competitor it handles twice as high a feedrate.
The prices of the feed and sales gas (0.10 Dfl/sm^3) are considered to be the same (per sm^3). The prices for C_3-product, C_4-product and C_5^+-product are 115.50, 144.80 and 144.4 Dfl/sm^3.
The total investment is 163.6 mln Dfl, with an economical plant life of 25 years. This results in a pay out time of 5.2 years, a rate of return of 19.2 % and a discounted cash-flow rate of return of 17.3 %. The process is economically feasible.
All the financial data used is in the "worst case". Product prices are historical low and are expected to rise. They have a large influence on process economics.
Problems encountered in the design are high compression costs, high cooling water streams and a large column diameter of the deethaniser.

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Appendices

1. Introduction

Natural gas is a mixture of hydrocarbons with methane as the main constituent. Natural gas liquids (NGL) are liquefied hydrocarbons, recovered from natural gases. Reasons to recover these natural gas liquids from natural gas could be:

- To adjust the gas to the required standards (e.g. dewpoint).
- The economic value of the natural gas liquids.
- Process safety hazard due to possible condensation.

The process presented in this report recovers liquefied propane and heavier hydrocarbons from a natural gas stream.
The main product is propane. Other products are a stream mainly consisting of methane and ethane (sales gas), butane and a heavy C_5^+-stream (condensates). For the pure component properties the reader is referred to appendix 5-1.

Except for a minor amount used as chemical feedstock, the sales gas is primarily used as clean heating fuel in industry, power plants, commercial installations and households [1].
Propane has many applications. Some of the most important are [2]:

- Use as a chemical feedstock for production of e.g. ethene, propene, glycerol and nitroalkanes (solvents).
- Use as a clean fuel for cooking and heating in the domestic and commercial market.
- Use as a feedstock for alkylation.

Butane is used as fuel and as a chemical feedstock (e.g. for nylon) [2].
Condensates can be added to gasoline to control volatility [3], serve as a refinery or chemical feedstock, or serve directly as fuel [2].

1.1 Environmental issues

Natural gas has a number of environmental advantages compared with other fossil fuels. Compared with coal and oil, natural gas has a 30-50 % lower emission of CO_2 and 60-90 % lower emissions of NO_X to produce the same amount of energy. In addition, natural gas produces only marginal emissions of SO_2 and particles [4].
The process developed here has only two waste streams:

- A gas stream from regenerating the molsievebeds, this stream will be flared.
- A stream of thermally contaminated water, which will be fed to the surface water.

Waste waters from piping and equipment cleaning could be collected and treated in a reclaimer.

Possible energy conservation techniques in natural gas treatment plants are: [5]

- Waste heat recovery from gas engines or gas turbine exhaust.
- Open cycle heat pumps on fractionating columns.
- Cascading.
- Increasing the number of trays in the columns.
- Using cooled plant feed.
- Production of steam in the fired heater

Where possible, attention will be paid to these items.

The hazards associated with NGL are that it is a low-temperature fluid, flammable, explosive with air and can create vapour clouds dense enough to cause asphyxiation. Non of these is a problem when NGL is contained in properly designed tanks or piping and equipment.

Outside the battery limits, the following safety aspects should be considered [5]:

- Storage tank location and orientation.
- Containment of the liquids in the immediate storage area by means of basins.
- Avoiding the use of possible ignition sources in the vicinity of the plant.

Inside the battery limits one could take into account [5]:

- Furnace/stack design.
- Stack location (relative to prevailing winds).
- Electrical area.
- Gas leak-, fire-, smokedetectors, etc.

A detailed safety study is given in chapter 10.

1.2 Marketsituation

Since 1989 the natural gas resources exceed the crude oil reserves (based on amount of energy). This is one of the reasons for increased interest in natural gas as a resource for liquid fuels and small alkanes.

The world gas reserves on 1 January 1990 totalled about $129*10^{12}$ sm^3.[1]

The world's 1989 gross gas production amounted to $2.5*10^{12}$ sm^3 of which about 83 % was marketed. The rest covers reinjection, flaring, losses, etc.

Consumption patterns in various regions of the world generally follow production profiles. Russia and Eastern Europe produce and consume about 40 % of the world's annual natural gas availability, followed by North America with about 30 %. [1].

The natural gas comes from a Northsea field. The price is not only dependent on the field of origin, but also on the distance between the gasfield and the plant. [3]

Fuel costs vary widely from one area to another because of the cost of the fuel itself and the cost of transportation. To maintain its competitiveness versus alternative energies, natural gas prices for the final consumer are generally determined by prices of competitive fuels such as light heating oil, heavy fuel oil, coal, LPG, or electricity (considering the inherent environmental, handling and financial advantages of gas utilisation). A large share of internationally traded natural gas is normally sold under long-term contracts up to 35 years' duration. Prices of these gas deliveries are generally tied to official international quotations. The market prices of the products can be found in chapter 11.

The plant's annual production is:

- 0.342 Mton/a C_3-product.
- 0.212 Mton/a C_4-product.
- 0.174 Mton/a C_5^+-product.

Comparing this to the production levels of 1985, it is about 3 % of the world's annual production. This is further explained in chapter 5.5.

There's little information about the competitor's developments.

There are about 12 turbo expander plants [6] throughout the world, and hundreds of plants using similar processes to yield propane and sales gas [6].

At this time, the Norwegian company Statoil is building a new plant at the Kårstø complex north of Stavanger, which will become operational in October 2000 [4]. The plant has a feedrate of 250 t/h.

1.3 Process selection

Four different process options have been compared:

- Reflux heat exchanger process [6,7].
- Cryomax process [6,8].
- Turbo expander with two expansion steps [6,9].
- Turbo expander with one expansion step.

These processes are all based on the following unit operations:

- Feed pre-treatment.
- NGL extraction.
- NGL fractionating, yielding C_3, C_4 and C_5^+.

For a detailed description and comparison of the processes, the reader is referred to chapter 2. Block schemes are given in appendices 2-1, 2-2, 2-3, 2-4.
Eventually the process with two expansion steps was chosen. The chosen process is based on a patent [9], but has been altered and improved [10]. It is unknown if this process is patented, as there are many patents with only slight differences.
Some aspects that are not included in the design are:

- CO_2 removal. At the temperatures used, no hydrates will form and the hydrocarbon dewpoint of the sales gas can be achieved without a problem.
- H_2S removal. H_2S levels in the feed are already 2.5 ppm. There is no need to lower these levels. [11]
- Mercury removal. The specified feed (appendix 8-1) does not contain mercury, but because of the use of an aluminium plate-fin heat exchanger, it is very important that there is no mercury in any of the process streams. In case of traces of mercury in the feed, the mercury could be removed by molecular sieves.[12,13]

Some problems encountered in the design are:

- The size of the deethaniser; because of the large mass flows in the bottom region, the bottom diameter was estimated to be over 6 meter. The problem could not be solved by adding trays, changing the feed tray locations or feed temperatures. By using different equations (e.g. the Hunt equation [14]) for calculation of column diameter, smaller diameters could be found.
- The large flows of feed and sales gas gave some problems in equipment design. In most cases this was solved by placing the units in parallel or adding extra units.
- For the design of the molsievebed, no relevant kinetic data were available. The length of the mass transfer zone was estimated, using alternative calculation methods. These methods were not meant for detailed designing.
- The fired heater/stack is used for heating a process stream with variable flow and for burning waste gas. Due to this dual nature it could not be designed.
- The use of an interstage reboiler was considered to lower the use of external heating. Simulation of an interstage reboiler was not possible. As there also was no literature found on the subject it has not been evaluated.

For a detailed explanation of the problems and their solutions, the reader is referred to the separate equipment design in chapter 8.

2. Process options and selection

Given the high feed rate of 5 Gsm^3/a of natural gas (appendix 9-1), it is advantageous to operate the process continuously. The process consists roughly of three steps:

- Feed pre-treatment.
- NGL extraction.
- NGL fractionation

In this chapter, these three steps will be considered. The process options are given and the selection will be explained.

2.1 Feed pre-treatment

Common to all natural gas liquefaction processes is the need to pre-treat the feed gas for removal of components that would either freeze out or lead to pollution upon combustion.

2.1.1 Process options for feed pre-treatment

The natural gas feed can contain several unwanted components, such as water, sour components and metals (e.g. mercury). Several processes are available to remove one ore more of these substances from a natural gas stream.

Removal of water
The low temperatures in the process would cause water to freeze out and cause the formation of hydrates. The literature [15,16] yields several different methods for separating water from a natural gas stream to prevent freezing. These include absorption in substances such as glycols, methanol (IFPEXOL), Selexol or adsorption in molecular sieves. Also separation over membranes is possible [6]

Removal of other substances
Sour components, such as CO_2 and H_2S can be removed by means of physical adsorption processes, chemical adsorption processes, or a combination of both. Possible processes are:

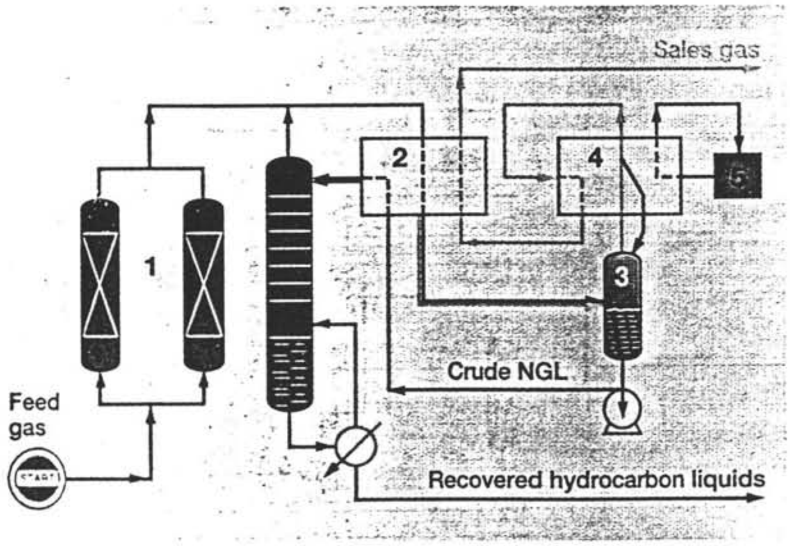
- Physical adsorption: Purisol, Selexol, Fluor solvent, Rectisol [1,6]
- Chemical adsorption: Amine wash [17], Sulfinol [5,6], MEA DEA, DGA [1,6]
- Combination of both: Dipa, MDEA, Amisol [1]
- Other: Membranes [5,17]

Possible traces of mercury, could be removed by a molsievebed, using molsieve 4A. [5,13]. Another option is adsorbtion followed by oxidation, cooling and separation [12]. Also adsorbtion on sulphur impregnated activated carbon [1,13] is a possibility.

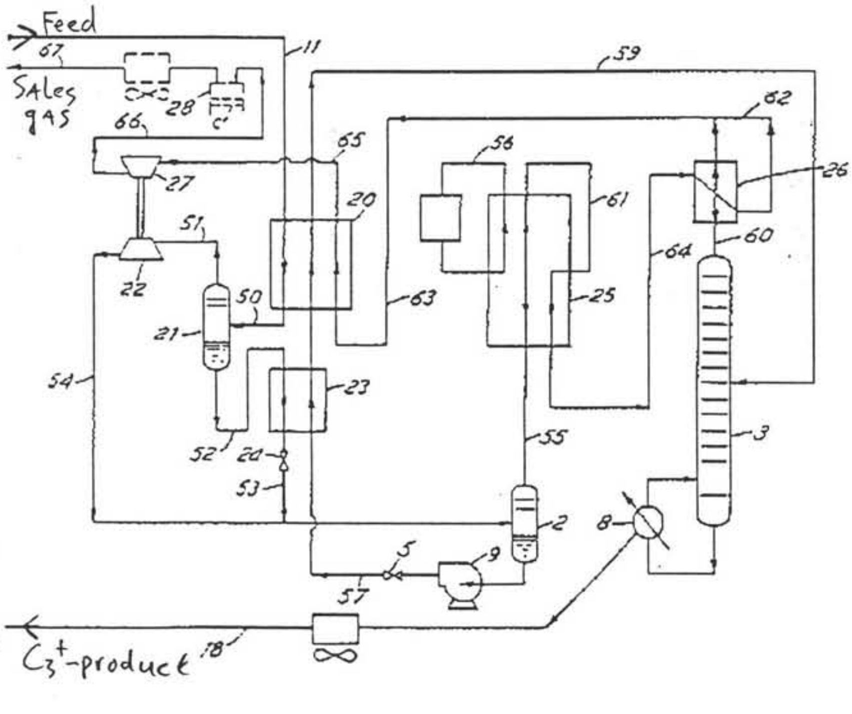
2.1.2 Selection of the feed pre-treatment process

Removal of water
Due to the high feedrate and the high pressure of the feed, the use of membranes in this process would require very large membranes and would lead to considerable

Reflux heat exchanger process



PetroFlux [6]



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losses of products. The Selexol (physical) absorbent can only dehydrate a gas stream to around 150 ppmwt [15], while the processes using glycol and methanol can achieve concentrations of <1 ppmwt [15]. As best solid absorbent, molecular sieves can achieve about 0.1 ppmwt [16].
The glycol-process is quite complex, using a lot of equipment. The methanol-process is operated at 305 K, too high above feed gas temperature. They are often used in the production of liquefied natural gas (LNG), for which it is important to achieve extremely high purity because of the very low operating temperature (113 K). Molsieves operate more efficiently at high pressures and low temperatures. Therefore molecular sieves are chosen to dehydrate the feed to a water concentration of about 0.1 ppmwt. It is interesting that most plants of this type use molsieves for dehydration of the feed [18]. There will be a need for more than one sieve so one (or more) can be regenerated while the other is dehydrating the feed.

Removal of other substances
Given the product specifications, the natural gas does not have to be liquefied. (appendix 9-2 and 9-3) Therefore, extremely low temperatures will not be needed and it will not be necessary to remove the CO_2 as long as the hydrocarbon dewpoint specification of the sales gas is met.
H_2S is already 2.5 ppm, there is no need to lower H_2S levels [11].
The given feed stream does not contain Hg, He, and COS, they do not have to be removed. It is very important that the feed does not contain any mercury, because of the use of an aluminium plate-fin heat exchanger. If mercury is present in the feed, it will most likely be removed by the molsieves. However, this shortens the life time of the molsieves for water removal and would have to be evaluated in depth.

2.2 NGL extraction

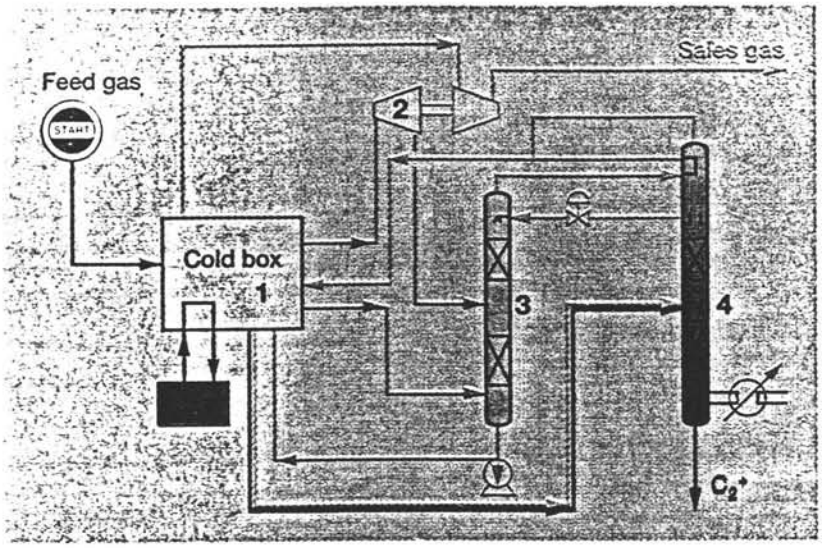
In the NGL extraction, the C_3^+-stream is separated from the sales gas.

2.2.1 Process options for NGL extraction

The simplest way of accomplishing the required extraction is the use of a deethaniser, a depropaniser and a debutaniser in series. This process was examined first, but does not give satisfactory results for application in this process. Literature and patents give several other process options, including the use of mixed refrigerants [6,19], expansion processes and use of a reflux heat exchanger. The mixed refrigerant-process is not chosen because CO_2 is present in the feed, the low process temperatures would cause problems due to the formation of CO_2 solids. The most promising processes have been compared, before choosing the final process. This paragraph gives a description of the options, and guidance through the selection procedure. A short description of the compared processes is given here. For details the reader is referred to appendices 6-4, 6-5, 6-6 and chapter 5.2.

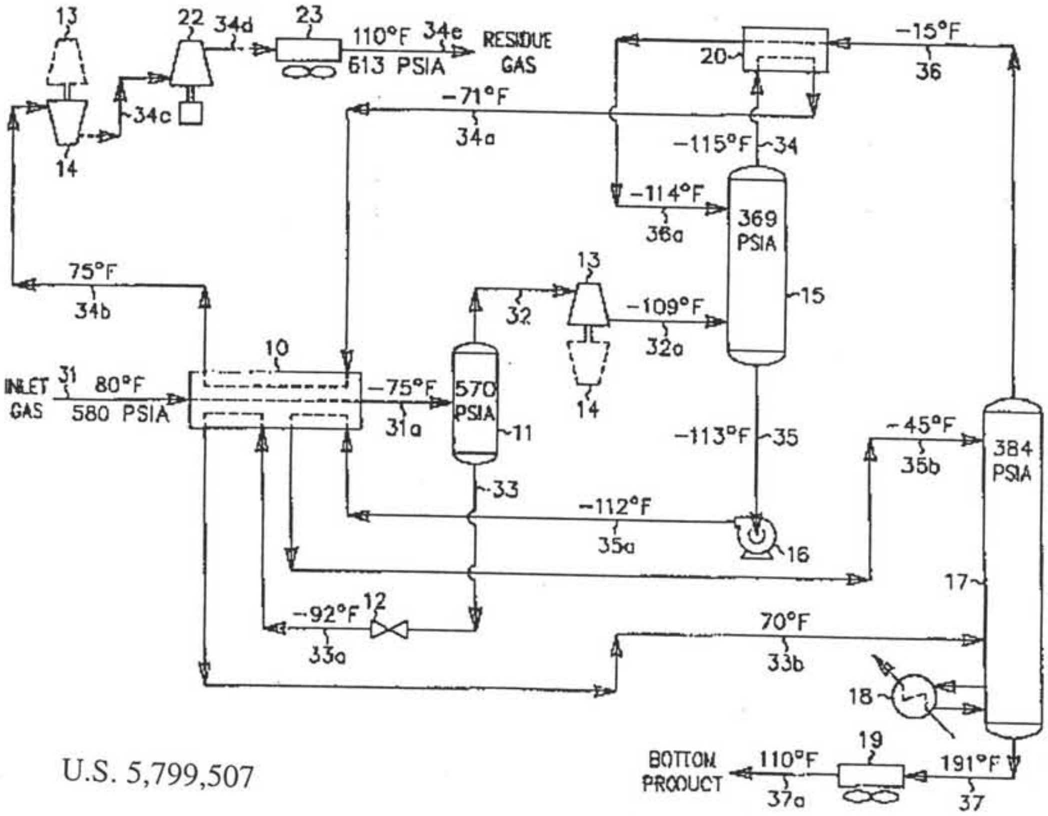
Reflux heat exchanger process
The reflux heat exchanger process, based on the Petroflux process (by Costain Petrocarbon Limited) [7], is characterised by a separator, after which the vapour is cooled. In this cooling step, the higher boiling components are condensed and wash the upcoming vapour while flowing back to the separator. The liquid is expanded in a

Cryomax [6]



CRYOMAX

Turbo-expander process



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valve and fed to a deethaniser, which yields the C_3^+-stream. The vapour is compressed, cooled/condensed and separated. The resulting liquid is fed to the overhead of the reflux heat exchanger. The vapour is mixed with the already existing vapour stream from the first cooling step, this stream is the sales gas stream. The blockscheme and description of the process are given in appendices
2-4 and 6-6.
Another process option is recycling the vapour from the deethaniser to the reflux heat exchanger [6].

Cryomax process
In the Cryomax process (by Technip) [6,8], the feed is expanded in one step to the desired pressure. Expanding leads to cooling and condensating of the feed. The mixture is washed in an absorber and the resulting vapour is the sales gas. The liquid is fed to a deethaniser, which produces the liquid C_3^+ product. The vapour of the deethaniser is cooled/condensed and separated. Part of the liquid from the separator acts as reflux to the deethaniser and the rest is mixed with the vapour from the separator to be fed to the absorber.
The blockscheme and description of the process are given in appendices 2-3 and 6-5.

Turbo-expander process with two expansion steps
In the Turbo-expander process the feed is expanded in two stages. After the first expansion step, the liquid-vapour mixture is separated. The liquid is fed to a deethaniser and the vapour is cooled, expanded and separated again. The vapour from this second separator is fed to the bottom of an absorber, the liquid stream is fed to the deethaniser. The deethaniser bottoms form the C_3^+-liquids. The vapour from the top of the deethaniser is cooled and separated. The liquid formed in the cooling step is partly used as a reflux for the deethaniser and partly as a washing liquid for the absorber. The vapour from the separator (mixed with the vapour from the second separator) is fed to the bottom of the absorber. The liquids, leaving the absorber are also fed to the deethaniser. The vapour from the absorber is the sales gas stream.
This process is based on [9] and has been altered (chapter 5) to give the process shown in appendix 2-2. A detailed description is given in appendix 6-4.

Turbo-expander process with one expansion step
This process is a simplification of the Turbo-expander process with two expansion steps. The difference with the preceding process is the use of one expansion step instead of two. This implies the use of less process equipment. The feed is expanded and separated. The vapour is fed to the bottom of the absorber, the liquid is fed to the deethaniser. The vapour from the deethaniser is cooled and separated. The liquid formed in the cooling step is partly used as a reflux for the deethaniser and partly as washing liquid for the absorber. The vapour is fed to the bottom of the absorber. The liquids from the absorber are also fed to the deethaniser, which produces the C_3^+-stream. The vapour from the absorber is the sales gas stream.
The description is given in appendices 2-2 and 6-4

2.2.2 Selection of the NGL extraction process

From these four processes, one process must be chosen for further process design. The selection is based on the following criteria:

- Product specifications.
- Sales gas specification.
- Compressor duty of the sales gas compressor.
- Reboiler duty of the deethaniser.
- Recovery of propane.
- Need for external refrigeration.
- Equipment costs.
- Recycle streams.

The main objective of the selection is to maximise propane recovery, while at the same time minimising duties and costs. Of course, product and sales gas specifications have to be met.

The first two criteria are stated in chapter 3, Basis of design.

The need for external refrigeration is taken into account as a selection criterion, because external refrigeration at low temperature is very complex and therefore expensive. The influence of equipment costs is very small, compared to compressor and (deethaniser-) reboiler duty. It has been found that (estimated) equipment costs have little influence on choosing between the four options.

Some designing aspects that have not been used here as criteria, include:

- Sales gas production.
- Turndown.

Sales gas is not the main product of this process (although it is one of the largest process streams), so the sales gas production has not been used to compare the different options.

Turndown is not yet considered in this stage of development, it will be taken into account in the final process.

The process options have been simulated in the flowsheeting program CHEMCAD III (chapter 8.1). In appendix 6-1 the four different options have been compared, without fully optimising the processes and assuming the feed does not contain any water. This first comparison gives the following advantages and disadvantages.

Table 2.1: Process pro's and con's

Process	+	-
Reflux heat exchanger	-C2/C3 can be adjusted, to influence reboiler duty, amount of trays in deethaniser and recovery. -Does not use a contactor -Low feedrates in the deethaniser -Operates at high pressure	-External refrigeration needed at low temperature -Unstable simulation due to multiple recycle loops -Difficulty and uncertainty of -Modelling a reflux heat exchanger -Much loss of potential work due to the use of valves

Cryomax	-Larger recovery of work from feedstream. -Stable simulation -High propane recoveries possible	-Large pressure difference in the process. -27.5 MW compressor duty. -Lowest temperature in process 198 K -92.3 % recovery
Turbo-expander Two expansion Steps	-97.7% recovery. -Less heat exchangers (many grouped in one coldbox). -24.1 MW compressor duty -No extra refrigeration needed -Stable simulation -Very high propane recoveries possible (up to 99.9 %)	-Loss of potential work due to the use of a valve
Turbo-expander One expansion Step	-99.3 % propane recovery -Only one turbo-expander -Stable simulation -Less equipment -No extra refrigeration needed	-24.6 MW compressor duty -Problems if one turbo-expander fails -Reboiler duty of 18 MW

Because of the need for very expensive external refrigeration and the unstable simulation, the reflux heat exchanger is not further developed from this point on. The three remaining processes have been fully optimised and extended with a distillation train. Also the possibilities for heat integration were examined. After optimising, the three completely developed processes were compared in appendix 6-3. Detailed equipment design was not considered yet (although reasonable estimations have been made), process equipment was compared based on number and estimated costs (Chemcad III). Table 2.2 gives a summary.

Table 2.2. Process comparison after optimisation

	Cryomax	Turbo-expander Two expanders	Turbo-expander One expander
Recovery (wt %)	99.8	98.4	99.95
C_2/C_3 (wt %)	1.0	1.0	1.0
Heat integration	Only at low pressures	Good possibilities	Good possibilities
Reboiler duty (MW)	9.58	16.71	17.8
Compressor duty (MW)	34.7	24.1	25.9

The two turbo-expander processes have been compared separately in appendix 6-3. It shows here that the higher propane recovery and lower equipment costs of the process with one expansion step, can not compare to the higher compression costs and reboiler duty. That leaves Cryomax and the turbo-expander process with two expansion steps as the final options. These will be financially optimised and compared.
Table 2.3 gives the results of this economical evaluation, in terms of savings.

Table 2.3. Economic comparison of the Cryomax and turbo-expander process

	Savings in comparison to each other (Mln Dfl)	
	Cryomax	Turbo-expander, two expanders
Recovery	0.95	
Reboiler duty	2.94	
Compressor duty		11.58
Total	3.89	11.58

As can be seen in table 2.3, the high compression costs are the main reason for not choosing Cryomax. The higher propane recovery and lower reboiler duty have relatively small influence on the total costs of this process.

After comparing the two remaining processes, it can be seen that compression costs are the most important aspect in choosing the final process. The turbo-expander process with two expanders has been selected as a basis for further process design.

2.3 NGL fractionation

After separating the NGL and the sales gas, the C_3^+-bottomproduct of the deethaniser, is split into a C_3, C_4 and a C_5^+-stream.

2.3.1 Process options for NGL fractionation

Recovery of heavy hydrocarbons is done by distillation. This can be done using a depropaniser and a debutaniser column in series, a divided-wall column (Petlyuk column [20,21,22]) or by absorption in heavy hydrocarbons.

2.3.2. Selection of the NGL fractionation process

Literature on the subject of a divided-wall column [20, 21, 22] gives 7.8 % higher equipment costs for a divided-wall column, but 16.8 % lower operating costs. The increase in equipment costs is due to the larger column diameter needed, along with thicker walls.

The Petlyuk column shows considerable energy savings in several cases. However, it can be competitive with the heat-integrated two-column system only in those cases where the concentration of the middle component is high, the split between the first and second components is harder than the split between the second and the third component, and the required separation is not too sharp.

In this case the concentration of the middle component is not very high and the required separation is sharp (recoveries over 99% and purities over 98%). Due to these facts, the use of a Petlyuk column is not possible in this process.

Using heavy hydrocarbons from the process for recovering the various products, did not lead to any improvements in the process. The C_3^+-stream will therefore be fractionated in a depropaniser and debutaniser in series.

3. Basis of design

3.1 Description of the process

Natural gas is a mixture of hydrocarbons with methane as the main constituent. Natural gas liquids are liquefied hydrocarbons, recovered from natural gases. This is done to adjust the gas to required standards, process safety considerations and the economic value of the natural gas liquids.
This process recovers liquefied propane (C_3) and heavier fractions from a natural gas stream. The main product is propane (C_3). Other products are sales gas, butane (C_3) and a heavy C_5^+-stream (condensates).

The process, as described in the basis of design, is based on the following unit operations:

- Feed pre-treatment,
- NGL extraction and
- NGL fractionation.

What will be done is:

- Removal of all free H_2O to prevent freezing and hydrate formation.
- Extraction of NGL in various cooling, expansion and separation steps.
- Fractionation of NGL, yielding propane (C_3), butane (C_4) and condensates (C_5^+) by distillation.
- Delivering sales gas, C_3-product, C_4-product and C_{5+}-product at the desired specifications.

What will not be done is:

- Removal of CO_2 from the feed stream. With the given specifications the natural gas does not have to be liquefied. Therefore extremely low temperatures will not be needed and it will not be necessary to remove the CO_2 as long as the hydrocarbon dewpoint in the sales gas is met.
- Removal of H_2S from the feed, because the fraction is already 2.5 ppm [11].
- Removal of Hg, He, and COS, because the feed stream does not contain these substances. If Hg is present, it could be removed with molecular sieve 4A.
- Design of fired heater/stack, because the dual nature of this piece of equipment makes the design non-standard and it is not part of the main process.
- Design of valves and controllers, as this goes beyond the scope of this report.

3.2 Process definition

3.2.1 Process concept chosen

Given the high feed rate of 5 Gsm^3/a it is advantageous to operate the process continuously.

The process consists of three steps:

- H_2O removal in a feed pre-treatment step to about 0.1 ppmwt.
- NGL extraction.
- NGL fractionation.

Feed pre-treatment

The only feed treatment step in this process consists of water removal by adsorption, to prevent ice and hydrate formation in the process. There are several different methods for separating water from a natural gas stream to prevent freezing [6,15]. These include absorption in substances such as glycols, methanol, Selexol or adsorption on molecular sieves. Also separation over membranes is possible. [6]. For more detail on selection the reader is referred to chapter 2.1.2.

Molecular sieves are chosen to dehydrate the feed to a water concentration of about 0.1 ppmwt, due to their efficiency and simplicity. For sufficient capacity and smaller equipment three beds will be chosen. Two operate simultaneously, while one is being regenerated.

Other components in the feed such as CO_2 and H_2S could be removed by adsorption processes. This is not necessary due to feed and product specifications, and will not be done.

NGL extraction

The literature on separating propane and higher hydrocarbons from natural gas yields several options, using turbo-expander(s) and separator(s), mixed refrigerants and a reflux heat exchanger with separator. The mixed refrigerant-process is not chosen because the low process temperatures will cause problems due to the formation of CO_2 solids. This process also requires external refrigeration. Both the turbo-expander processes and the process using the reflux heat exchanger could make use of external refrigeration loops, giving them an element of the mixed refrigerant process. These processes operate at higher temperature (203 K), where a mixture containing CO_2 does not pose as great a problem. Therefore the choice of a sales gas separation system has focussed on processes using one or more turbo-expanders with separators and processes using a reflux heat exchanger with separator.

Four process options are presented for sales gas separation:

- Reflux heat exchanger process
- Cryomax process
- Turbo expander process with one expansion step
- Turbo expander process with two expansion steps

Block schemes and descriptions are given in appendices 2-1, 2-2, 2-3, 2-4, 6-4, 6-5 and 6-6.

All four processes have been simulated and evaluated (chapter 2). The reflux heat exchanger process was not chosen because of expensive refrigeration and unstable simulation. After fully optimising the three remaining processes, the Turbo-expander processes were compared separately, and with the Cryomax process. The latter was not chosen due to the large sales gas compressor duty. The turbo expander process with one expansion step was also discarded, due to high compressor duties. Therefore the Turbo-expander process with two expansion steps has been chosen. It was most economical, mainly due to smaller duties of the sales gas compressor.

NGL fractionating

Recovery of heavy hydrocarbons is usually done by distillation. Membrane separation is also possible, but not appropriate in this process due to high methane losses and technical difficulties for large streams [23]. Distillation can be done using a depropaniser and a debutaniser column, a divided-wall column (Petlyuk column) or absorption in heavy hydrocarbons.

Literature on the subject of a divided-wall column [20,21,22] gives higher equipment costs for a divided-wall column, but lower operating costs. The increase in equipment costs is caused by the larger column diameter needed, along with thicker walls. With the feed and product specifications in this process, the use of a Petlyuk column will not be adequate and will therefore not be simulated.
Using heavy hydrocarbons from the process as a washing agent, did not show any improvement in the yield.
Therefore NGL fractionation will be done in a depropaniser followed by a debutaniser column.

3.2.2 Block scheme

For the block schemes of the four evaluated processes the reader is referred to appendices 2-1, 2-2, 2-3 and 2-4. Appendix 2-2 contains the block scheme of the finally chosen process.

3.2.3 Thermodynamic properties

The system is characterised by high pressures, low temperatures and hydrocarbons with low carbon numbers. For such systems there are many different thermodynamic models available. Among these the cubic equations of state developed by Soave-Redlich-Kwong (1972) and Peng-Robinson (1976) and their modifications are the most frequently used [24]. They are considered to be the most accurate of the cubic equations of state [25]. The Peng-Robinson equation of state is the best for use at low temperature [26]. It is also used for N_2, CO_2 and H_2S, the major non-hydrocarbons in the system.
Therefore the Peng-Robinson cubic equation of state is the thermodynamic model selected for the process.
Peng and Robinson have validated their equation of state for all components in the feed mixture for temperature ranges of 116.5-283.2 K for methane and nitrogen to 297.0-588.7 K for n-octane, using pressures ranging from 13 up to about 100 bar [27]. Though the highest pressure in the system is above this value, no separations (apart from dehydration) are performed at pressure above 100 bar. Comparison of various Pxy-diagrams generated by Chemcad III with the Peng-Robinson model and diagrams from the literature [28] showed no great deviations (appendix 5-7). Therefore it is assumed that the Peng-Robinson model will give an adequate description of the processes that occur.

3.2.4 Pure component properties

For the list of pure component properties the reader is referred to appendix 5-1.

3.2.5 Process stream summary and mass balance

Using the block scheme presented in appendix 2-2 and assuming completely dehydrated feed, the process streams can be calculated using a flowsheeting program (Chemcad III). The process stream summary is given in appendix 5-2. The heat and mass balances are given in appendix 5-3. The utility usage is given in appendix 5-4. The process yields are given in appendix 5-5. All product specifications were met. (see appendix 8-12)

3.3 Basic assumptions

The contents of this chapter are derived from appendix 9-1. The standard conditions (sm^3) are 293.15 K and 1 bara.

3.3.1 Plant capacity

Product and feed specifications

Feed gas:

Feed rate:	5.00 Gsm^3/a, based on 8400 operating hours per year (on-stream factor: 0.96)

Composition of Feed Gas (see appendix 8-1 for calculation of the feedstream):

Table 3.1: Composition of feed

	mol%	wt%	kg/s
N_2	0.400	0.543	0.78
CO_2	3.365	7.174	10.27
C_1	81.430	63.282	90.59
C_2	8.230	11.988	17.16
C_3	3.730	7.967	11.41
i-C_4	0.630	1.774	2.54
n-C_4	1.130	3.182	4.55
i-C_5	0.325	1.136	1.63
n-C_5	0.330	1.153	1.65
n-C_6	0.245	1.023	1.46
n-C_7	0.120	0.583	0.83
n-C_8	0.030	0.166	0.24
H_2S	0.00024	0.00041	0.00
H_2O	0.035	0.031	0.04
Total			143.16

All hydrocarbons are considered to be saturated, e.g. C_3 is assumed to be propane, not propylene. In the following table the feed gas specifications are given.

Table 3.2: Feed gas specifications

Temperature [K]	278
Pressure [barg]	109
Phase [V/L/S]	V
Manner of transportation	Pipe line
Price [Dfl/sm^3] [ch. 11]	0.10

The plant turndown is assumed to be 50 % maximum.

Products:

The specifications of the sales gas are shown in the following table:

Table 3.3: Sales gas specification

In or out	Out	Specifications		
Temperature [K]	< 323		Commercial	Design
Pressure [barg]	189	Hydrocarbon dewpoint (at 50 bara):	258 K	< 258 K
Phase [V/L/S]	V			
Manner of transportation	Pipe line	Water dewpoint:	< 0.030 g/sm^3 ± 213 K	173 K
Price [Dfl/sm^3] [ch. 11]	0.10	Temparature	< 323 K	322 K

The specifications of the C_3-product, C_4- product and C_5^+- product streams are shown in the following tables. The recovery of C_3 has to be > 80 %.

Table 3.4: C_3-product specifications

In or out	Out	Specifications		
Temperature [K]	< 311		Commercial	Design
Pressure [barg]	20	Conc. C_2	< 1.5 wt%	1.4 wt%
Phase [V/L/S]	L	Conc. C_4	< 1.0 wt%	0.9 wt%
Manner of transportation	Pipe line	Conc. H_2S	< 40 ppmwt	35 ppmwt
Price [Dfl/sm^3] [ch. 11]	115.50	Temperature	< 311 K	310 K

Table 3.5: C_4-product specifications

In or out	Out	Specifications		
Temperature [K]	< 311		Commercial	Design
Pressure [barg]	8.0	Conc. C_3	< 0.2 wt%	0.1 wt%
Phase [V/L/S]	L	Conc. H_2S	< 15 ppmwt	10 ppmwt
Manner of transportation	Pipe line	Temperature	< 311 K	310 K
Price [Dfl/sm^3] [ch. 11]	144.80			

Table 3.6: C_{5+}-product specifications

In or out	Out	Specifications		
Temperature [K]	< 311		commercial	Design
Pressure [barg]	7.0	Conc. C_4	< 1.0 wt%	0.9 wt%
Phase [V/L/S]	L	Temperature	< 311 K	310 K
Manner of transportation	Pipe line			
Price [Dfl/sm^3] [ch. 11]	144.4			

Wastes

The described process has only three waste streams:

- A gas stream from regenerating the molsievebeds, which will be flared.
- A stream of thermally contaminated water, which will be fed to the public sewage system, as will be the reboiler- and condenser streams.
- Used zeolite, from the molsieve beds.

Waste waters from piping and equipment cleaning could be collected and treated in a reclaimer.

3.3.2 Economical plant life

The economical plant life is 25 years (as given in appendix 9-2).

3.3.3 Location

The plant is located in Køllsnes, a place on the Norwegian west coast. Norway is chosen due to the close vicinity to gas fields such as Troll and possibly lower electricity prices.

3.3.4 Battery limit

Inside the battery limit the plant will contain the following important equipment:

- Molecular sieves to remove the water from the feed.
- A cold box to optimise heat-integration.
- Distillation columns and an absorber to separate the hydrocarbons.
- Turbo-expanders to reduce the pressure of feed gas and compress sales gas.
- Liquid-vapour separators.
- Compressor.
- Pumps.
- Heat exchangers for product cooling.
- Heater/stack.

The following external facilities are assumed to be available outside the battery limit:

- All equipment necessary to provide the utilities, such as steam and water supply (see appendix 5-6 for all available utilities).
- Possibilities for waste water and other waste disposal.
- Piping and/or storage for all products.

3.3.5 Definition of all in- and outgoing streams/substances

For the specifications of the feed gas, sales gas and hydrocarbon product streams the reader is referred to chapter 3.3.1. The design values are used as estimations. After the finishing the design, it will be checked if the criteria are met. Temperatures were chosen 1 K below specification, except for the water dewpoint. The feed is dehydrated to about 0.1 ppmwt. This means a water dewpoint of 173 K.

3.4 Margin

The margins have been calculated under the following assumptions:
Feed gas has been pre-treated and is assumed to be of the same quality as the sales gas and therefore the same price per sm^3.
Flows are used according to appendix 5-2.
Densities according to appendix 5-1 are used.

Table 3.7: Margin calculation

	Dfl/sm^3	Flow (Msm^3/a)	Mln. Dfl/a
Feed gas	0.10 (chapter 11)	5000	500.0
Sales gas	0.10 (chapter 11)	4635	463.5
Total consumed gas		365	36.5
C_3	115.50 [58]	0.5838	67.4
C_4	144.80 [58]	0.3736	54.1
C_5^+	144.40 [58]	0.2630	38.0
Total sales			159.5
Margin			123.0

With the data from table 3 the maximum allowable investment can be calculated, as can be seen in appendix 11-6. The maximum allowable investment is 1453 million Dfl. This investment does not take into account the annual operation costs (utilities, operating labour, taxes, etc.). Therefore this value is (far) too optimistic, but shows that the proposed process can be viable.

sharper estimate of max. allow. investment could be made by assuming that about 50 % of the margin is available for capital investment.

4. Thermodynamic properties

The system is characterised by high pressures, low temperatures and hydrocarbons with low carbon numbers.

4.1 Thermodynamic model selection

There are many thermodynamic models available for systems containing hydrocarbons at various pressures. Of these, the cubic equations of state developed by Soave-Redlich-Kwong (1972) and Peng-Robinson (1976) and their modifications are the most frequently used [24] and are the most accurate of the cubic equations of state [25]. The Peng-Robinson equation of state is the best for use at low temperature [26]. It is also used for N_2, CO_2 and H_2S, the major non-hydrocarbons in the system. Given these facts the Peng-Robinson cubic equation of state is the thermodynamic model selected for the process.

4.2 Data validation

Peng and Robinson have validated their equation of state for all components in the feed mixture for temperature ranges of 116.5-283.2 K for methane and nitrogen to 297.0-588.7 K for n-octane, using pressures ranging from 13 up to about 100 bar [27]. The highest pressure in the system is 190 bara for the sales gas. Though this exceeds the maximum pressures at which Peng and Robinson validated their equation no separations are performed at this high pressure, only compression and expansion. The feed also has a pressure above 100 bar. Again, only dehydration and expansion occur at this pressure. Most separation steps in the process are performed at or below 30 bara, well within the range of pressure validated by [28]. The temperatures at the various separation steps are well within the validated range. Therefore it is assumed that the Peng-Robinson model will give an adequate description of the processes that occur, even above the pressure range of [27].
Comparison of various Pxy-diagrams generated by Chemcad III with the Peng-Robinson model and diagrams from the literature [28] showed no great deviations. For an arbitrarily chosen temperature of 283.15 K the Pxy-diagrams are given in appendix 5-7, figures 1-4, comparing literature and calculated values for ethane/propane and CO_2/ethane mixtures. The deviations are not significant and the model can be used.

5. Process structure and description

5.1 Criteria and selection

The basic block scheme, presented in chapter 2 and appendix 2-2, can now be used to develop a process flow scheme. In the previous chapters, various design criteria and aspects have been mentioned. This information is now used to determine the choice of the following design elements:

- Unit operations.
- Process conditions.
- Heat integration.
- Start up, shut down and turndown.

5.1.1 Unit operations

The appropriate equipment can now be selected, following the block scheme. For detailed selection procedures of the equipment and design aspects, the reader is referred to chapter 8.

what is the task to be done $\Delta(\mathrm{P}, \mathrm{T}, \underline{x})$

Dehydration unit

The dehydration unit has to be designed to remove all water from the feed, to prevent freezing out and the formation of hydrates.

how deep

Methods to remove water from the feed include [6,15] absorption in substances such as glycol, methanol (IFPEXOL) or selexol. Also adsorption on molecular sieves or other adsorbents is possible. A relatively new method is membrane separation.
Membranes are not suitable for application in this process. Because of the high feedrates they would be very large and cause a big pressure drop. Membranes also give a considerable loss in methane and ethane [23].
The Selexol (physical) absorbent can only dehydrate a gas stream to around 150 ppmwt [6], while the processes using glycol and methanol can achieve concentrations of <1 ppmwt [6]. Molecular sieves, as the best solid adsorbents, can also achieve concentrations below 0.1 ppmwt [15,29].
The glycol-processes are quite complex, using a lot of equipment. The methanol-process is operated at 305 K, too high above feed gas temperature.
Molecular sieves can remove all free water in the process. Also, they operate more efficiently at low temperatures and high pressures [31].
Therefore molecular sieves are chosen to remove the water from the feed.
There is also a choice in the type of sieve used [30]. The most widely used synthetic zeolites are 3A, 4A and 5A. These are molecular sieves with different cations, respectively potassium, sodium and calcium. These different cations result in different sizes of openings into the main cavity of the zeolite due to the sizes and valences of the cations. In these zeolites the openings are about 3, 4 and 5 Å respectively. Due to the small opening in molecular sieve 3A it is the only one that excludes most molecules, except water. The 4A adsorbs methane, ethane and propane and the 5A also absorbs higher hydrocarbons. The 3A molecular sieve is a little more expensive, but lasts a year longer (four instead of three years) [30].
To reduce the propane and other hydrocarbon loss, while still removing all the free water from the feedstream, the 3A molecular sieve will be chosen.

Design criteria for the molsieves include a maximum pressure drop of 0.5 bar and a minimum need for sales gas to regenerate. Also a minimum L/D has been defined, to prevent the bed from becoming "pancake" shaped.
The influence on following process will be a small temperature rise (< 1 K), loss of methane, ethane, propane (due to hold up of the feed) and the need for part of the sales gas stream for regeneration.

$\Delta(p, T, \underline{x})$

Turbo-expanders

The expansion of the feed from very high to moderate pressures can be done by a flash or a Joule Thompson valve. These methods have one major disadvantage: all of the work is lost. The use of normal expanders would be a solution, but the expansion causes temperature drop, which causes the formation of a liquid phase. Normal expanders can only handle vapour phase. Therefore a special type of expander is needed: the turbo-expander. This apparatus can handle liquid fractions up to 40 %wt (appendix 9-3). The work of the expander can either be used to generate electricity or to compress another process stream. In this process two expanders are used, both integrated with a sales gas compressor. The feed is expanded in two steps because of the higher efficiency obtained when using two expansionsteps. This was evaluated using Chemcad III.
The expanders are designed for a maximum liquid fraction of 25 %wt. (appendix 9-2)
The maximum suction capacity is 10,000 m^3/h.
The efficiency of the expansion process is estimated to be 0.85, the efficiency of the compression 0.75 [32].
Because of the limited suction capacity, the two turbo-expanders are placed in parallel to prevent the use of extra compressors.

Sales gas compressor

The two turbo-expanders do not have enough power to compress the sales gas to the specified 190 bar. Therefore, an extra compressor is needed. The efficiency of the compression is 0.85 [32].

Plate-fin heat exchanger

When optimising the process, the possibility of combining several streams into one heat exchanger was examined. Because the temperature of the sales gas stream is not limited, five streams could be combined in one heat exchanger. This heat exchanger not only provides heat integration, but also saves the use of separate heat exchangers. The use of a compact heat exchanger combines 5 heat exchangers in one piece of equipment. Plate-fin heat exchangers are a type of compact heat exchangers, consisting of a stack of flat plates (parting sheets) and corrugations, brazed together as a block [33]. A Plate-fin heat exchanger has several advantages over other types of heat exchangers, including [33,34,35]:

- High thermal effectiveness and very close temperature approaches.
- Large heat transfer surface per unit volume.
- Compactness and low weight, which reduces space and foundation requirements.
- Possibility of heat exchange between many process streams, including evaporating or condensing streams.
- Freedom in design with multi-stream and several possible geometries.

The design of the plate-fin heat exchanger is based on a minimum temperature difference of 5 K and maximum block dimensions of 0.6x0.9x6.2 m. [33]

Separators

The separators serve to separate the liquid from the vapour after liquid has formed in the expansion or cooling steps. Three separators are needed in this process. Two separators are placed after the two turbo-expanders. The third one is used to create the reflux to the deethaniser. Frequently used types of liquid-vapour separators are gravity separators, impingement separators (e.g. vane-type) and filter separators. The separators used in this process are vane-type separators, because higher possible velocities enable the use of smaller columns. They also have a low pressure drop.

Absorber

To remove more propane from the sales gas and remove ethane from the C_3^+-stream at the same time, an absorber is used. The topstream of the deethaniser is cooled and the gas fraction is mixed with the already existing sales gas stream. This stream still contains propane, while the liquidstream from the second expansion step still contains too much ethane. To remove excess propane from the sales gas and ethane from the stream entering the deethaniser, an absorber is installed in the process. It is designed to recover as much propane as possible from the sales gas stream.

Deethaniser

To separate the sales gas from the C_3^+-stream, a deethaniser is used. Membranes have been considered, but selectivity would not be sufficient and there would be considerable losses of methane, ethane and propane. Secondly, the plant capacity is much larger (about 2 times) than the largest plant, currently operating with membrane technology. The deethaniser will separate sales gas from the C_3^+-stream, with a C_2/C_3 ratio of 1 %wt. This is estimated to make sure that the C_2/C_3-ratio in C_3-product is within the specifications. The distillation column will be optimised for high recovery and low reboiler duty.

C_3^+ fractionating

The bottom product from the deethaniser is fractionated in a depropaniser and a debutaniser. It is possible to carry out this separation in one column, a divided wall column (Petlyuk). This Petlyuk column shows considerable energy savings in several cases [20]. But it can only compete with the two-column system in those cases where the concentration of the middle component is high, the split between the first and second components is harder than the split between the second and the third component, and the required separation is not too sharp [21,22].

In this case the concentration of the middle component is not very high and the required separation is sharp (recoveries over 99% and purities over 98%). Due to these facts, the use of a Petlyuk column is not possible in this process.

Depropaniser

In the depropaniser, the C_3- fraction is recovered from the heavy hydrocarbon stream. The depropaniser is designed to give a high propane recovery in combination with low reboiler and condenser duties.

Debutaniser

To separate the butanes from the C_4^+-stream, a debutaniser is used. The bottom product is sold as condensates. The design is based on high recovery with low reboiler and condenser duties.

Pumps

Six pumps in this process are used to transport fluids. Only one pump is used to raise the pressure of the propane product stream. Standard pumps are used in the process; there's no need for special equipment in this case.
The design criteria for the pumps are:

- Maximum capacity of a centrifugal pump is 1000 m^3/h.
- Minimum capacity of a centrifugal pump is 0.25 m^3/h.

Product coolers

To meet their specifications (chapter 3.3.1), the products and the sales gas stream have to be cooled before leaving the plant. Cooling can be done by using air- or water coolers. Air-cooled exchangers are the best choice for minimum process temperatures above 338 K. Since all product streams have to be cooled below 338 K, water-cooled exchangers are needed to cool the products.
The heat exchangers are designed with the following criteria:

- Minimum temperature difference of 7 K.
- Maximum allowable pressure drop of 0.2 bar.
- Maximum exchanger area of 600 m^2.

With the selected equipment, the block scheme has now developed into a process flow scheme. It is presented in appendix 1-1.

5.1.2 Process conditions

The process conditions in the flow scheme were estimated by reasonable engineering judgement and comparison with literature and patents. Using the first estimations [6,9,10], the process was optimised for maximum propane recovery. After a rough optimum in propane recovery was reached, the process was optimised for minimum compressor and reboiler duties, including basic economical evaluation. In appendix 7-1, a sensitivity analysis is shown. It also shows the final optimisation steps.
Process improvements (compared to patents and literature) are mainly achieved due to the use of two expansion steps, various feedstreams to the deethaniser, heat integration, splitting the reflux from the deethaniser to use as reflux and wash liquid in the absorber and finding the optimal process conditions for these items.
The results will be explained in the following paragraph.

Pressures

The product streams have to meet with their specified output pressures. The sales gas has to be pressurised to 190 bara, propane has a pressure specification of 21 bara. Butane and C_5^+ require pressures of 9 and 8 bara respectively. The sales gas compressor is used to compress the sales gas after it has been compressed in the turbo-expanders. The pressure in the depropaniser is not the same as the specified pressure for the propane product. The feed of the depropaniser is depressurised to 16 bara. This pressure level was chosen to have a sufficiently high temperature in the top of the depropaniser to be able to operate the condenser with surface water.
The pressure of the feed of the debutaniser is the same as the specified output pressure of the butanes.

The operating pressure of the process can not be too low, because sales gas compression is a very expensive process step, and the sales gas has to be recompressed to 190 bar. The selected pressures are given in table 5.1.

Table 5.1: Selected process pressures

Process condition	Setpoint
Pressure,$_{out}$ first expander	72.5 bara
Pressure,$_{out}$ second expander	30 bara
Pressure,$_{out}$ sales gas compressor	190 bara
Pressure,$_{in}$ depropaniser	16 bara
Pressure,$_{in}$ debutaniser	9 bara

Temperatures

Not all temperatures in the process can be influenced. The ones that can be influenced are bound by certain limits. The temperatures of the streams in the plate fin heat exchanger must be carefully chosen to avoid a pinch. Also, a minimum temperature difference of 5 K between the process streams in the coldbox has to be taken into account. The process temperatures are optimised in appendix 7-1, a summary is given in table 5.2.

Table 5.2: Selected process temperatures

Process conditions	Setpoint
Temperature stream,out <3>	245 K
Temperature stream,out <7>	250 K
Temperature stream,out <12>	235 K
Temperature stream,out <23>	220 K

Trays

In the different columns, the number of trays is optimised to gain the highest recovery. The location of the feedtrays is optimised by comparing feed compositions with the actual tray compositions. The feedtrays are located in parts of the column with matching compositions. For this optimisation, the theoretical number of trays and location of these trays is considered. For actual number of trays and their locations, the reader is referred to chapter 8.2.8.

The number of trays and feed tray locations are optimised in appendix 7-1, the results are given in table 5.3.

Table 5.3: Tray amounts and locations

Process variable	Theoretical Number	Stream ID	Theoretical Location
Trays in absorber	10	<18>	1
		<17>	10
Trays in deethaniser	28	<19>	1
		<22>	8
		<13>	12
		<8>	16
Trays in depropaniser	45	<45>	20
Trays in debutaniser	31	<55>	15

Splitters

Two splitters are used in the process. One to create the deethaniser top reflux stream, and one to match the work of the turbo-expanders.

The splitfactor for the deethaniser was optimised to achieve the highest recovery, before the sales gas compressor was considered. The splitfactor is not optimised for the lowest overall costs, so this may not be the optimal choice. The used splitfactors are given in table 5.4. For detailes the reader is referred to appendix 7-1.

Table 5.4: Splitfactors

Splitter	Stream number	Split factor
Reflux/ absorber	<19>	0.53
	<18>	0.47
Turbo-compressors	<29>	0.61
	<30>	39

5.1.3 Heat integration

For heat integration the process can be split into three parts:

- NGL extraction (the part up to the deethaniser).
- NGL fractionation (deethaniser, depropaniser, debutaniser and all appendages).
- Compression and cooling of the sales gas.

The last two parts can be grouped for heat integration.

During the design stage of the process all process streams for the NGL extraction (the part up to the deethaniser) were equipped with heat exchangers to control the temperature. At first, the process was roughly optimised. At this point the duties of the various exchangers were compared and temperatures matched to eliminate the need for external refrigeration.

In the first part complete heat integration was realised in a coldbox. The optimal conditions for integrated outlet temperatures were established and the need for external refrigeration was eliminated. This can be seen in the composite heat curves in appendix 7-2. This appendix also shows the pinch temperatures of 257 K hot and 250 K cold.

In the remaining two parts all temperatures are above the pinch temperature.

All heat is added in the reboilers. There are no streams that need to be condensed at temperatures above any of the temperatures of the reboilers. Since heat transfer to a reboiler using only sensible heat (e.g. partially heating reboiler E02 with hot sales gas) will lead to huge pieces of equipment, there is no room for heat integration on the hot side.

All cooling is performed in two condensers and four coolers, operating at temperatures above 310 K, with reasonably large duties. There are only several streams in this section that could be heated (<45> and <55>, the feed streams to the depropaniser and the deethaniser.) The sales gas cooler and both condensers need more cooling than can be given by <45> and <55>. Secondly, the condensers operate at temperatures too low to make use of <45> or <55> for cooling. Also the C_3- and C_4-product coolers could not be cooled by these streams, because they operate at temperatures below that of <45> and <55>. Only the sales gas cooler and C_5+-product coolers could be partially cooled by stream <45> or <55>. Because this would not be

complete and would mean extra heat exchangers or complicated multistream exchangers, this has not been further evaluated.
Therefore all heat is supplied by LP-steam and all cooling is achieved by cooling water in the NGL fractionation part and in the sales gas treating part. This will be further discussed in chapter 5-4.

5.1.4 Start up, shut down and turndown

Start up and shut down
During start up special care has to be taken to get the molsieve beds into the correct adsorption and regeneration cycle. If start up takes longer than 10 hours, dry, warm gas will be needed. It may be necessary to expand the feed to a lower pressure to generate more cold during the start up.
During shut down the sales gas compressor will have a higher duty, due to the absence of compression by the turbo-expanders.
Special (dynamic) control will be necessary during start up and shut down. This has not been taken into consideration in the design. The most important element in the start up and shut down will be knowledgeable and experienced personnel.

Turn down
In case of lower feedrates the process can be turned down. A ration of 50 % is taken as maximum turndown.
For the molecular sieve beds it means less flow and therefore lower superficial velocities. This results in longer adsorption times, or shorter regeneration times. A higher pressure drop will result (0.65 bar max.) from this, but its influence is small.
For turbo-expanders and compressors turndown is no problem. Their operational range is 0-125 % (appendix 9-3) of the design flow.
Assuming all flows decrease by the same factor the cold box will only suffer during turndown, but will operate with no problems in the next steady-state. Because there are several cold box-units operating in parallel it is possible to shut down one or two.
Heat exchangers can operate with smaller velocities and lower duties, without any problems.
Separators are not significantly influenced by turndown, except for operating more efficient, due to lower vapour velocities.
Pumps and compressors can run at lower duties. They can operate within the range of 0-125 % of the design.
The difficulties arise from smaller flows through the distillation columns and the absorber. Therefore valve trays have been chosen for use as column internals. With smaller flow part of the valves will close to keep approximately the same flow.
Reboiler and condenser duties could decrease, e.g. by not using parallel units, and reflux ratios could be adjusted. The reflux to the deethaniser can be adjusted by flow control in <19>.

5.2 Process flow scheme

The process flow scheme can be found in appendix 1-1 and is inserted in this report for convenience. The process feed , <01>, is directed into the battery of molecular sieve beds (S01, S02, and S03). Two of these beds will be in use at a given time, while one is being regenerated by stream <43>. Therefore the feed is split in two equal streams to be fed to the beds. The temperature and pressure of regeneration stream <43> are not defined, because they vary with time, due to the regeneration cycle.

The dehydrated feed, <02>, is fed to the expander side of turbo-expander 1 (K01) and expanded to 72.5 bara. This is done to cool the feed, condense the heavier components and lower the pressure for the plate fin heat exchanger. Condensing could be accomplished by cooling alone, but the necessary refrigeration has to come from the process to eliminate the need for external refrigeration. This refrigeration comes from the vapour stream from C01, <20>, and the liquid stream <07>. If, however, <20> is fed to the sales gas compressor (K02) at too high a temperature the compression is less efficient, resulting in a higher compressor duty for K02. Expansion to 72.5 bara gives sufficient cooling that the temperature of the sales gas does not increase unnecessarily and is within the range of the plate fin heat exchanger.

The expanded feed from K01, <03>, is cooled in the cold box (E01). This is done by transferring heat to the expanded liquid from V01, <07>. Extra cooling is provided by the vapour from C01, <20>. This cold box also acts as condenser for the vapour from the deethaniser (C02), <23>. The liquids from V02, <12>, and the vapour from C01, <20>, supply the extra cooling needed. The dotted lines show the streams within the cold box.

The cooled feed, <04>, is partially liquid. In L/V separator 1 (V01) <04> is separated in vapour, <05>, and liquid, <06>. The pressure of <06> is reduced to 30 bara, by using a Joules-Thompson valve, <07>. This causes a decrease in temperature. Due to this lower temperature, <07> can be used for cooling the feed <03> in the cold box (E01).

The vapour from V01, <05>, is expanded in turbo-expander 2 (K03) to 30 bara. At this pressure separation in the deethaniser (C02) is possible, but the pressure is high enough as not to cause excessive duties in the sales gas compressor (K02).

The expanded stream from K03, <09>, is partially liquid. This is separated in L/V separator 2 (V02). The liquid from V02, <11>, is fed to the cold box (E01) to condense the vapour from C01, <23>. The vapour, <10>, is mixed with the vapour from L/V separator 3 (V03), <14>, and fed to absorber C01, <17>.

L/V separator 3, V03, receives vapour, <23>, from the deethaniser (C02), after being partially condensed in the cold box (E01), <26>. The vapour from V03, <14>, is mixed with the vapour from V02, <10>, and fed to absorber C01, <17>. The liquid from V03, <15>, is split 0.47/0.53 into streams <18> and <19>. Stream <18> is fed to the absorber as wash liquid. Stream <19> is fed to the deethaniser (C02) as reflux.

The absorber (C01) is fed by liquid stream <18> at the top and vapour stream <17> at the bottom. The vapour stream, <20>, constitutes the sales gas and is fed to cold box E01 to provide part of the necessary refrigeration. The liquid bottoms, <21>, are fed to the deethaniser (C02).

The deethaniser (C02) is fed by four streams, <19>, <22>, <13> and <08> at theoretical trays 1, 8, 12 and 16. The vapour topproduct, <23>, is sent to the cold box (E01) and fed to L/V separator V03, <26>. The bottoms, <44>, are fed to the depropaniser (C03), after being depressurised to 16 bara, <45>. This pressure level was chosen to have a sufficiently high temperature in the top of the depropaniser (C03) to be able to cool with surface water. The reboiler E02 supplies the heat necessary for the separation in deethaniser C02.

The sales gas stream, <24>, exiting the cold box (E01) is split and fed to the two turbo-expanders (K01 and K03). The split is 0.60/0.40 for streams <27> and <29>. During the cooling cycle there will also be gas flowing through <25> back to the molecular sieve beds, <43>. Streams <27> and <29> are combined to <31>, which is compressed in sales gas compressor (K02). Part of the resulting sales gas, <32>, is drawn off, <35>, to be used as dry, hot heating stream for the molecular sieve beds and possibly as fuel. The rest, <33>, is cooled in the sales gas cooler (E05), <34> and is the sales gas product.

The sales gas used for heating, <35>, can be split to use part of it, <36>, for fuel, after reducing the pressure, <38>. The rest, <37>, is heated in the furnace/stack (F01), after the pressure has been let down, <39>. The resulting hot stream, <42>/<43>, is fed to the molecular sieve bed being regenerated and picks up the water from the bed, <41>. This stream, <40>, is burned in the furnace/stack (F01), along with any fuelgas, <38>. If the molecular sieve bed being regenerated is dry, it has to be cooled. This is done by stream <25>/<43>, which is heated in the process to <41>. This stream is also burned. This means furnace F01 operates as fired heater and stack.

The depropaniser (C03) is fed by <45>. The vapours, <46>, are condensed in condenser E03, <48>, and stored in reflux accumulator V04. Stream <49>, exiting the accumulator is split into reflux, <50>, and low pressure, hot C3-product, <52>. The required reflux ratio is 1.988. The low pressure, hot C3-product, <52>, is raised in pressure by pump P05, <53>, and cooled in E09 to C3-product specification, <54>. The bottoms, <47>, of the depropaniser (C03) are let down in pressure to 9 bara, <55>, and fed to the debutaniser (C04). This is the pressure required for the C5+-product. Reboiler E04 supplies the heat for separation in the depropaniser (C03).

The vapours, <56>, of the debutaniser (C04) are condensed in condenser E06, <58>, and stored in reflux accumulator V03. Stream <59>, exiting the accumulator, is split into reflux, <62>, and hot C_4-product, <60>. The reflux ration is 1.956. The hot C_4-product, <60>, is cooled in E10 to product specification, <61>. The bottoms from the debutaniser (C04), <57>, are cooled in E08, <63>, and let down in pressure to C5+-product specification, <64>. Reboiler E07 supplies the heat for separation in the debutaniser (C04).

Pumps P01 through P04 and P06 are used solely to correct for pressure drop and supply the pressure to raise the liquids to the various pieces of equipment. The bottoms of V01, C02, C03 and C04 do not require pumps as there is pressure drop available to cause the liquids to flow.

In the process flow scheme given in appendix 1-1, no pressure drops are shown for ease of reading. For true pressures, the reader is referred to appendix 5-2, process stream summary.

5.3 Process stream summary

The process stream summary is given in appendix 5-2 and the calculation of the process feed is given in appendix 8-1. Table 5.5 shows the total mass balance.

Table 5.5: Overall Component Mass Balance & Stream Heat balance

Stream	No. :	1		34+40+54+61+64			
	Name:	Total Plant, In		Total Plant, Out		Total Plant, Out-In	
Comp.	Mol.wt	kg/s	Kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.78	0.0277	0.78	0.0277	0.00	0.0000
CH4	16.04	90.59	5.6479	90.59	5.6479	0.00	0.0000
C2H6	30.07	17.16	0.5707	17.16	0.5707	0.00	0.0000
CO2	44.01	10.27	0.2333	10.27	0.2334	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	11.41	0.2586	11.41	0.2586	0.00	0.0000
I-C4H10	58.12	2.54	0.0437	2.54	0.0437	0.00	0.0000
N-C4H10	58.12	4.55	0.0784	4.55	0.0784	0.00	0.0000
I-C5H12	72.15	1.63	0.0225	1.63	0.0225	0.00	0.0000
N-C5H12	72.15	1.65	0.0229	1.65	0.0229	0.00	0.0000
N-C6H14	86.18	1.46	0.0170	1.46	0.0170	0.00	0.0000
N-C7H16	100.20	0.83	0.0083	0.83	0.0083	0.00	0.0000
H2O	18.02	0.04	0.0024	0.04	0.0024	0.00	0.0000
N-C8H18	114.23	0.24	0.0021	0.24	0.0021	0.00	0.0000
Total		143.16	6.9357	143.16	6.9357	0.00	0.0000
Enthalpy	MW	-648		-635		13	

It can be seen that the design is consistent. Also it can be seen that a total of 13 MW is added to the streams. This amount is the sum of all heat removed in the condensers and coolers subtracted from the heat added by reboilers, the sales gas compressor and the heat evolved by burning the regeneration stream. In appendix 5-3 the heat and mass balances can be found.

5.4 Utilities

Appendix 5-4 gives the summary of the utilities used.
Heat is normally fed to a process in the form of low-, medium- or high-pressure steam (LP-, MP- and HP-steam). Because the reboilers (E02, E04 and E07) are operated at 378, 390 and 401 K respectively the heat can be supplied by LP-steam. This has a temperature of 423 K (appendix 5-6).
Reboilers E02, E04 and E07 use respectively 55, 27 and 18 % of the total heat fed to the process. It is clearly shown here that reducing the duty of reboiler E02 has great influence on the process economics.
MP- and HP-steam could be used to decrease the size of the reboilers, but it is assumed that the operating costs are most important and therefore LP-steam is chosen to supply heat to the reboilers.
Cooling in a process is normally achieved by air- or water-cooling, if possible. Only at low temperatures a refrigeration loop should be installed. It is assumed that a refrigeration loop is expensive to operate and not necessary for the temperatures required . It is usually economical to apply air-cooling when process temperatures

exceed 338 K [36, p. 689] this is therefore not applicable to this process. At temperatures up to 323 K it is most economical to use cooling water [36, p. 689]. The condensers (E02 and E06) operate at 319 and 343 K and the coolers (E05, E08, E09 and E10) at 322, 310, 310 and 310 K at the outlet respectively. Cooling water is available at 303 K and can be discarded at a maximum of 311 K. Therefore all cooling can be accomplished by cooling water. In this text, cooling water means surface water, not chilled water, as chilled water is assumed to be too expensive. Condensers E02, E06 and coolers E05, E08, E09 and E10 account for 23, 14, 58, 3, 1 and 1% of the total heat removed from the process respectively. This shows that the effect of amount of heat removed in the sales gas cooler has a great influence on the process economics. This is directly coupled to the duty of the sales gas compressor, as it causes the temperature rise of the sales gas.

Given the allowable temperature rise of the cooling water, the condenser E02 and sales gas cooler E05 use large streams of cooling water (1.1 and 2.8 ton/h respectively) and are actually several heat exchangers in parallel (3 and 2 parallel units respectively). It could be economical to make condenser E06 an air-cooled condenser, due to the high temperature, but that would have to follow from a detailed economical evaluation. This is beyond the scope of this report. Also sales gas cooler E05, C5+ cooler E08 and C4 cooler E10 could be part air cooler, part water cooler, or partially cooled by stream <45> and <55>. It might even be economical to install a separate loop to produce cooling water, in which cooling water of a higher temperature could be allowed. Even returning the water at 313 K reduces the amount of cooling water by ~20 %. Once again this would require a detailed economic evaluation beyond the scope of this report.

The electricity requirements in the process are completely dominated by the duty of the sales gas compressor K02. This compressor accounts for practically 100 % of the electricity usage. The six pumps in the process combined only use 52 kW.

Given the large electricity consumption in the process the sales gas compressor K02 is the place to achieve cost reductions. The usage of the six pumps is negligible and not a place to look for cost reduction.

It is therefore important not to feed the compressor with gas that is too hot. This will cause the temperature to rise too far, due to a larger volume at higher temperature. Furthermore, it is important to match the streams to the turbocompressors with their respective duties, so their outlet pressures are equal. Use of the sales gas compressor can not be avoided, as the sales gas has to be discharged at 190 bara. Only operating the process at higher pressure and using more efficient turbo-expanders/compressors reduces the duty of the sales gas compressor.

5.5 Process yields

The process yields, with simplified block scheme can be found in appendix 2-2. It can be seen that a large amount of electricity and cooling water is used. The reasons for this are given above and in appendix 5-5.

The overall C3-recovery is 97 %. This is high compared to other similar plant recoveries [37]. The C3-product has a flow rate of 80 sm^3/h. Compared to the total propane produced in the US in 1985 [38], this is 6.5 % and compared to the total propane production outside the US, it is 3.4 %. The production of propane by this plant is 2.3 % of the total production of propane in the world. (This is under the assumption that the abbreviation prop. denotes propane and its units are given under the same standard conditions as the calculated values.) The total propane production is produced in 1323 plants world-wide [38].

The overall C_4-recovery is 99 %, with a flow rate of 44 sm^3/h. This is 6.1 % of the total production of C_4 in the US and 3.0 % of the C_4 production in the world (in 1985) [38].

The overall C_5^+-recovery is 98 %, with a flow rate of 32 sm^3/h. This can not be compared to US or world production, as C_5^+ is not as defined as the other streams.

These amounts appear to be quite large and would lead to a huge impact on the market if less than 0.1 % of the plants has a market share of over 6 %. It should however be kept in mind, that these figures are given for 1985. In comparison, the production of NGL increased by 21 % between 1989 and 1998 [39]. Therefore this production level, though still quite large, should not have such a drastic effect on the market situation.

According to [6] there are 12 plants using the turbo-expander process in operation, with capacities up to 400 MMscfd and propane recoveries of 95 %. The process presented in this report is slightly better in performance and has a higher capacity (around 500 MMscfd).

6. Process control

In the process the following type of controllers are used:

- Flow control.
- Pressure control.
- Quality control.
- Temperature control.
- Level control.

The incoming gas stream, <01>, is flow controlled, so the specified turndown ratio can be accomplished and to prevent a feed stream larger than the plant can handle. The feed gas is led through the molecular sieve beds (S01, S02 and S03). These beds are quality controlled by a hygrometer to make sure that the water concentration in the stream, <02>, is low enough to prevent freezing of water in the process. If the concentration of water is too high, the molecular sieve bed in use is shutdown and another molecular sieve is used. The quality control will also start the regeneration cycle.

The turbo-expanders (K01 and K03) have pressure controls on the expander-side to be able to adjust the outlet pressure of the expander. The outlet pressure of the expander determines the temperature of the outlet stream, which is very important for the cryogenic cooling in the onstream process. Also, in case of an expander trip, the pressure control will register the trip and a Joule-Thompson valve will be used to make it possible to keep the process running.

All vessels or columns containing liquid have a level control to prevent gas leaving at the bottom of the vessel. The reflux accumulator vessels (V04 and V05) have a level control, which adjusts the product streams (<52> and <61>). By adjusting these streams, the reflux-control of the distillation column can be maintained at original settings.

The accumulated vapour streams of the liquid-vapour separators (<10> and <14>) are flow controlled to prevent entrainment in the absorber.

The absorber (C01) and deethaniser (C02) are pressure controlled to be able to adjust the process pressure in combination with the turbo-expanders.

The distillation columns (C02, C03 and C04) are temperature controlled to be able to adjust the specifications of the bottom stream leaving the column. The temperature control will adjust the setpoint of the flowcontrol of the reboiler to accomplish a change of column temperature. The setpoint of the temperature control has to be given by the engineers. (A more advanced option is using inline quality controls in the product streams, which can adjust the setpoints of the temperature controls of the columns.)

All reboilers (E02, E04 and E07) are flow controlled to be able to adjust the amount of heat transferred to the distillation column. The setpoint of the flow controller is given by the temperature control of the column.

The depropaniser (C03) and the debutaniser (C04) have a pressure control to make sure all vapour leaving the column is condensed in the condensor. Therefore the pressure control can adjust the cooling water flow. The pressure control can not be used to adjust the top specifications directly.
The reflux streams of the depropaniser and the debutaniser (<51> and <62>) are flow-controlled to be able to adjust the reflux stream. Adjustment of the reflux stream is the only direct way of changing the top specifications of the columns

All product stream heat exchangers (E05, E08, E09 and E10) are temperature controlled to make sure the product streams are cooled to the specified delivery temperature. The temperature control will adjust the amount of cooling water used.

Pressure controls are used at the discharges of the C3 pump (P05) and the sales gas compressor (K02) to ensure the discharge pressure is equal to the specified pressure. In case of a pressure different from specification pressure the pressure control can adjust the duty of the pump or compressor.

The temperature control of the furnace (F01) is placed to make sure the temperature of the gas used to dehydrate the molecular sieves is the right temperature. If the temperature is different of the specified temperature the amount of gas going through the furnace is adjusted.

Detailed control design has to be done by a control expert.

7. Heat and mass balances

Table 7.1 shows the total mass and heat balance for the total process streams.

Table 7.1. Overall component mass balance & stream heat balance

STREAM No. :		1		34+40+54+61+64			
Name :		Total Plant, In		Total Plant, Out		Total Plant, Out-In	
COMP	Mol. wt.	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.78	0.0277	0.78	0.0277	0.00	0.0000
CH4	16.04	90.59	5.6479	90.59	5.6479	0.00	0.0000
C2H6	30.07	17.16	0.5707	17.16	0.5707	0.00	0.0000
CO2	44.01	10.27	0.2333	10.27	0.2334	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	11.41	0.2586	11.41	0.2586	0.00	0.0000
I-C4H10	58.12	2.54	0.0437	2.54	0.0437	0.00	0.0000
N-C4H10	58.12	4.55	0.0784	4.55	0.0784	0.00	0.0000
I-C5H12	72.15	1.63	0.0225	1.63	0.0225	0.00	0.0000
N-C5H12	72.15	1.65	0.0229	1.65	0.0229	0.00	0.0000
N-C6H14	86.18	1.46	0.0170	1.46	0.0170	0.00	0.0000
N-C7H16	100.20	0.83	0.0083	0.83	0.0083	0.00	0.0000
H2O	18.02	0.04	0.0024	0.04	0.0024	0.00	0.0000
N-C8H18	114.23	0.24	0.0021	0.24	0.0021	0.00	0.0000
Total		143.16	6.9357	143.16	6.9357	0.00	0.0000
Enthalpy	MW	-648		-635		13	

It can be seen that it is consistent. Also it can be seen that a total of 13 MW is added to the streams. This amount is the sum of all heat removed in the condensers and coolers subtracted from the heat added by reboilers, the sales gas compressor and the heat evolved by burning the regeneration stream.

This is shown in appendix 5-3, where the heat and mass balances for every piece of equipment can be seen. Here the mass and heat flows in the process are also in balance, at least within 0.1 % error, the chosen tolerance in Chemcad III. Therefore the design is consistent.

8. Process and equipment design

In chapter 5 the equipment was selected and the design criteria were given. In this chapter the process is simulated and the equipment will be sized according to the performance data of the flowsheet calculations. Furthermore, background information about decision points for process and equipment design will be given. Summaries of all used equipment can be found in appendix 3.

8.1 Integration by process simulation

Chemcad III, version 3.3 is used for all process simulations from the dehydrated feed onward. Only simulation of the two turbo-expanders could not be accomplished in one piece of equipment. It had to be simulated by selecting an expander and a compressor and manually supplying the actual power of the expander to its compressor counterpart. The compounds used are well known and their properties are well documented. No problems occurred in using Chemcad III, except for simulation of the reflux heat exchanger and interstage reboilers in the distillation columns. For applications above the critical temperature the Boston-Mathias extrapolation for alpha-function was used. By generating phase envelopes (appendix 7-3) it was checked that retrograde condensation did not occur in the expanders or valves.

8.2 Equipment selection and design

The equipment will be sized based on the stated design criteria. Only results are given in this chapter, detailed calculations are given in appendix 8.

8.2.1 Design of a molecular sieve bed to remove water from the feed

The molecular sieve bed removes all free water (exit concentration 0.1 ppmwt) at entrance temperature and pressure. The design characteristics for the molecular sieve bed are given in the following table:

Table 8.1: Characteristics of molecular sieve 3A

Adsorption capacity [16]	kg/kg	0.113
Porosity of bed	-	0.3
Density of bed [16]	kg/m^3	689
Cost of molsieve [30]	\$/kg	3.18
Desorption heat required [16]	kJ/kg	4187
Heat capacity of the bed [31]	kJ/kgK	1.05

Equipment selection

The most widely used synthetic zeolites are 3A, 4A and 5A, which have openings of about 3, 4 and 5 Å respectively. Due to the small opening in molecular sieve 3A it is the only one that excludes most molecules, except water. The 4A adsorbs methane, ethane and propane and the 5A also absorbs higher hydrocarbons. Though the 3A molecular sieve is more expensive (25 %), it lasts a year longer (4 years instead of 3) and hydrocarbon losses are only 11 % of the losses with a 4A sieve [30].

To reduce the propane and other hydrocarbon loss, while still removing all the free water from the feedstream, the 3A molecular sieve will be chosen.

Material of construction

The feed of the molecular sieve contains water and hydrogen sulphide. Therefore the molecular sieve is placed in a vessel consisting of carbon steel, because it is resistant to water and the low concentrations of hydrogen sulphide, is strong enough and relatively cheap.

Calculation method

The molecular sieve beds have been designed using estimations/equations given by Trent [16,30]. The literature by Trent is not meant as a design manual, but is used due to complete lack of experimental data for the given stream.
The feed gas/regeneration gas properties are calculated by the flowsheeting program Chemcad III. Adsorption is the basis of the equations given by Trent [16,30]. Adsorption takes place in downflow [16]. This is done to ensure that the bed will not be fluidised.

Design criteria

The following criteria have been used for the design of the molecular sieve beds:

- Superficial gas velocity is 90 % of the maximum superficial velocity.
- Length/diameter ratio of is bed >1.
- Maximum pressures drop of 0.5 bar.
- Minimisation of total flows for regeneration.

Calculation

The equilibrium adsorption capacity as given in table 8.1 is estimated from fig. 2 in [16] to be 0.19 kg/kg. In [31], fig. 5-19 it is given that the capacity drops 30 % in the first 200 cycles, but then levels off to last through 2000 cycles. Also a correction of 0.02 kg/kg has to be taken into account for residual water in the sieve. This leads to the assumed adsorption capacity of 0.113 kg/kg.
The porosity of the bed is estimated. The density of the bed is taken to be the average of the density range given in [16] and the rest of the characteristic variables are taken from the literature, except the heat capacity, for which the heat capacity of a 4A molecular sieve [31] is used. The sieve is chosen to be 4-8 mesh beads [16].
The heat necessary for desorption of water is multiplied by 2.5 to correct for heat losses and heating of the bed [16]. The heating and cooling times are chosen to add up to half of the adsorbing time, so continuous operation is possible.

This results in the following operating conditions, with accompanying results (see also appendix 4):

Table 8.2: Operating conditions and results for the molsieve beds

Operating conditions	Unit	Amount
Adsorption time	h	20.00
Heating time	h	7.16
Cooling time	h	2.84

* Gantt-diagram
* Design criteria and constraints ⇒ st451 task.
* Design decision variables.

Results of Calculation		
Adsorption		
V,max	m/s	0.10
V,chosen	m/s	0.09
MTZ	m	0.92
Diameter	m^2	2.53
Bed mass	ton	16.15
Bed length	m	6.65
L/D	-	2.63
Pressure drop	bar	0.44
Water adsorbed	kg	1573.2
Total cost of sieve (per four years)	k$	154.04
Regeneration		
Tgas, in, 110 bar	K	700
Tgas, out, 110 bar	K	380
Tregeneration, bed	K	375
Tcool, final	K	278
Tcool, average	K	327
Tgas, cool, in, 30 bar	K	240
Tgas, cool, out, 30 bar	K	270
flow gas, heating	kg/s	2.43
flow gas, cooling	kg/s	0.96

The heat liberated during adsorption is taken up by the feedgas. Due to the large flow of feedgas and high pressure the temperature in the bed does not increase [31]. Also the temperature of the feedgas does not increase appreciably (<1 K).

Regeneration gasses flow upward through the bed, opposite to the direction of the feedgas being dehydrated. This is done to assure complete regeneration of the end of the bed, so the feedgas is totally dehydrated. Secondly, the superficial velocities during regeneration are smaller and therefore the risk of fluidising the bed is less. The regeneration temperature has been chosen above the atmospheric boiling point of water, but considering the fact that the regeneration gas is dry, this might not be necessary. On the other hand, normal regeneration temperatures are between 450 and 600 K [16]. The regeneration temperature is used to calculate the temperature difference between the bed and the heating stream. Therefore, initially heat will be transferred at a higher rate. During the cooling, the time-averaged temperature is used as the temperature of the bed. The bed will not be of uniform temperature, so initially more heat will be removed though towards the end of the cooling the heat-flux will be less. Due to temperature differences the adsorption might not be uniform throughout the bed. This is another reason for the direction of flow during regeneration to be opposite to the direction of the feedgas being dehydrated. Any loss of efficiency at the entrance of the bed is compensated by increased adsorption at the exit of the bed and the unused part of the bed.

Given the fact that a well-designed bed has a mass transfer zone (MTZ) of about 1 m [16] and the superficial velocity is below any fluidisation velocity, the design is satisfactory. Only the relatively large diameter of the bed is a cause of concern. In comparison, usage of 4 or 5 beds in total leads to diameters of 2.1 and 1.8 m respectively. As stated above, this also gives a reduction of unused bed mass and a reduction in costs. This reduction is significant compared to the costs of extra beds, but the differences are in the order of several k$ per load of sieve, which is negligible. The choice for the amount of beds has been given earlier. The large diameter is caused by a given volumetric flowrate and calculated superficial velocity. Other literature [31] gives velocities up to twice as high as the calculated maximum superficial velocity. This would cause the bed diameter to be smaller by a factor $\sqrt{2}$. Therefore a specialist in the field of molecular sieve design should still be consulted for the design.

Regeneration will be accomplished by two different streams, as can be seen for the heating and cooling steps in the flow scheme appendix 1-2. Heating will take place by using part of the sales gas stream <35>, and cooling will be done by stream <25>. This choice is based on the high pressure and dryness of the sales gas stream and the desire for a more constant feed to the main compressor. It has to be heated to 700 K in a heater, fired with sales gas, the water-rich regeneration or cooling gas.

The cooling stream has been chosen for its low temperature. Cooling can not be achieved by the feedstream, because a temperature increase may lead to changes in coldbox design, compressor duty and water content of the feed to the process. Also, the warmed cooling stream is not returned to the compressors, because of its elevated temperature. The elevated temperature increases the duty of the main compressor and the extra compression costs exceed the value of the "wasted" gas.

A problem, especially in control, is the difference in pressure between the heating and cooling streams (110 and 30 bara). This will require a controlled release of pressure from the bed before the cooling stream enters. Also the bed will have to be brought to the pressure of the feedstream by letting part of the feedstream in while not opening the exit valve. This will not lead to easy control.

Heating of the regeneration stream will be done in a heater fired by the waste stream from regeneration. These streams have net heating values of over 40 MJ/kg, as calculated using Chemcad III. Therefore these streams should be sufficient to heat the stream needed for heating the bed during regeneration. It also means there should be no flow through the heater during the cooling step. If necessary, during for instance the start up of the plant, the sales gas stream can be used in the heater to generate heat.

The bed-life [30] is four years. Other literature [31] gives up to 4000 cycles as bed-life. This would mean the bed life is 9.5 years. Therefore the costs are a worst case. It is also possible to install a spare bed to use while the content of one of the beds is being replaced. This does not give a significant increase in costs (about 50 k$), but allows continuous production in the process, even during sieve replacement. It does require the use of a larger plot size.

8.2.2 Design of a plate-fin heat exchanger (E01)

Plate-fin heat exchangers (PFHE) are a type of compact heat exchangers, consisting of a stack of flat plates (called parting sheets) and corrugations, brazed together as a block. [33]
Plate-fin heat exchangers have the following advantages over other types of heat exchangers [34,35]:

- High thermal effectiveness and very close temperature approaches.(<1K)
- Large heat transfer surface per unit volume.
- Compactness which reduces space and foundation requirements.
- Low weight per unit heat transfer.
- Possibility of heat exchange between many process streams.
- Freedom in design with multi-stream and several possible geometries.
- No problems for two-phase process streams.
- A plate-fin heat exchanger can operate in true counterflow.

The main disadvantages of compact heat exchangers are [35]:

- The PFHE is not necessarily cheaper for a given duty than other heat exchangers, because constructing PFHEs is complex.
- A major constraint of compact heat exchangers is their limited size.
- Little operating experience.
- Aluminium is vulnerable to thermal stress.
- Possible maintenance problems of compact units.

The PFHE can be specified for temperatures ranging from near absolute zero to more than 1073 K, and for pressures up to at least 90 bar.
Compared to a shell and tube heat exchanger (carbon steel), an aluminium PFHE designed for a pressure around 70 bar has the following advantages:[33]

- About 1000 m^2 surface per m^3, compared with around 300 m^2/m^3 for the shell and tube.
- One third of the equipment weight for a given volume.
- Significantly better heat transfer per surface.

Combination of these features gives the PFHE about 25 times more surface per equipment weight, than the shell and tube heat exchanger. In case of a small overall mean temperature difference compared to the temperature range of the streams, the PFHE has the advantage of true counterflow operation, unlike the shell and tube exchanger whose shellside flow is a mixture of cross- and counter-flow.

A plate fin heat exchanger has been selected for this process, because of

- High thermal efficiency required.
- Clean service in the process.
- Multiple streams to heat/cool.
- Condensing/evaporating streams.
- High-pressure operation.
- Low temperature.

Material selection plate-fin heat exchanger.

Common materials used in heat exchangers [33] are stainless steel, nickel alloys, copper and aluminium. Titanium has been proposed, but brazing of the separate titanium parts is very expensive. Some points of interest, concerning the metal are:

- Stainless steels, and most nickel alloys are used for PFHEs in high temperature service.
- Stainless steels have poor conductivity, but their strength allows thinner plates than in aluminium, which offsets some of the reduction in heat transfer. When compared with aluminium, stainless steel has lower fin efficiency.
- Aluminium has a relatively high thermal conductivity and maintains excellent strength and ductility to temperatures as low as 4.3 K. Aluminium looses its strength at temperatures above 373 K.
- Aluminium is susceptible to damage by rough handling, excessive vibration and localised unrelieved stresses.
- Aluminium parts are relatively cheap.

The advantages in weight and thermal conductivity make aluminium a good choice in this design. For cryogenic service, aluminium alloy 3003 is generally used [33].

Corrosion aspects.

Mercury and caustic soda are extremely corrosive, as is pure methanol over a longer period. Wet ammonia is also not suitable for aluminium.
Neither H_2S nor CO_2 corrode aluminium in any concentration [33,45,35], this makes aluminium a suitable material for sour natural gas streams (provided that they are dry). Aluminium and its alloys are readily attacked by hydrochloric acid. Degreasing agents such as methylene chloride, trichloroethylene, etc. are hydrolysed by traces of water and should therefore only be used if an inhibitor is added.
Aluminium and its alloys in electrical contact with more noble metals, such as copper, will rapidly corrode. Electrical insulation should therefore be provided between aluminium and other metals used in the system.

Calculation method

The PFHE has been rated, following the rating example 6.4 of [33].
The method followed is especially made for the design of PHFE. The stream properties of the process streams are within the limits described in the rating example

Design criteria

The design of the heat exchanger is based on the following design criteria:

- Complete heat integration of the NGL extracting process. This is possible, because the sales gas stream leaving the absorber has no design temperature. This stream is used for removal of excess process heat.
- The temperature difference between cold and hot streams must not be less than 5 K.
- The maximum pressure drop for each stream is 0.3 bar.
- Heat loss to the environment has been accounted for, by introducing an efficiency factor of 0.9 in the total heat flow.
- Due to the high gas flows there is negligible effect from longitudinal heat conduction. [33]

Calculation

To calculate the dimensions of the coldbox, various stream properties are needed. Physical properties have been taken from Chemcad III at the zone temperatures. The other relevant properties are calculated per stream, per zone in appendix 8-4

In appendix 8-3 the stream properties from Chemcad III are given in process data sheets. In table 8.3 some basic stream properties are given, together with the design pressures.

Table 8.3: Data input plate-fin heat exchanger

Stream no.,in	T_{in} [K]	T_{out} [K]	Flow [kg/s]	ΔP_{design} [kPa]
<3>	257.1	245.0	143.11	26
<23>	240.2	220.0	39.99	23
<12>	207.5	235.0	19.93	26
<20>	206.1	240.3	118.91	26
<7>	223.7	250.0	28.06	28

The very large process streams are a major disadvantage for the use of a PFHE, this is solved by using four units in parallel. The unit is calculated as one apparatus. Also the existence of six zones consisting of four streams is a difficult problem, because the example is based on two zones with only three streams.

In appendix 8-4, the steps 1-17 are executed following the rating example as accurate as possible. The results are given in table 8.4 (see also appendix 4). In table 8.4, N,tot equals the total amount of layers and dP <stream> is the actual pressure drop for each stream per zone. A,m and A,b are the metal cross section and total cross section respectively. From these data the prosity p,b has been calculated. These calculations only include rough design features, stacking arrangement and rating the geometry have not been considered. The results are therefore not sufficient to actually size and determine the geometry of the plate-fin heat exchanger. Calculated data have to be sent to a manufacturer for detailed design.

Table 8.4: Results PFHE design

		Zone 1	Zone 2	Zone 3	Zone 4	Zone 5	Zone 6	**Total**
N,tot	[-]	171	664	273	400	326	88	
H	m	0.407	1.888	2.843	4.084	3.287	0.809	4.084
W	m	0.6	0.6	0.6	0.6	0.6	0.6	0.6
L	m	0.404	5.022	1.139	3.120	2.064	2.396	14.15
dP <20>	Pa	2996	3113	5994	6923	8286	-	27312
<23>	Pa	7042	6152	9994	-	-	-	23187
<12>	Pa	-	6177	9995	10180	-	-	26353
<7>	Pa	-	-	7060	8039	8214	5744	29057
<3>	Pa	-	-	-	9468	9084	6474	25026
A,m	m^2	0.154	0.598	0.246	0.360	0.294	0.0792	1.73
A,b	m^2	0.244	1.133	1.706	2.451	1.972	0.485	2.451
p,b	[-]	0.631	0.528	0.144	0.147	0.149	0.163	0.706

8.2.3 Design of product coolers

All product streams, leaving the distillation columns or the compressor have to be cooled to meet delivery specifications. The important data of the streams are given in the following table:

Table 8.5: Heat exchangers data

Stream name	Stream no.	Equipment no.	Flow (kg/s)	P (bara)	Tin (K)	Tout (K)	Cooling duty (MW)
Sales gas	<33>	E05	118.3	190	394	322	25.79
C3 product	<53>	E09	11.3	21	319	310	0.33
C4 product	<60>	E10	7.0	9	343	310	0.63
C5+ product	<57>	E08	5.7	8	401	310	1.34

Equipment selection

The following main types of heat exchangers are available ([36], p. 584-598, 691-695, 698-699):

- Shell and tube heat exchanger.
- Plate heat exchanger.
- Air-cooled exchanger.

Air-cooled exchangers are the best choice for minimum process temperatures above 338 K. To meet with delivery specifications, however all product streams have to be cooled below 338 K. Therefore air-cooling is not assumed to be a feasible option.
A plate heat exchanger is hard to clean and can not be used with the available cooling water, which could cause fouling. Chilled water is not used as it is assumed to be too expensive (see chapter 5.4). Therefore the use of a plate heat exchanger is not advisable. Shell and tube heat exchangers will be used with cooling water as coolant.

Materials of construction

All heat exchangers use water as cooling medium. Water is corrosive, so the material of construction has to be resistant to water to prevent corrosion. The sales gas, <30>, also contains hydrogen sulphide, so the construction material for the sales gas heat exchanger has to be able to resist hydrogen sulphide to prevent corrosion.
For all heat exchangers carbon steel is chosen as construction material, because it is resistant to water and hydrogen sulphide, strong enough for existing process conditions and relatively cheap. A disadvantage of carbon steel is the rather poor thermal conductivity.

Calculation method

Kern's method has been used as calculation method. This method is based on experimental work on commercial exchangers with standard tolerances and will give a reasonably satisfactory prediction of the heat-transfer coefficient for standard designs. The prediction of pressure drop is less satisfactory, as pressure drop is more affected by leakage and bypassing than heat-transfer. The calculation method is described in [36], p. 614-631.

Design criteria

The following design criteria have been used for the design of the heat exchangers:

- The minimum temperature difference is 7 K.
- The maximum allowable pressure drop is 0.2 bar at the shell and the tube side.
- The maximum exchanger area is 600 m^2.
- The used shell and tube velocities are in the range given by [36], p. 603-604.

The specifications of the cooling water are given in the following table:

Table 8.6: Specifications cooling water

T_{design}	303 K
T_{max}	311 K
Fouling coefficient	2.0 $kW/m^2.K$

Calculation

For the different product streams, different heat exchanger configurations have been calculated; counter-current flow and "one shell and two tube" passes configuration. For the cooling of streams <53>, <57> and <60> the "one shell and two tube" passes configuration is not a suitable configuration, because the minimum temperature difference will either be smaller than 7 K or more cooling water will be needed. This is not the cheapest option. For cooling of the Sales gas, <33>, both configurations can be used. The configuration with the smallest area has been chosen, as this configuration is the cheapest.

Leading the stream with the highest pressure through the tubes gives the lowest costs. Therefore the product streams are sent through the tubes. Because the cooling water can cause fouling, a square tube pitch is chosen to reduce the chance of obstructions of the shell.

The cooling of the sales gas, <30>, requires an exchanger area larger than the maximum allowable area, therefore two heat exchangers are used in parallel.

The diameters of the pipes have been varied, using standard diameters, until the design met with all design criteria. The main reason for changing the pipe diameters was to meet tube and shell velocities and pressure drop restrictions. The calculations in appendix 8-5 follow the method as described in [36], p. 623.

The results of the calculations are given in table 8.7.

Table 8.7: Results heat exchanger calculation

Configuration		**Sales gas**[1]	**C3 product**	**C4 product**	**C5+ product**
Counter-current	U ($W/m^2.K$)	620	550	570	550
	ΔTm (K)	43.4	7.6	16.4	32.4
	A (m^2)	479	77	66	75
One shell and two tube passes	U ($W/m^2.K$)	625	-	-	-
	ΔTm (K)	40.8	-	-	-
	A (m^2)	505	-	-	-

[1] Area for one of the two heat exchangers.

The counter-current configuration will be chosen for all heat exchangers.

For detailed design specifications see specification sheets in appendix 4.

8.2.4 Design of turbo-expanders

Two turbo-expanders are used in the process. The most important design data are given in table 8.8.

Table 8.8: Design data turbo-expanders

Turbo-expander	Equipment no.	Side	Stream no.	P_{in} (bara)	T_{in} (K)	P_{uit} (bara)
1	K01	Compressor	<29>	30.0	240	[1]
		Expander	<02>	110.0	278	72.5
2	K03	Compressor	<27>	30.0	240	[1]
		Expander	<05>	72.5	245	30.0

[1] Pressure not specified. Depends on expander side conditions.

Equipment selection

For expansion purpose two types of expander can be used ([40]):

- Reciprocating expansion engines
- Turbo-expanders

Reciprocating expanders have been used for 50 years or more, and are still used to some extent. Their problems are high maintenance, excessive size, valve problems and the fact that liquid will damage the valves. The turboexpanders have high efficiency, low maintenance, reduced size, are able to handle liquid and generate low temperatures.

In the process turbo-expanders are used, because the low temperatures are needed to accomplish the required separation and the high efficiency, so utility costs can be reduced. They also lower pressure for the coldbox. The turbo-expanders are linked with a compressor to make maximum use of the energy generated by the expander.

Materials of construction

The materials of construction for the turbo-expander-compressor system have been specified by Mafi-Trench; standard materials have been used. The most important parts of the turbo-expander are made of:

Expander housing	:	Stainless steel
Compressor housing	:	Carbon steel
Shaft	:	Stainless steel
Expander wheel	:	Aluminium alloy
Compressor wheel	:	Aluminium alloy

For a more specific description see appendix 9-3.

Design criteria

The following design criteria have been used:

- The maximum allowable suction volume is 10,000 m^3/h. (appendix 9-2)
- The maximum power generated is 7500 kW. (appendix 9-2)
- The maximum allowable liquid formation is 25% weight. (appendix 9-2)

Calculation

The compressors of the turbo-expanders are used in parallel to meet with the design criteria (maximum suction volume). Using Chemcad the duty of the expander has been calculated, using an efficiency of 85% (appendix 9-2 and 9-3). The compressor duty was matched with the expander duty using an efficiency of 75% (appendix 9-3). The results are given in table 8.9.

Table 8.9: Results turbo-expander design

Turbo-expander	Equipment no.	Side	η (%)	Duty (kW)	P_{out} (bara)	T_{out} (K)
1	K01	Compressor	75	3363	52.4	284.5
		Expander	85	3363	72.5	257.1
2	K03	Compressor	75	5252	52.4	284.5
		Expander	85	5252	30.0	207.5

More design data are given in appendix 4.

Mafi-Trench also designed the turbo-expander, using the stream summaries from the process. Their calculations give other results than Chemcad. The liquid weight fraction is larger and the discharge temperatures of the expanders are lower. And according to Mafi-Trench the two compressors should be placed in series, because placing the compressors in series allows a better match between expander and compressor impellers. From this fact it follows that the maximum suction volume, which was assumed, proves to be too low. The specifications for the compressors in series are given in appendix 9-3. Implementation of the new configuration was not possible due to lack of time and conflicting inputs between Chemcad and Mafi-Trench.

8.2.5 Design of the compressor

One compressor is used to compress the sales gas to the specified delivery pressure. The important design data are given in table 8.10.

Table 8.10: Compressor data

Compressor:	Equipment no.	Stream no.	Capacity (m^3/h)	P_{in} (bara)	P_{uit} (bara)	T_{in} (K)
Sales gas	K02	<31>	8941	52.2	190.0	284.8

Equipment selection

There are two main types of compressors used in the industry:

- Reciprocating compressor
- Centrifugal pump

The required capacity in combination with the necessary discharge pressure, is higher than a single reciprocating compressor can handle ([36], p. 432). Therefore a centrifugal compressor is used. The compressor is a single stage compressor as the compression factor is 3.6.

Materials of construction

Using [41] p. 80 the following main materials of construction are used:

- Shaft : heat-treated and stress-relieved forged alloy steel
- Casing : steel
- Rotor : low alloy steel

All are used for strength and tear resistance. More specific material choices must be done in co-operation with the compressor manufacturer.

Design criteria

The following design criteria have been used designing the compressor:

- the maximum allowable suction capacity is 340,000 m3/h. ([41], p. 81)

Calculation

The duty of the compressor has been calculated using Chemcad, using a efficiency of 85 % ([32]). No further design has taken place. The results are given in table 8.11.

Table 8.11: Results compressor design

Compressor:	Equipment no.	Duty (MW)	T_{out} (K)
Sales gas	K02	24.1	394.3

More design data are given in appendix 4.

8.2.6 Design of pumps

In the process six pumps are necessary to "keep the liquid moving" and elevate the pressure (to overcome pressure drop and/or higher-pressure operation in the following unit(s)). The important input data are given in table 8.12:

Table 8.12: Input data pumps

Pump:	Equipment no.	Pumped liquid	Capacity (m^3/s)	P_{in} (bara)	T_{in} (K)
Bottom V02	P01	<12>	562	30.0	207.5
Bottom V03	P02	<16>	204	29.7	220.0
Bottom C01	P03	<22>	54	30.2	208.7
Reflux C03	P04	<51>	179	16.0	318.8
C3 Pump	P05	<53>	90	16.0	318.8
Reflux C04	P06	<60>,<62>	148	9.0	343.0

Equipment selection

A single-stage centrifugal pump is used, because there are no special requirements for the pump. The advantages of a centrifugal pump are simplicity, low first cost, uniform (non-pulsating) flow, small floor space, low maintenance expense, quiet operation and adaptability for use with a motor or turbine drive ([42], p. 6-7). The required capacities are all in the operating range of a centrifugal pump.

Material selection

In the chemical industry, the selection of pump materials of construction is dictated by considerations of corrosion, erosion, personnel safety and liquid contamination. The experience of pump manufacturers is often valuable in selecting materials.
The materials of construction are mild steel (MS) and high tensile steel (HT Steel). Mild steel is used for the pump house and high tensile steel is used for the pump rotor and the shaft.

Design method

The design method is given in [36], p. 434-435

Design Criteria

The following design criteria have been used for the design of the pumps ([36], p. 434):

- Maximum capacity of a centrifugal pump is 1000 m^3/h.
- Minimum capacity of a centrifugal pump is 0.25 m^3/h.

Calculation

For calculation of the required power [36], p. 434-435 is used.
The calculations are given in appendix 8-6.

The results are given in table 8.13.

Table 8.13: Results pump design

Pump:	Equipment no.	P_s (bara)	P_d (bara)	Capacity (m^3/h)	η_p (%)	Power (kW)
Bottom V02	P01	30.0	30.6	562	83	11.8
Bottom V03	P02	29.7	30.8	204	80	7.6
Bottom C01	P03	30.2	30.7	54	75	1.0
Reflux C03	P04	16.0	17.5	179	80	9.2
C3 Pump	P05	16.0	21.0	90	76	16.4
Reflux C04	P06	9.0	10.2	148	78	6.3

For more specific design data see appendix 4.
If the $NPSH_{Available}$ is smaller than the $NPSH_{Required}$, it is an option to lower the pump.

8.2.7 Design of liquid-vapour separators

Three liquid-vapour separators are used.
The input data for the liquid-vapour separators are given in table 8.14:

Table 8.14:Input data liquid-vapour separators

L/V separator:	Equipment no.	Stream no.	P (bara)	T (K)
1	V01	<04>	72.5	245.0
2	V02	<09>	30	207.5
3	V03	<26>	30	220.0

Equipment selection

For the liquid-vapour separation several types of separators can be used. The most important types of liquid-vapour separators are gravity separators, centrifugal separators, impingement separators (e.g. vane-type) and filter separators
An impingement separator is used, because this type of separator is relatively cheap and has good separation efficiencies. A vane-type demister is used. The disadvantage of using a vane-type demister is that the maximum separation efficiency is lower then that of a knitted wire mesh. The advantage is that the maximum velocity is higher, so the column has a smaller diameter and is cheaper. Apart from that a vane-type demister has a lower pressure drop [43].

Material selection

All feeds of the liquid-vapour separators contain the full spectrum of hydrocarbons, nitrogen and hydrogen sulphide. Hydrogen sulphide is corrosive, so the material of construction has to be able to resist the prevalent hydrogen sulphide concentrations to prevent corrosion. For all liquid-vapour separators carbon steel is chosen as construction material, because it is able to resist hydrogen sulphide, strong enough and relatively cheap.

Calculation method

For designing the liquid-vapour separators three references have been used ([44], p. 154-155, [45], p 144-149, [46]). A description of the method used is given in appendix 8-7.

Design criteria

The following design criteria have been used for the design of the liquid-vapour separators:

- A liquid hold-up time of five minutes based on [46].
- A vapour disengaging height of 1.5 times the column diameter ([44], p. 156).

Calculation

The calculation method is described in appendix 8-7. The calculations are given in appendix 8-7. For V03 the superficial gas velocity is chosen lower than the maximum allowable gas velocity, because the height-diameter ratio would be very large using the maximum gas velocity. And a vessel with a large height-diameter ratio is very expensive.
The results of the design are given in table 8.15:

Table 8.15 : Results liquid-vapour separators design

L/V separator	Equipment no.	$U_{g,max}$ (m/s)	$U_{g,design}$ (m/s)	D (m)	H (m)	V (m^3)
1	V01	0.16	0.16	2.84	8.16	52
2	V02	0.33	0.33	2.97	6.80	47
3	V03	031	0.20	1.65	10.40	22

For detailed design information see the specification sheets in appendix 4.

8.2.8 Design of distillation columns and absorber

The process requires design of three distillation columns and one absorber.

Equipment selection

The selection to be made for the distillation columns mainly consists of the choice of the internals. Internals can be divided into two categories: trays and packing. Taking into account the high pressure at which the separation takes place and the high liquid flow-rates, it is clear that trays should be used in this separation. When using packing under these conditions the packing would be flooded and compressed and therefore operate less efficient [47]. The liquid would run straight down the packing, rendering it almost ineffective. Making the column diameter larger could prevent this. In that case, the cross-section of the packing is also enlarged, decreasing the amount of liquid per unit of square packing. However, this would lead to a column diameter larger than the one for a tray column. Also using multiple feedstreams significantly increases the height of the packed beds, due to extra liquid distributors. Therefore trays are the preferred internals for the separations in this process.

Trays can be divided into different types for different uses, the main types being [48]:

- Bubble-cap trays
- Valve trays
- Sieve trays

One of the considerations taken into account in selecting trays is the turndown of 50%. Bubble-cap trays are excellent trays for this purpose. One disadvantage however of this type of tray is the relatively high pressure drop, though this may vary with design. Valve trays perform almost as well as bubble-cap trays with respect to turndown, while suffering less from pressure drop. Sieve trays are not suited for high turndown, while their pressure drop equals that of valve trays. Efficiencies of these tray types do not differ significantly [36, p. 504]. Bubble-caps are more expensive than valve trays, which on their turn are more expensive than sieve trays [36, p. 514]. Therefore valve trays are used.

Material of construction

All flows through the columns are relatively clean and contain only traces of hydrogen sulphide. For all columns carbon steel is chosen as construction material, because it is resistant to hydrogen sulphide in low concentrations, it is strong enough and relatively cheap. It also has a low thermal conductivity. This material is also selected for the trays.

Calculation method

The valve tray design was done using [36, p. 535].
The column was sized using the Souders-Brown method, as given in [49]. This method gives conservative results, as it is based on no liquid entrainment between trays. In practice some entrainment can be tolerated at negligible loss in efficiency or capacity [48].
In order to convert the theoretically found number of plates to the actual needed number of plates, the plate efficiency is required. Although experimentally determined efficiencies are available in literature, these do not cover all the separations performed in the process. Therefore, a correlation for plate efficiency was used [36, p. 503].
For the absorber an alternative calculation method was used [36, p. 504].

Design criteria

The design of the columns is based on the following criteria:

- Separation sharpness.
- Maximum allowable impurity concentrations in product streams.
- Turndown of 50 %.
- Maximum recovery of valuable components.

The specifications of the product streams are given in table 8.16.

Table 8.16: Specifications of product streams

Stream name	Stream number	C2-spec. (wt%)	C3-spec. (wt%)	C4-spec. (wt%)	C5+-spec. (wt%)
C3-product	<54>	<1.5	max.	<1.0	
C4-product	<61>		<0.2	max.	<1.5
C5+-product	<64>			<1.0	max.

Calculation

First of all, it has to be mentioned that all calculations concerning the design have been carried without taking the pressure drop into account. This has been done, because the pressure drop can only be estimated when the number of plates per column is known. The number of plates is a function of the efficiency, for which a detailed calculation is necessary. As a result, the pressure drop is known only in a late stadium of design. However, considering the high pressures used in the process, the pressure drop will only have a very limited influence on the component properties [49].
The method of calculation for the plate efficiencies uses the molar average liquid viscosity and the average relative volatility of the light key. Calculation of the efficiencies is done in appendix 8-8; the results are given in table 8.17.
These calculated efficiencies are quite conservative, considering the fact that the calculated efficiency for the propane/butane separation is about 73%, while the percentage found in literature is 100% [49]. Absorbers have lower efficiencies than distillation columns [36, p. 504].
For the design of the valve trays a few assumptions had to be made. The weir height was chosen to be 75 mm [49, p. 180]. The weir length, although not mentioned in appendix 4, can be estimated at 0.77 times the column diameter [36, p. 527]. The active area was taken to be 75% of the column area [36, p. 526]. The perforated area

was estimated to be 10% of the active area [36, p. 535]. Results are presented in appendix 4.

The Souders-Brown method uses various equations relating the vapour and liquid streams to the diameter of the column. The required values for vapour and liquid streams and surface tension were taken from Chemcad. The detailed calculation is given in appendix 8-8 and some results are shown in table 8.17 (see also appendix 4). The height of the column was calculated using a tray spacing of 0.45 m [36, p. 510] and further using the equation in [49, p. 144].

The pressure drop has been estimated to be 8 mbar per tray [49]. In [49] it is given that pressure drop is not as pronounced for high pressures. The design pressure of the columns specified is the top pressure.

Table 8.17: Results column calculation

Equipment Name	Equipment number	Largest diameter (m)	Height (m)	Tray eff. (%)	Actual no. of trays (-)	Top pressure (bar)	Pressure Drop (bar)	Reflux Ratio [2]
Absorber	C01	5.7	16.4	43	24	30	0.2	-
Deethaniser	C02	6.5	[1] 23.1	69	39	30	0.3	-
Depropaniser	C03	2.9	33.5	73	61	16	0.5	1.97
Debutaniser	C04	2.5	24	76	40	9	0.3	1.93

1) The deethaniser is divided in two parts: one with a diameter of 3.5 m and a total height of 6.2 m and one with a diameter of 6.5 m and a height of 16.9 m.
2) Values taken from Chemcad

8.2.9 Design of the condensers for distillation columns

Two of the three top-streams (<46> and <56>) of the distillation columns have to be totally condensed. The product streams are condensed with cooling water. The important input data for the condensers are given in table 8.18:

Table 8.18: Input data condensers

Condenser:	Equipment no.	Stream no.	Condensation rate (kg/s)	P_{in} (bara)	T_{in} (K)	Heat load (MW)
Depropaniser	E03	<46>	33.77	16	318.75	9.94
Debutaniser	E06	<56>	20.24	9	343.01	6.18

Equipment selection

There are two main types of condensers: direct contact condensers and shell and tube exchangers. In a direct contact condenser the condensing gas is brought into direct contact with a liquid stream. The condensate will flow with the cooling liquid. This is not desirable in the process, because the product specifications allow little amount of impurities from water or other cooling medium. Therefore shell and tube exchangers are used as condensers.

Material selection

All condensers use water as cooling medium. Water is corrosive, so the material of construction has to be able to resist steam to prevent corrosion. For all condensers carbon steel is chosen as construction material, because it resistant to water, strong enough and relatively cheap. A disadvantage of carbon steel is the rather poor thermal conductivity.

Calculation method

For design of the condensers the method described in [36], p. 650-669 is used.

Design criteria

The following design criteria have been used for the design of the condensers:

- The minimum temperature difference is 7 K.
- The maximum allowable pressure drop is 0.2 bar at the shell and the tube side.
- The maximum exchanger area is 600 m^2.
- The shell and tube velocities are in the range given by [36], p. 603-604.

The specifications of the cooling water are given in the following table.

Table 8.19 : Cooling water specifications

T_{design}	303 K
T_{max}	311 K
Fouling coefficient	2.0 $kW/m^2.K$

Calculation

The condensers have been calculated using [36], p. 664-669.
Condensation of the top streams (<46> and <56>) takes place in the shell, because condensation in the tubes leads to larger pressure drops at the shell-side and pressure drops are limited by the design criteria. To reduce the area used by the condensers the condensers are placed vertical. Because the condensers are placed vertical figure 12.43 is used in stead of equation 12.50 in [36], p. 652-653.

The exchanger area required to condense <46> is larger than the maximum exchanger area per unit. Therefore three units are used in parallel. The calculations are given in appendix 8-9.

The most important results are given in table 8.20.

Table 8.20:Results condenser design

Condenser:	Equipment no.	U ($W/m^2.K$)	ΔTm (K)	Area (m^2)
Depropaniser[1]	E03	600	11.1	497
Debutaniser	E06	575	35.7	301

[1] Area given for one of three heat exchangers

For more design information see the specification sheets in appendix 4.

8.2.10 Design of the reboilers of the distillation columns

The process uses three distillation columns for the separation of the product streams. Each of the columns uses a reboiler. The reboilers are heated by LP steam. In the following table the important input-data are given, necessary for the design of the reboilers.

Table 8.21: Input data reboilers

Reboiler	Equipment no.	Vaporisation rate (kg/s)	P (bara)	T_{bottom} (K)	Heat load (MW)
Deethaniser	E02	74.9	30	377.9	16.71
Depropaniser	E04	25.6	16	389.7	8.09
Debutaniser	E07	19.5	9	400.8	5.37

Equipment selection

The following main types of reboilers can be used ([36], p.669-671):

- Forced circulation reboilers
- Thermosyphon reboilers
- Kettle type reboilers

The forced circulation reboilers are especially suitable for handling viscous and heavily fouling process fluids. The advantages of a forced reboiler are predictable and high velocities, suitability for low vacuum operations and suitability for low rates of vaporisation. The major disadvantage of a forced circulation reboiler is that a pump is required and the pumping cost will be high.

Thermosyphon reboilers are the most economical type for most applications, but are not suitable for high viscosity fluids or high vacuum pressure. A disadvantage of this type is that the column base must be elevated to provide the hydrostatic head required for the thermosyphon effect.

Kettle reboilers have lower heat-transfer coefficients than the other two types, as there is no liquid circulation. They are not suitable for fouling materials and have a high residence time. They are suitable for vacuum operation and for high rates of vaporisation.

The fluids, leaving the distillation column have high pressures and rather low viscosity, so a thermosyphon reboiler is the best type of reboiler. To reduce the space used by the reboilers vertical reboilers are chosen.

Material selection

All reboilers use LP steam as heating medium. Steam (water) is corrosive, so the material of construction has to be able to resist steam to prevent corrosion. For all reboilers carbon steel is chosen as construction material, because it is able to resist water, strong enough and relatively cheap. A disadvantage of carbon steel is the rather poor thermal conductivity.

Calculation method

For the design of the reboilers a general correlation of heat-transfer rate with reduced temperature for vertical thermosyphon reboilers is used. This method has been developed by Frank and Pickett [50]. They programmed Fair's rigorous design method for computer solution and used it, together with operating data on commercial exchangers, to derive the correlation. The design method is given in [36], p.681-686.

Design criteria

The following design criteria have been used for the design of the reboilers:

- The minimum temperature difference is 10 K.
- The maximum exchanger area is 600 m^2.
- The maximum heat flux should not exceed 37.9 kW/m^2.

The specifications of the steam used are given in the following table:

Table 8.22: Steam specifications

Category	Low pressure
$T_{condensation}$	423 K
ΔH_{vap}	2200 kJ/kg
Fouling coefficient	10 $kW/m^2.K$

Calculation

For all three reboilers the calculations are given in appendix 8-10. The calculation method used is given in [36], p. 685-686. Using figure 12.29 for designing all reboilers design, heat fluxes are found larger than the maximum heat flux specified in the design criteria. Therefore all reboilers have been calculated with the maximum design heat flux. The shell inside diameter has also been calculated, using pipes with a length of 3.66 m, an inside diameter of 25 mm and an outside diameter of 30 mm. For the vertical thermosyphon reboilers a fixed tube sheet is chosen, with a 1.25 square pitch. The results of the calculations are given in table 8.23:

Table 8.23: Results reboiler calculations

Unit:	Equipment no.	Area (m^2)	Shell inside diameter (m)
Deethaniser	E02	441	1.70
Depropaniser	E04	213	1.22
Debutaniser	E07	142	1.02

For more design information see the specification sheets in appendix 4.

8.2.11 Design of liquid accumulators

The condensed vapour of the condensers (E03 and E06) is collected in a liquid accumulator. By using liquid accumulators it is possible to adjust the reflux rate without influencing the whole process. The input data of the liquid accumulators are given in table 8.24:

Table 8.24: Input data liquid accumulators

Accumulator:	Equipment no.	Feed	P (bara)	T (K)
Reflux C03	V04	<48>	16	318.8
Reflux C04	V05	<58>	9	343.0

Equipment selection

For economy of space, a vertically placed vessel is used as accumulator.

Material selection

All feeds of the liquid accumulators contain the full spectrum of the hydrocarbons in the feed, nitrogen and a trace of hydrogen sulphide. Hydrogen sulphide is corrosive, so the material of construction has to be able to resist hydrogen sulphide to prevent corrosion. For all accumulators carbon steel is chosen as construction material, because it is resistant to hydrogen sulphide, strong enough and relatively cheap.

Design criteria

The following design criteria have been used for the design of the liquid accumulators:

- A liquid hold-up time of 5 minutes [46].
- A liquid-volume of 65% of the total vessel volume [46].

Calculation

Using the flow rate and the hold-up time a liquid volume is calculated. The liquid-volume is 65% of the total accumulator volume and the total volume is calculated. Using [51] a diameter is chosen and the height is calculated.
The vessel is placed vertical to reduce the area used by the vessel. In appendix 8-11 the calculations are given. The results are given in table 8.25.

Table 8.25: Results liquid accumulator design

Accumulator:	Equipment no.	V (m^3)	D (m)	H (m)
Depropaniser	V04	35.0	3.0	4.6
Debutaniser	V05	20.0	2.3	4.5

For more specific design data see appendix 4.

8.2.12 Design of a fired heater/stack

The fired heater, used for the heating the regeneration stream <39> for the molecular sieve bed has a dual purpose. It is also used for burning the waste streams from the molecular sieve beds.
The amount of heat to be transferred is 1 MW. Given the fact that the streams entering the molecular sieve beds have a net heating value of 44 MJ/kg, more than enough heat is available for use in the heater even if an efficiency of 80 % is chosen. These streams have to be burnt, because they contain a large amount of (flammable) hydrocarbons, but are relatively small process stream. Also it has not been examined what the influence of the batch-nature of the regeneration streams is on the heating capacity and temperatures in this piece of equipment.
Design calculations can be made for a fired heater or a flare [44]. Due to the difficulty in designing the dual nature of this equipment (heater and flare) and the fact that it is not a major part of the process it has not been designed.

8.3 Special issues

8.3.1 Expander trip

Due to wear and tear, e.g. fluids in the stream, it is possible that problems can occur with the turbo expanders. It is assumed that the problems will mostly occur on the expander side. An expander trip will be defined here as a blockage in the expander side of the turbo expander. If this happens it will mean no flow and therefore possibly no cooling downstream in the process. To decrease the influence this will have on the process bypass valve have been installed. If the pressure decreases then the valve will be opened. This will decrease the efficiency of the process. Recovery will drop, as the separations in the L/V separators will not be as efficient due to higher temperatures. Also the sales gas compressor will require a much larger duty to bring the sales gas up to specification without the use of the turbo compressors. Due to higher temperatures extra cooling water will be needed. To ensure proper product specifications it may be necessary to increase reboiler duty in the deethaniser.

8.3.2 Variations in feed composition

For the feed it has been assumed that the C_3, C_4, C_5^+ fractions are all alkanes, not alkenes. This has an influence on the optimisation of the process, as can be seen in table 8.26.

Table 8.26: Process indices with alternative feed

Modification	Recovery C3 (%)	K02 duty (MW)	E02 duty (MW)
All C3 = propylene	95.9	25	17.1
All C3 = propylene, all C2 = ethylene	77.9	21.1	10.7

The difference due to variations in propane/propylene content is minimal. If all C_3 consists of propylene and all C_2 of ethylene the process is not running optimally. Judging by the reduced duties of the sales gas compressor (K02) and the C02 reboiler (E02) the recovery could still be raised.

9. Wastes

All processes produce wastes, directly and indirectly. Indirect wastes are beyond the scope of this report. Only the direct wastes are considered in this chapter.

The process produces three waste streams:

- Wastewater from cooling operations.
- Off-gas, produced by the burning of regeneration gas after regenerating the molecular sieves.
- Used 3A zeolite of the molecular sieves.

The streams are considered separately.

9.1 Wastewater

Wastewater from industries includes employees' sanitary wastes, process waters from manufacturing, wash waters and relatively uncontaminated water from heating and cooling operations. In this process wastewater from heating and cooling is the major type of wastewater, other kinds of wastewater can be neglected.

In contrast to gaseous effluents, for which specific emission standards exist for many substances, less emissions standards exists for liquid effluents [52]. In general authorities require to apply at least the "best available technology" (BAT). The final, quantitative emissions standards, which must be met with, are usually decided by negotiations with (local) government authorities.

For some types of water pollutants specific emissions standards exist. These are based on e.g. concentration of organic material in the water.

The produced wastewater can be separated in heating and cooling water.

Wastewater from heating (total amount equals 49.5 t/h at 423 K) is returned to the supplier to be recycled.

Cooling wastewater will be discharged to the surface water. The data of the produced cooling wastewater is given in table 8.1.

Table 8.1: Data wastewater

Amount (tons/s)	1.32
Temperature difference (K)	8.0
Thermal pollution (MW)	44.4

The effect of the thermal pollution will not be very large, but the amount of thermal pollution is rather large. For a plant like Shell Pernis the total thermal pollution of the water is about 200 MW [62]. The specific emission standards are not very important because the cooling water is not contaminated with organic material; a leak of one of the heat exchangers is the only possibility for the water to become polluted with organic materials. The chance of leakage is assumed to be very small.

9.2 Off-gas

Off-gas is produced in this process by the combustion of gas used for regenerating the molecular sieves. The sales gas, used for the regeneration contains mainly methane and ethane. But it also contains some nitrogen, carbon dioxide, propane and heavier hydrocarbons and small traces of hydrogen sulphide. Combustion of the sales gas gives CO, CO_2, NO_x, SO_x and H_2O.

In table 8.2 the emission limits for incinerators in the Netherlands for some of these components are given [52].

Table 8.2: Emission limits

Component	Emission limit
CO	50 mg/m^3
SO_2	40 mg/m^3
NO_x	70 mg/m^3

Currently there are no standards for the emission of CO_2. Because of the relatively small amounts of N_2 and H_2S in the sales gas, NO_x en SO_2 emissions will not exceed the emission limits. Norwegian standards have not been found, but the standards in the Netherlands are assumed to be stricter than the Norwegian standards. Therefore the emissions will be within the emission limits of Norway.

9.3 Used 3A zeolite

In the process there will be a discontinuous zeolite wastestream. Once in four years the molecular sieve beds have to be refreshed. In four years time the total waste of zeolite is almost 50 ton. The molecular sieve only absorbs water, therefore no other compounds will be found in the used zeolite. The used molsieve can therefore be used as landfill. The only costs will be transportation costs; these can be neglected in the economical review.

9.4 Process and equipment choice.

An open cooling water system is chosen in the process. To reduce the amount of thermal water pollution the use of a closed cooling water system, using a cooling tower, is a possibility. Due to lack of time this option has not been evaluated. Depending on an economical evaluation, an eventual choice for a closed cooling water system can be made. The use of a closed cooling water system does not mean that the total amount of thermal pollution is reduced, it is only transferred to the air instead of the surface water.

In the molecular sieves the 3A zeolite is used, because it lasts longer and has lower losses of hydrocarbons than other zeolites. By using the 3A zeolite the amount of waste is being reduced.

Regenerating with part of the sales gas leads to CO, CO_2, NO_x, SO_x and H_2O formation, while other ways of regenerating the molecular sieves may lead to different exhaust gasses. Other ways of regenerating the dehydrator could cause other wastestreams (e.g. liquid hydrocarbon wastes).

10. Process safety

To reduce the safety risks of the process an assessment of the Fire and Explosion Index and a Hazard and Operability study (Hazop) are carried out. The outcome of these methods is used to make recommendations for the safe operation of the process.

10.1 Fire and Explosion Index

The Fire and Explosion Index is determined using the Dow guide [53]. Calculation of this index is shown in appendices 10-1 to 10-5. For this, four pieces of equipment were selected. The criteria for the selection included the size of the equipment, the pressure in and the pressure drop over the equipment, the mass flow through the equipment and interaction between the streams of the equipment. As can be seen in the appendices, the following pieces of equipment were selected: the first turbo-expander (K01), the coldbox (E01), the deethaniser (C02) and the compressor (K02). The turbo-expander was selected, because of the high pressure (110 bara), the high pressure drop (37.5 bar) and the high rotating velocities of the expander and compressor turbines (10.000 rpm). The coldbox was selected because of the large mass flow and the interaction between the streams: an explosion in the coldbox results in the rupture of at least the five feed pipelines of the coldbox. The deethaniser is the largest amongst the vessels and columns. The compressor (K02) is a unit which has a large output pressure (190 bara) and a large increase in pressure (137.6 bar).

Table 10.1: F&E Index and damage factor for the four most critical units

Equipm. Number	Name	F&E Index	Damage Factor
K01	Turbo-expander	132	0.78
E01	Coldbox	117	0.77
C02	Deethaniser	95	0.71
K02	Compressor	113	0.76

As can be seen in table 10.1, the turbo-expander is the unit with the largest Fire and Explosion Index and therefore the largest damage factor. When the Dow Guide is used, it is not possible to calculate the F&EI for a whole process, only for a unit. The turbo-expander is the most dangerous and from that the whole process can be classified as having a heavy degree of hazard ([53], p.28).

Loss control credit factors were determined, with the turbo-expander as the unit under investigation. The results are shown in appendix 10-2, table 10.2.2. From this the Base and Actual Maximum Probable Property Damage can be calculated. These values are given in appendix 10-2, table 10.2.3. The radius of exposure is 34 m. Preferably other equipment should be placed outside this radius, to avoid extra damage when an accident occurs. For calculation of the Base MMPD is it assumed however, that the second turbo-expander is within this radius. The assumption that other equipment is not within this radius is based on [54]. Therefore the costs of two turbo-expanders and their throughput (i.e. the sales gas flow) during 15 minutes are included in this Base MPPD. The 15 minutes are taken to be the average time required to shut the flows to the turbo-expanders down [53].

Protective measures should in include, leak detectors, with an alarm for a natural gas concentration (of e.g. 25%) of the lower explosion limit (LEL) and should also activate a protective or shutdown system at, for example, 70% of the LEL. Also remote control valves are necessary to isolated units which are malfunctioning or which are threatened by fire, explosions, etc.

In case of malfunction or an accident (also depending on the severity of the case), the entire contents of the process should be flared. It should be recognised, that it takes time to flare, but at that the same time it is critical to dispose of the gas as quickly as possible, so that leakage will not last while the flare is burning for e.g. 15 minutes. Therefore the gas should be temporarily stored. This should be done, of course, in a remote dump, which is not to be affected by malfunctioning of other process equipment. Another option is the isolation of the unit, which is malfunctioning by using valves. The piping in the plant should be designed to make this measure possible, i.e. shortcuts must be provided.

Also, the flare must be at a remote enough distance not to ignite a cloud of vapour, which could possibly be hanging over the plant.

The temperature of the components is well below the autoignition temperature (see appendix 10-6, table 10.6.1), so the risk of spontaneous combustion minimal.

Finally it must be noted that if any oxygen (e.g. from the air) enters the process a highly dangerous situation occurs. Although the UEL (see appendix 10-6, table 10.6.1) of the components is not that high, local risk of explosion is very high. It must be stated though, that air can not enter the process because this is operating on overpressure. Leakage of process contents is into the air is much more likely. There is also a risk of rupture of equipment due to high pressures in the process. Equipment must be sufficiently strong to withstand these pressures. If rupture still should occur, the amount of vapour and liquids, which are released, is high. This is also an effect of the high pressure. The mentioned effects are accounted for in the Fire and Explosion Index calculation as well as in the design of the equipment.

10.2 Hazop

After developing a Fire and Explosion Index, it is evident, that the first turbo-expander is the most critical piece of equipment. It is therefore decided to carry out a hazard and operability study on this turbo-expander.

In appendix 10-6, table 10.6.2 possible deviations from normal operation are mentioned and their impact on the turbo-expander and the process are considered.

A few points may need some elaboration. First, if the compressor part of the turbo-expander has no flow, while, although maybe temporarily, the expander part still has feed, the compressor part continues to compress non-existent flow. This is because the compressor shaft is linked with the expander shaft. If this happens, the compressor will reduce the pressure in its feed pipeline, until further reduction is no longer possible. While reaching this stage, vibrations caused by friction and internal torsion will destabilise the whole turbo-expander. The turbo-expander typically runs at 10,000 rpm. Slight disturbance at these conditions will cause the turbo-expander to crash: in the worst possible case the turbo-expander separates from its skid, thereby rupturing feed pipelines and surrounding objects. At least, the friction will upset expander internals, like pressure seal rings. This may cause leakage and shortens the lifetime of the turbo-expander. To avoid this type of disaster, a bypass from expander to the compressor must be installed. In this way, when the feed pipeline of the compressor part gets blocked for whatever reason, the compressor gets feed from the expander, which, on its turn, runs at lower speed. In this way, the risk of a destabilised turbo-expander is greatly reduced. The plant should be shut down in this case anyway, rendering the loss of product streams by this measure insignificant. The amount of damage avoided is much higher than the loss of C_3^+ products, which are directly guided into the sales gas stream for a short time (until complete shutdown). When the feed stream of the expander part is blocked, the consequences are less critical. In that case the driving force for the compressor part is no longer available. The turbo-expander will automatically stop. Risk of internal vibrations is far less than in case of feed blockage of the compressor part of the turbo-expander. The output pressure of the compressor part is reduced, causing the main compressor at the end of the sales gas stream (K02) to require an increased duty to compress the sales gas to 190 bara. This is of course an insignificant problem, because in this case the whole process must be shut down.

To avoid disaster in case of gas leakage (LEL of ethane is 3v%; for LEL and UEL of other components, see appendix 10-6, table 10.6.1) a gas detection system should be installed. It is evident that any gas leakage in any part of the plant can be disastrous. Leakage may lead to fire and/or explosions. Therefore, if the detection system detects gas, it must sound the alarm. This must be done when the gas concentration reaches, for example, 25% of the LEL. That level is quite easily reached considering the low LEL of the components and the high pressure involved. At gas concentrations of for example 70% LEL, the gas detection systems should automatically initiate plant shut down. The use of pressure controllers is not sufficient in this case. Gas pressure in the system is always subject to slight deviations, making it impossible to detect gas leakage just by measuring pressure in the system.

11. Economy

The economics of the process are shown in table 11.1. For the calculation of the values given in this table the reader is referred to appendices 11-1 and 11-2.

Table 11.1: Total Investment Costs, Licence Costs en Working Capital

Type of costs	Mln Dfl @ 1999
Fixed Capital	125.88
Licence [1]	12.59
Working Capital [2]	25.18
Total Investment [3]	163.64

[1] Licence Costs = 0,10 * Fixed Capital
[2] Working Capital = 0,20 * Fixed Capital
[3] Total Investment Costs = 1.3 * Fixed Capital Costs

The working capital is selected at 20 % of the fixed capital. This is done because of the large consumption of utilities. These utilities require a large of amount of working capital to cover the first month of utility costs made. That is because of the fact, that from the point of delivering the first month's supply of C_3-product, C_4-product and C_5^+-product, the first payment is not to be expected for a month.
The licence costs are taken to be 10% of the fixed capital. Because this is part of the "worst case" scenario approach for the financial evaluation, it will probably be less.

11.1 Sales prices

The prices of the Feed, Sales Gas and the C_3, C_4 and C_5^+-products are given in table 11.2.

Table 11.2: prices of Feed, Sales Gas and Products

Components	Price (Dfl/sm^3)	Price (Mln Dfl/kton)
Feed	0.10	0.115
Sales Gas	0.10	0.130
C_3-product	115.50	0.197
C_4-product	144.80	0.255
C_5^+-product	144.40	0.219

The large difference in price (in Dfl/sm^3) between the feed and the sales gas and the product can be explained by considering the fact, that at these standard conditions (293.15 K, 1 bara) their densities are quite different (see appendix 5-1). The price per kton, however, is of the same magnitude.
It must be noted that the price per volume of the feed and the sales gas is the same. This is a result of the "worst case" scenario: it is assumed that, although the sales gas is delivered at a higher pressure (190 bara) than the feed gas (110 bara) and completely dehydrated (< 0.1 ppmwt), the price is the same as the price of feed gas. It also has to be noted though that the energy density (volume based) of sales gas is a little lower than that of the feed gas. This is caused by the removal of the higher hydrocarbons, which have considerable higher heats of combustion than the lower hydrocarbons present ([55], p. 5-75 and 5-78). It is therefore assumed that the price of

the sales gas equals that of the feed gas on a volume base. The price per unit of mass is higher, because of the lower density of the feed gas. Furthermore, it is possible that the heat of combustion per unit mass in sales gas is actually higher than in feed gas. That may be the case if one considers the fact that the heat of combustion per unit mass of methane is higher than that of butane ([55], p. 5-75 and 5-78). It was not further researched, if this is actually true. When dealing with such large feeds as in this plant, the common way to calculate the costs of the feed is the amount of energy extracted from the feed gas. For ease of calculation it was decided to use a mass based fee for cost calculation [39].
The actual price for the sales and feed gas was taken to be 0.10 Dfl/sm^3. This price is based on two sources. The first one, a Gasunie newsletter [57], yields a price of 0.145 Dfl/sm^3. The price is based on a purchase of more than 50 mln sm^3 per year. It is obvious that a throughput of hundred times this amount should have a reducing influence on this price. Also this gas has specifications for the "end" user, e.g. no water. The feed of this process consists of raw natural gas, and should therefore be cheaper than the partly treated gas from Gasunie. In [58] an average wellhead price of about Dfl 0.0001/sm^3 was found. Therefore it seemes reasonable to assume a sales and feed gas price of Dfl 0.10/sm^3.
Prices for the C_3, C_4 and C_5^+-products are taken from [58].

11.2 Overall economics

The overall economics of the process are calculated in appendices 11-3 to 11-5. The main results are shown in table 11.3. These results can be characterised as reasonably good. The rate of return (ROR; 19.2%) is not sufficient for today's standards. For example, Shell demands a ROR of at least 20%, otherwise the project will not be realised. The pay out time is 5.2 years. Compared to a plant life of 25 years, this is reasonable.

Table 11.3: Net cash flow, rate of return en pay out time (Mln Dfl @ 1999)

Item	Unit	Result	Remarks
C_3-product Sales	kton/a	67.45	
Sales price	Mln Dfl/kton	0.20	
C_4-product Sales	Kton/a	54.10	
Sales price	Mln Dfl/kton	0.26	
C_5-product Sales	kton/a	37.98	
Sales price	Mln Dfl/kton	0.22	
Gross Income	Mln Dfl/a	159.53	
Production Costs	Mln Dfl/a	128.16	
Net Cash Flow [1] (before tax)	Mln Dfl/jaar	31.37	= (A)
Total Investment	Mln Dfl	163.64	= (B)
Pay-Out-Time (POT)	Years	5.2	= (B)/(A)
Rate of Return (ROR)	%	19.2	= (A)/(B)*100%
Earning power	%	17.3	

[1] Net Cash Flow = sales * Sales Price – Production Costs.

In table11.4 a sensitivity analysis is carried out. It can be seen here that the fluctuation in product prices has the largest impact on the economics of the process. Overall it can

be stated that the process is quite sensitive to deviations of prices. It should be noted that at the moment the natural gas price, and hence the prices of natural gas products, are (very) low due to the low crude oil prices: the price of natural gas is bound to the price of crude oil. Further decrease in natural gas price is not to be expected, since the crude oil price is rising from its historically low point. Therefore the chance of decreasing prices of natural gas product is small. A rise is more likely [58].
The prices of natural gas products are, although not completely, bound to the price of natural gas. An increase in the natural gas price is therefore likely to increase the prices of the products. An analysis has been made, in which both feed price as well as product prices are varied. The increase in prices has a very healthy influence in the process economics.
The location of the plant in Norway should have reducing effect on the utility costs. These consist mainly of electricity cost for the compressor. The electricity in Norway is estimated to be cheaper than in the Netherlands, due to the presence of hydropower electricity stations. This has a positive effect on the process economics.
Using the Norwegian prices for electricity can thus reduce costs. Other utility costs reduction may be found in the implementation of a closed cooling loop, to reduce the amount of cooling water. Without a full economical evaluation, it is not certain whether or not true reduction can be achieved. But such an evaluation is beyond the scope of this report. Rise in costs of feed will almost certainly be more than compensated for by the rise in revenue from product sales, due to higher product prices.

Table 11.4: Sensitivity analysis

		Pay Out Time		Rate Of Return		Total Investment		DCFROR (earning power)	
		Years	% dev. [1]	%	% dev.	Mln Dfl	% dev.	%	% dev.
Base Case		5.2	0	19.2	0	163.6	0	17.3	0
Feed	+10%	6.0	15	16.6	-14	163.6	0	15.0	-13
	-10%	4.6	-12	21.7	13	163.6	0	19.6	13
Feed+ prod. Prices	+10%	3.8	-27	26.4	38	163.6	0	23.5	36
	-10%	8.3	60	12.0	-38	163.6	0	10.4	-40
Util. Costs	+10%	6.1	17	16.3	-15	163.6	0	14.7	-15
	-10%	4.5	-13	22.0	15	163.6	0	19.8	14
Prod. Prices	+10%	3.5	-33	28.9	51	163.6	0	25.6	48
	-10%	10.6	104	9.4	-51	163.6	0	7.6	-56
PCE [2]	+10%	6.3	21	15.9	-17	180.0	10	14.3	-17
	-10%	4.3	-17	23.2	21	147.3	-10	20.8	20

[1] % deviation from Base Case

[2] Purchase Costs Equipment

It can be concluded that the process is most sensitive to deviations in product prices.

It must be stated though, that this financial evaluation is based on a "worst case" scenario: feed and utility prices and total investment costs estimations are quite high. In combination with the rise in crude oil prices, the process is economically feasible.

12. Conclusions and recommendations

The process option using two expansion steps gives the lowest compression costs, does not need external refrigeration, has a C_3-recovery of 97 % and meets all product specifications. Molecular sieves are the best option for dehydration of the feed and the use of distillation columns for NGL fractionation is possible and simple.
No significant waste streams are produced.
The production levels are high, as are the compression costs, cooling water streams and column diameter of the deethaniser. The process is not new, but C_3-recovery (97 %) is high compared to other processes.
The process is economically feasible.

Due to several uncertainties several recommendations can be given:

- Using two parallel trains if equipment sizing becomes a problem and if feed rates vary significantly.
- Acquiring more detailed information of use of a reflux heat exchanger and its performance in modelling and operation to see if it is possible to operate the process at a higher pressure level to reduce compressor costs.
- Researching if it is better to feed all the liquid obtained from cooling the vapour from the deethaniser back as reflux and use heavy hydrocarbons to wash the vapour in the absorber to reduce the reboiler duty and possibly the flows in the deethaniser.
- Using a closed cooling water loop so the maximum temperature of the cooling water can be raised.
- Acquiring more detailed information about the use of turbo-expanders from manufacturers. The contacted manufacturer gave high efficiencies, higher possible suction volumes and larger transferable duties, than were assumed, and lower exit temperatures after expansion than were calculated, possibly eliminating the need for two turbo expansion steps, or influencing the heat integration.
- Researching kinetic data for the molecular sieves and regeneration conditions in more depth.

13. Literature

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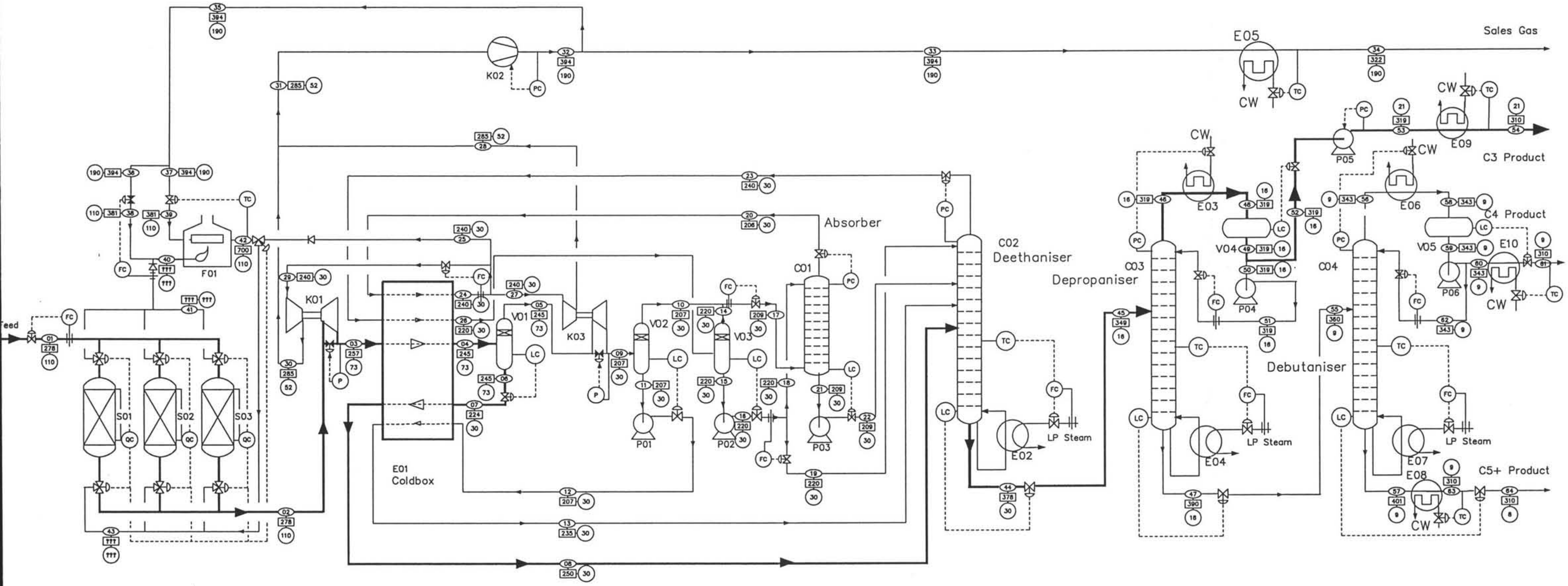
14. List of symbols used

Symbol	**Units**	**Description**
α	W/m^2.K	Heat transfer coefficient
α_a	-	average relative volatility of the light key
δ	-	Correction factor in LMTD calculation (Taborek)
ΔPdesign	kPa	designed pressure difference
ΔP <stream>	Pa	Actual pressure difference
ε,eff	-	Fin efficiency of heat transfer
λ	W/m^2.K	Thermal conductivity
μ	cP	Viscosity of fluid
μ_a	mNs/m^2	molar average liquid viscosity
μ,g	Pa.s	Vapour viscosity
μ,l	Pa.s	Liquid viscosity
μ,s	mPa.s	Solvent viscosity
ρ_gas	lb/ft^3	Gas density
ρ,g	kg/m^3	Vapour density
ρ,l	kg/m^3	Liquid density
ρ,s	kg/m^3	Solvent density
σ,f	N/m	Surface tension
σ	kg/s^2	Surface tension
φ	-	Vapour correction in condensing stream
a	J/mol	Constant in Peng-Robinson equation
A	m^2	Area
A	psimin/cPft^2	Dimensional constant for pressure drop calculation
A,b	m^2	Total cross sectional block area
A,m	m^2	Total cross sectional metal area.
Amin	m^2	Minimum area
Ax	m^2	cross sectional area per layer
Ax,t	m^2	Total cross sectional area per zone
b	m	gap between sheets plate-fin
b	m^3/mol	Constant in Peng-Robinson equation
B	-	Banking factor
B	psimin/cPft^2	Dimensional constant for pressure drop calculation
C	-	Parameter for different vane type demisters
C_2/C_3	-	C_2/C_3 ratio
symbol	Units	Description
Cp	J/kg.K	Heat capacity
Cp,g	J/kg.K	Vapour heat capacity
Cp,l	J/kg.K	Liquid heat capacity
C_{tr}	m/s	Corrected capacity gas load coefficient
C_{tr}^*	m/s	Capacity gas load coefficient
CW		Cooling water
D	m	Diameter
D, column	m	Column diameter

d,h	m	hydraulic fin diameter
d,hole	m	diameter of holes in fins
De	mm	Equivalent diameter
D,min	m	Minimum diameter
dP	bar	Pressure difference
dP,f	Pa	Frictional pressure loss between plates
dP/L	Psi/ft	Pressure drop per bed length
dQ/MTD	W/K	Incremental heat duty from heat curves/Temp driving force
dT	K	Increase in temperature
dTlm	K	Logarithmic mean temperature difference
dTm	K	Corrected logarithmic mean temperature difference
dTmn	K	Logarithmic mean temperature difference
E	-	Heat transfer effectiveness
E_0	%	Column efficiency
F	-	Convective boiling enhancement factor
f	-	Fanning friction factor
F_{lg}	-	Flow parameter
Ft	K	Log-mean temperature difference correction factor
F_omega	-	Constant in Peng-Robinson equation
g	m/s2	Gravitational acceleration
Gs	$kg/s.m^2$	Shell side mass flow rate per unit area
H	m	Height
hf(finh)	m	Fin height
Hi	-	Inside heat transfer coefficient
H,liquid	m	Liquid height
hs	$W/m^2.K$	Heat transfer coefficient shell side
I.d.	mm	Inner diameter
j,g	-	Colburn j-factor
Jh	-	Heat transfer factor
K	-	Equilibrium constant for the solute
Kw	W/m.K	Thermal conductivity of tube wall material
L	m	Length
L,2	m	Length of zone
L/D	-	Length/diameter ratio
L/di	-	Length/internal diameter ratio
Lambda, max	m/s	Maximum gas-load factor
LD50	g	Dose which kills 50 % of the test subjects
LMTD	K	Log mean temperature difference
LUB	ft	Length of unused bed
M_s	kg/kmol	Solvent molecular weight
m'	$kg/s.m^2$	Mass flow per unit area
M'	kg/s	Mass flow
MAC	mg/m^3	Maximum allowable concentration
Mol. Wt	g/mol	Molecular weight
MTZ	ft	Length of the mass transfer zone

MW	g/mol	Molecular weight
n	m^{-1}	Number of fins per meter
N,tot	-	Total number of layers per zone
NTU	-	Number of transfer units needed
Nr	-	Number of tubes in a vertical row
Nu	-	Nusselt number
O.d.	mm	Outer diameter
Omega	-	Constant in Peng-Robinson equation
p,h	m	Area fraction of perforated fin corrugations
Pr	-	Prandtl number
Pr,g	-	Prandtl number, vapour phase
Pr,l	-	Prandtl number, liquid phase
Pt	mm	Tube pitch
q	kg/kg	Vapour quality (kg vapour/kg total)
Ql	m^3/s	Liquid flow rate
Qsi	-	Pressure drop coefficient
Qv	m^3/s	Vapour flow rate
Qv	m^3/s	Vapour flow rate
R	-	Dimensionless temperature ratio
rd	$m^2.K/W$	Fouling heat transfer resistance
R	-	Thermal capacity ratio
Re	-	Reynolds number
Re,g	-	Reynolds number, vapour phase
Re,l	-	Reynolds number, liquid phase
Re,Tl	-	Reynolds number, liquid phase in a condensing stream
S	-	Dimensionless temperature ratio
S,eff	m^2/ m^2	Overall effective heat transfer surface per parting sheet surface area
St	-	Stanton number
t	m	Fin/corrugation thickness
T	K	Temperature
t,b	m	Minimum thickness of edge bar
Tc	K	Critical temperature
th	s	Liquid hold-up time
Tin	K	Entrance temperature
Tr	-	Reduced temperature
T_s	m	Tray spacing
Tuit	K	Exit temperature
Tw	K	Wall temperature
U	$W/m^2.K$	Heat-transfer coefficient
Ug, max	m/s	Maximum superficial gas velocity
UO	$W/m^2.K$	Overall heat transfer coefficient
Us	m/s	Shell side fluid velocity
V	m^3	Volume
V	m^3/mol	Molecular volume
V,chosen	m/s	Chosen superficial gas velocity

V,max	m/s, ft/min	Maximum superficial gas velocity
W	m	Effective width for heat transfer
x	-	Component fraction in liquid phase
x	-	parameter in fin efficiency equation
x	mPa.s.kmol	parameter in absorber efficiency equation
X	-	Parameter in equation for convective boiling
y	-	Component fraction in vapour phase
Z	-	Parameter in equation for multi-components



01	C3 ABSORBER	E03	CONDENSER C03	E09	C3 COOLER	P01	BOTTOMS V02	S01	MOLSIEVE BED	V04	REFLUX ACC. C03
02	DEETHANISER	E04	REBOILER C03	E10	C4 COOLER	P02	BOTTOMS V03	S02	MOLSIEVE BED	V05	REFLUX ACC. C04
03	DEPROPANISER	E05	SALES GAS COOLER	F01	FURNACE/STACK	P03	BOTTOMS C01	S03	MOLSIEVE BED		
04	DEBUTANISER	E06	CONDENSER C04	K01	TURBO EXPANDER 1	P04	REFLUX C03	V01	L/V SEPARATOR 1		
01	COLDBOX	E07	REBOILER C04	K02	SALES GAS COMPRESSOR	P05	C3 PUMP	V02	L/V SEPARATOR 2		
02	REBOILER C02	E08	C5+ COOLER	K03	TURBO EXPANDER 2	P06	REFLUX C04	V03	L/V SEPARATOR 3		

Process Equipment Summary

Designers

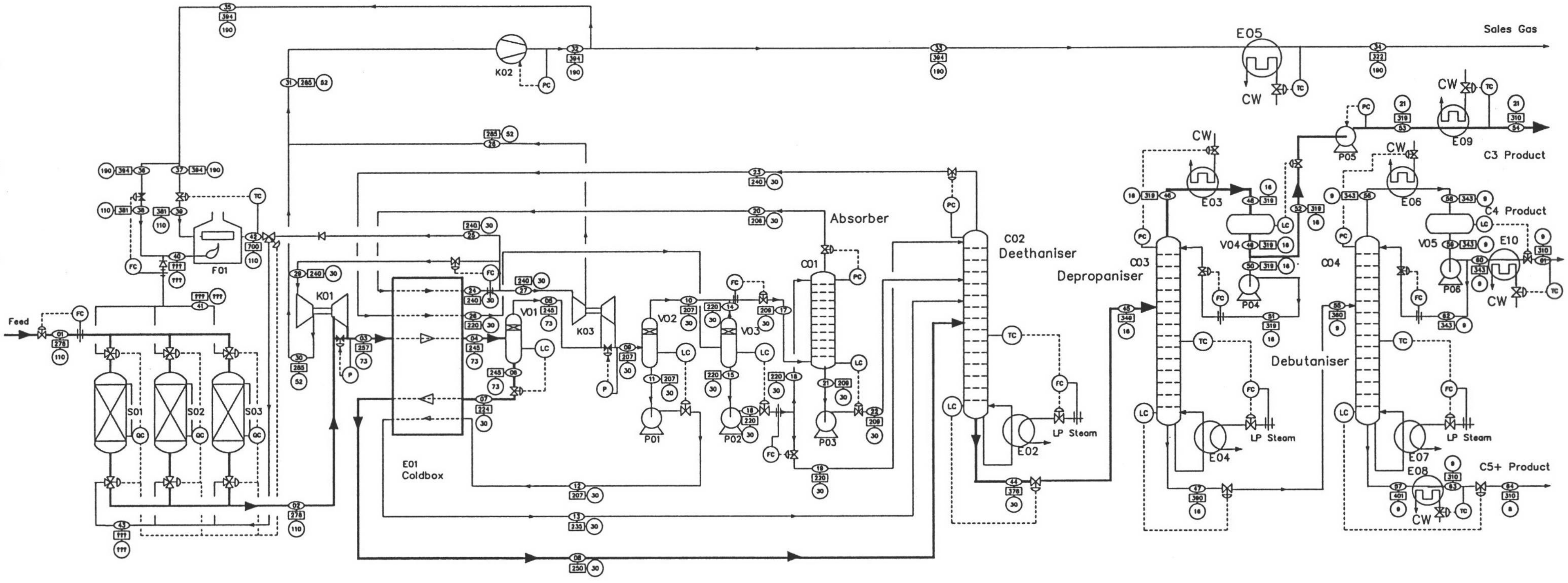
R.F. Eilers
K.M. de Lathouder
J.R. Law
B.A. Tulleken

Process Flow Scheme

Project	:	Recovery of Hydrocarbon Products from a 5 Gsm3/a Natural Gas Stream
Project ID Number	:	CPD 3228
Completion Date	:	23 April 1999

⬭ Stream Number ▭ Temperature [K] ○ Pressure [bara]

Appendix 1-1 Process Flow Scheme



Process Equipment Summary

C01	C3 ABSORBER	E03	CONDENSER C03	E09	C3 COOLER	P01	BOTTOMS V02	S01	MOLSIEVE BED	V04	REFLUX ACC. C03
C02	DEETHANISER	E04	REBOILER C03	E10	C4 COOLER	P02	BOTTOMS V03	S02	MOLSIEVE BED	V05	REFLUX ACC. C04
C03	DEPROPANISER	E05	SALES GAS COOLER	F01	FURNACE/STACK	P03	BOTTOMS C01	S03	MOLSIEVE BED		
C04	DEBUTANISER	E06	CONDENSER C04	K01	TURBO EXPANDER 1	P04	REFLUX C03	V01	L/V SEPARATOR 1		
E01	COLDBOX	E07	REBOILER C04	K02	SALES GAS COMPRESSOR	P05	C3 PUMP	V02	L/V SEPARATOR 2		
E02	REBOILER C02	E08	C5+ COOLER	K03	TURBO EXPANDER 2	P06	REFLUX C04	V03	L/V SEPARATOR 3		

Designers

R.F. Eilers
K.M. de Lathouder
J.R. Law
B.A. Tulleken

Process Flow Scheme

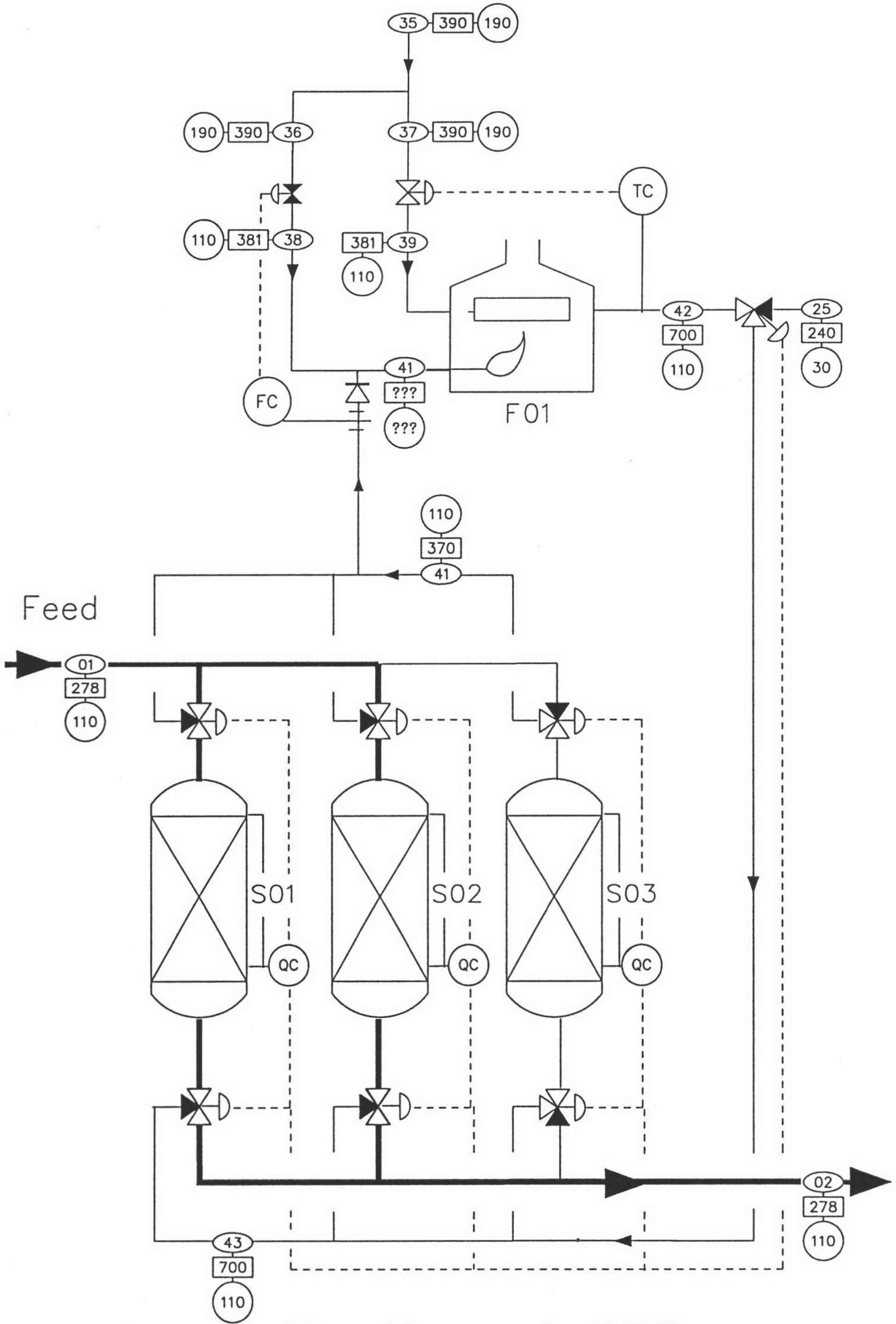
Project	:	Recovery of Hydrocarbon Products from a 5 Gsm3/a Natural Gas Stream
Project ID Number	:	CPD 3228
Completion Date	:	23 April 1999

⬭ Stream Number ▭ Temperature [K] ○ Pressure [bara]

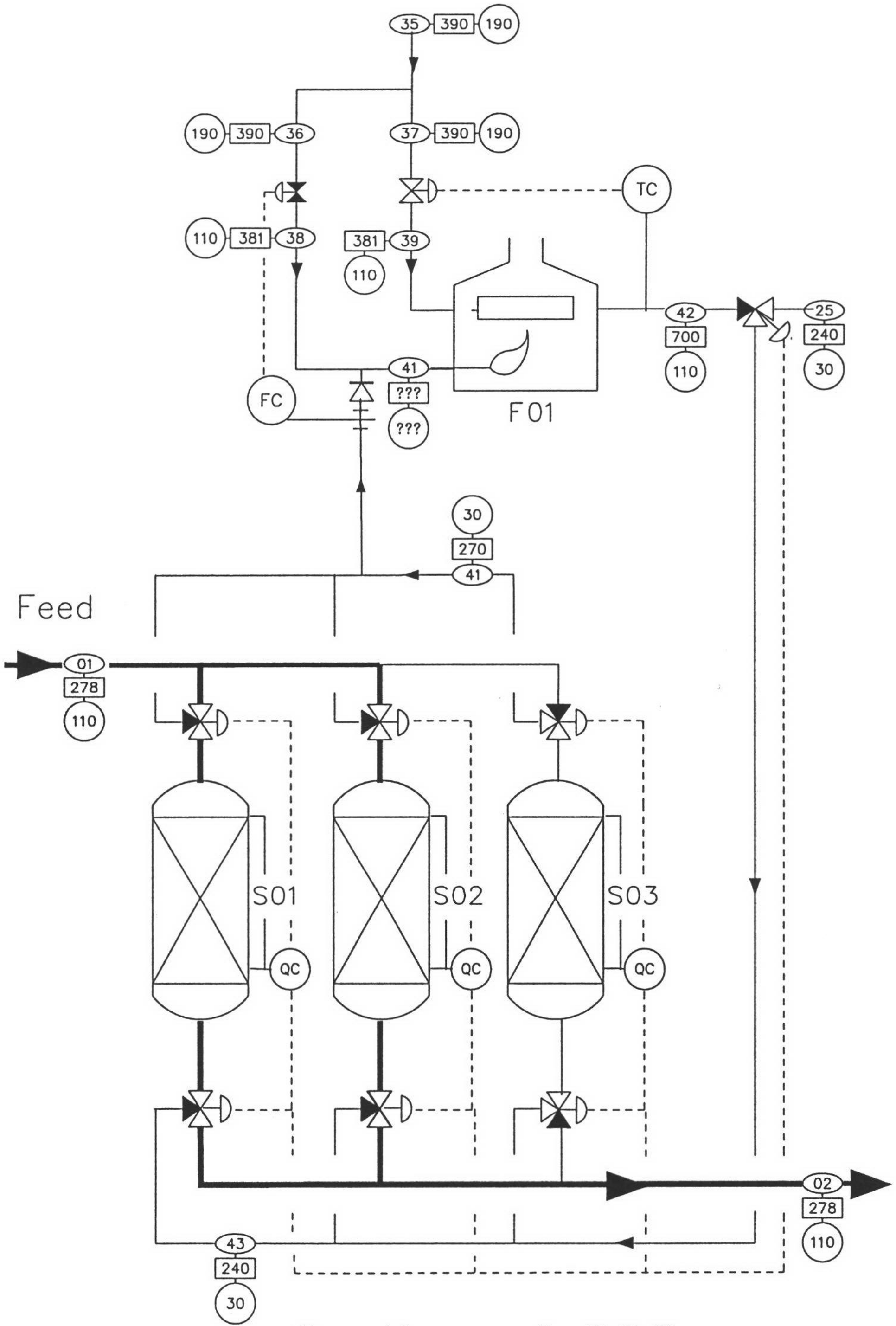
Appendix – 2

Appendix 1–2

Flowscheme of the regeneration cycle



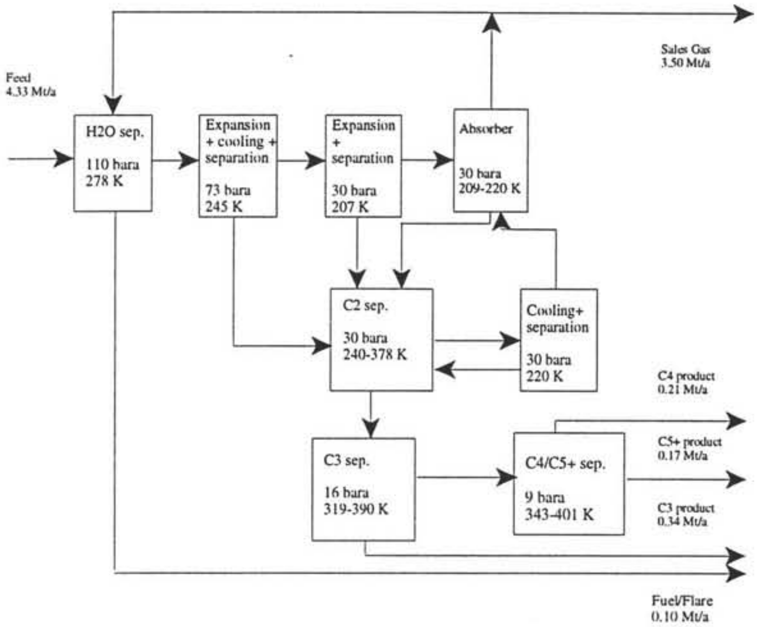
Heating of S03



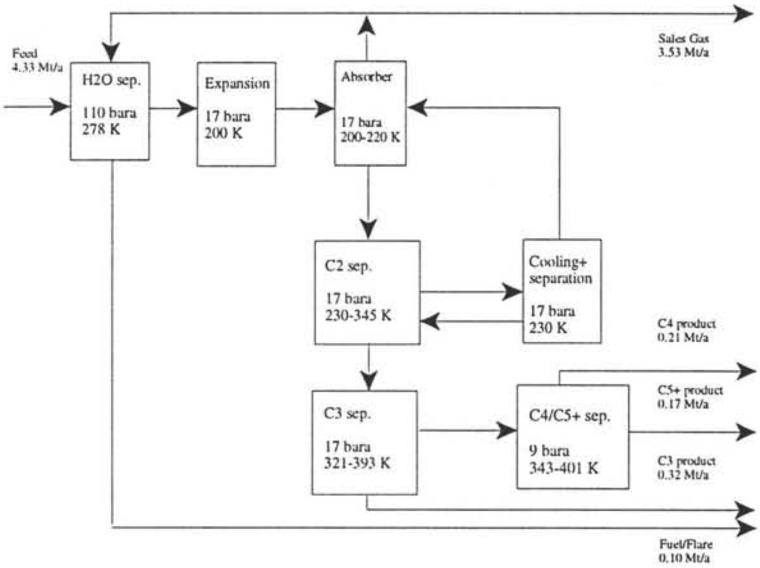
Cooling of S03

Appendix 2-1

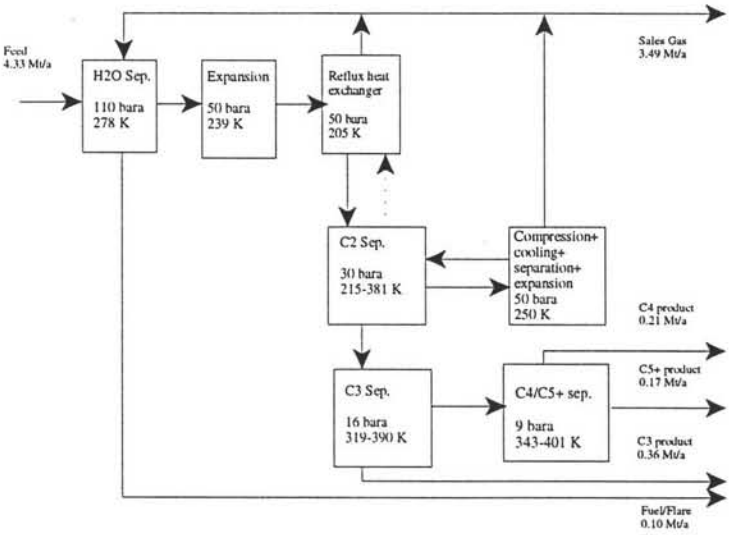
Combined blockschemes



Blockscheme of the turbo-expander process
(see also appendix 2-2 for t/t values and 6-4 for a description of the process)



Blockscheme of the Cryomax process
(see also appendix 2-3 for t/t values and 6-5 for a description of the process)



Blockscheme of the Reflux heatexchanger process, based on Patent US 4.846.863
(see also appendix 2-4 for t/t values and 6-6 for a description of the process)

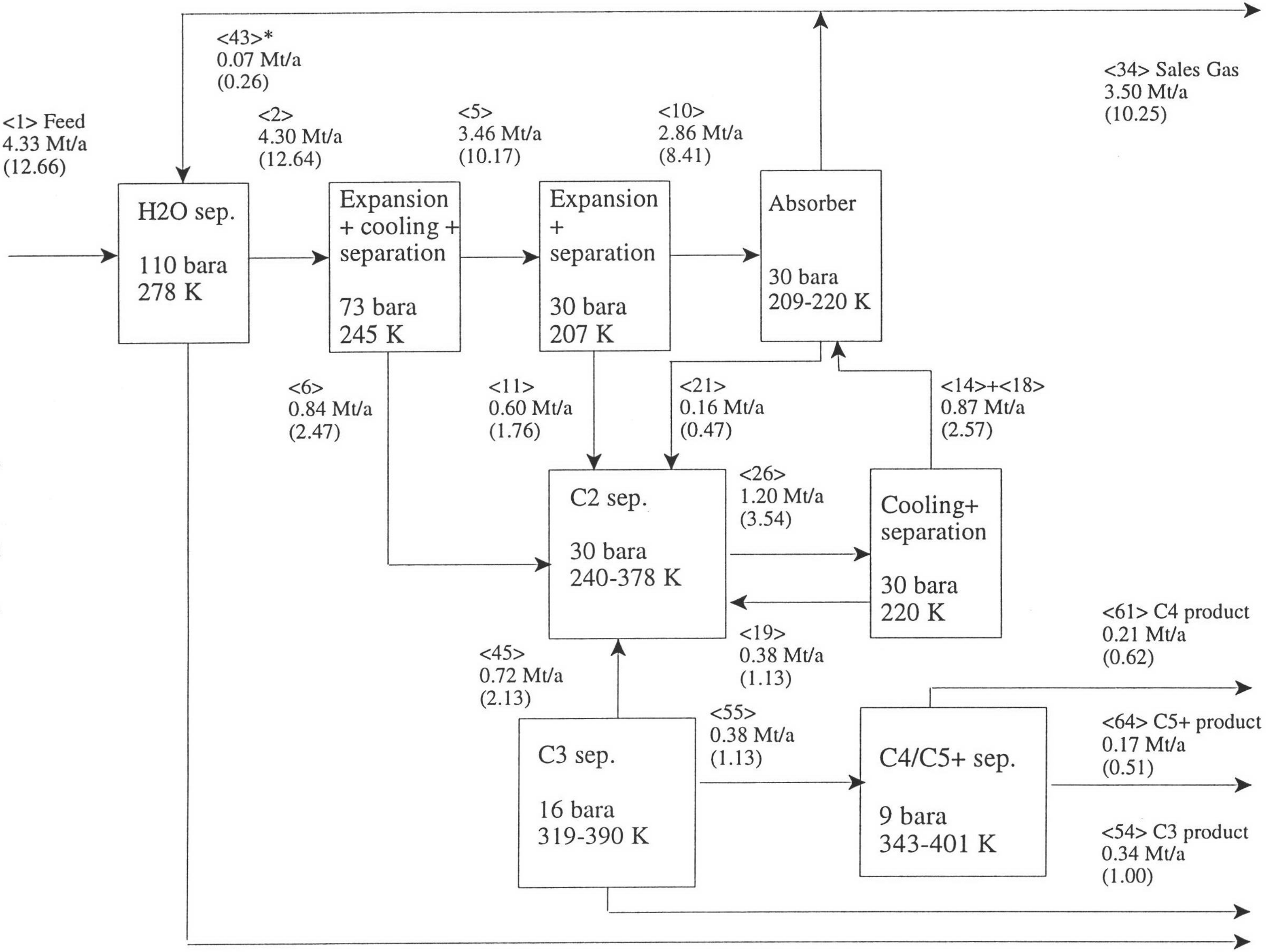
Total in:
4.33 Mt/a

N.B. Flows are in Mt/a

Total out
4.33 Mt/a

Appendix 2-2

Blockscheme of the turbo-expander process



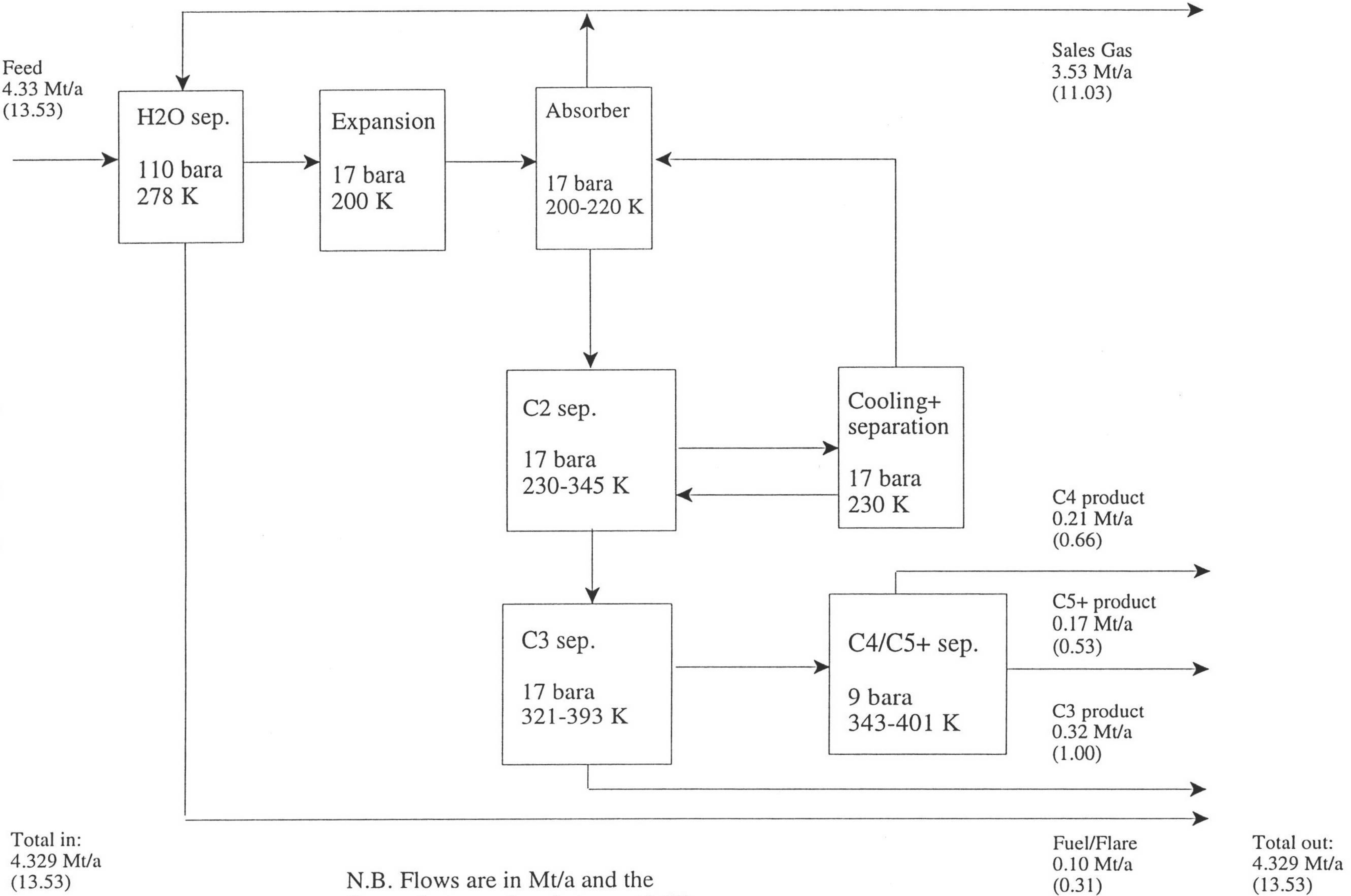
Total in:
4.33 Mt/a
(12.66)

Total out:
4.33 Mt/a
(12.66)

N.B. Flows are in Mt/a and the figures between brackets are t/t C3-product values.
*If in heating cycle of regeneration

Appendix 2-3

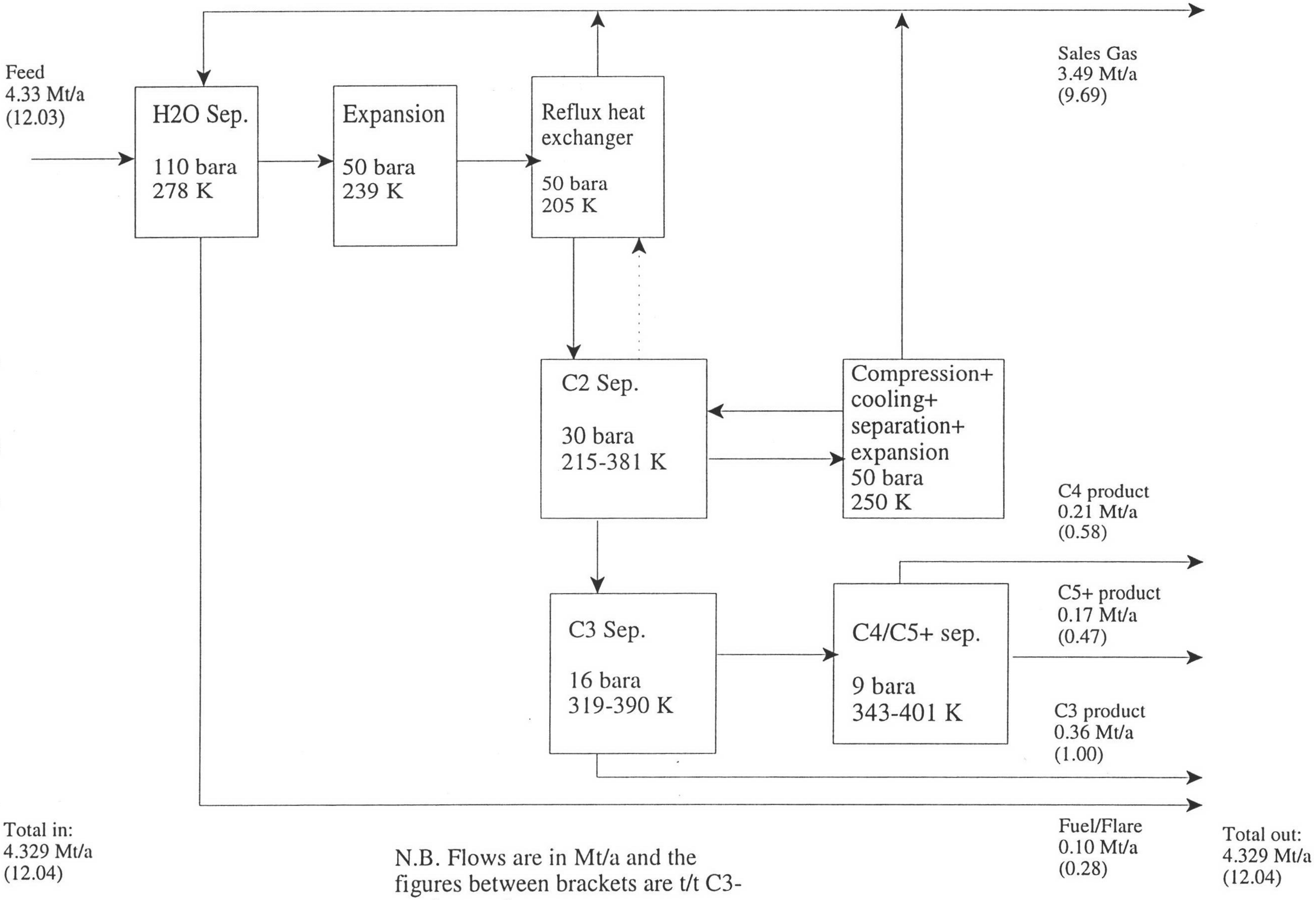
Blockscheme of the Cryomax process



N.B. Flows are in Mt/a and the figures between brackets are t/t C3-product values.

Appendix 2-4

Blockscheme of the reflux heatexchanger process.
Based on patent US 4.846.863.



N.B. Flows are in Mt/a and the figures between brackets are t/t C3-product values.

Appendix 3

Equipment Summary Sheets

Technical University Delft,
Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. : **NAME :**	**C-01** **Absorber**	**C-02** **Deethaniser**	**C-03** **Depropaniser**	**C-04** **Debutaniser**	**S-01/0** **Molsiev** **beds**
	Tray Column	Tray Column	Tray Column	Tray Column	
Pressure [bara] :	30.0/30.2	30.0/30.3	16.0/16.5	9.0/9.3	110/30
Temp. [K] :	206.1/208.7	240.2/377.9	318.8/389.7	343.0/400.8	278-37
Volume [m^3] :					33.5
Diameter [m] :	5.70	3.50/ 6.50	2.90	2.50	2.53
L or H [m] :	16.40	6.20/16.90	33.50	24.00	6.65
Internals					n.a.
-Tray Type :	Valve	Valve	Valve	Valve	
-Number of trays :	24	39	61	40	
-Fixed Packing					
Type :	n.a.	n.a.	n.a.	n.a.	
Shape :	n.a.	n.a.	n.a.	n.a.	
-Catalyst					
Type :	n.a.	n.a.	n.a.	n.a.	
Shape :	n.a.	n.a.	n.a.	n.a.	
- Vane-type demister					
- Angle :	n.a.	n.a.	n.a.	n.a.	
- Baffle Distance :	n.a.	n.a.	n.a.	n.a.	
Materials of Construction *(2)* **:**	Column: CS Trays: CS	Column: CS Trays: CS	Column: CS Trays: CS	Column: CS Trays: CS	CS
Number in					
- series :	1	1	1	1	-
- parallel :	-	-	-	-	3 *(3)*
Other :					Contains zeolite

Remarks :
(1) Depending on adsorption, heating or cooling cycle
(2) CS = Carbon Steel
(3) Molsieve beds S01, S02 and S03 are three units, operated in parallel

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Sheet 1

Technical University Delft,
Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. : NAME :	V-01 L/V Separator 1	V-02 L/V Separator 2	V-03 L/V Separator 3	V-04 Reflux accumulator C-03	V-05 Reflux accumulator C-04
	Vertical	Vertical	Vertical	Vertical	Vertical
Pressure [bara] :	72.5	30.0	30.0	16.0	9.0
Temp. [K] :	245.0	207.5	220.0	318.8	343.0
Volume [m^3] :	51.6	47.0	22.3	35.0	20.0
Diameter [m] :	2.84	2.97	1.65	3.0	2.3
L or H [m] :	8.16	6.80	10.40	4.6	4.5
Internals					
-Tray Type :	n.a.	n.a.	n.a.	n.a.	n.a.
-Tray number :	n.a.	n.a.	n.a.	n.a.	n.a.
-Fixed Packing					
Type :	n.a.	n.a.	n.a.	n.a.	n.a.
Shape :	n.a.	n.a.	n.a.	n.a.	n.a.
-Catalist					
Type :	n.a.	n.a.	n.a.	n.a.	n.a.
Shape :	n.a.	n.a.	n.a.	n.a.	n.a.
- Vane-type demister					
- Angle :	45°	45°	45°	n.a.	n.a.
- Baffle Distance :	0.02 m	0.02 m	0.02 m	n.a.	n.a.
Materials of Construction :	CS	CS	CS	CS	CS
Number in					
- series :	1	1	1	1	1
- parallel :	-	-	-	-	-
Other :					

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Sheet 2 of 7

Technical University Delft,
Process Systems Engineering

HEAT EXCHANGERS & FURNACES - SUMMARY

EQUIPMENT NR. : NAME :	E-01 coldbox zone 1	E-01 coldbox zone 2	E-01 coldbox zone 3	E-01 coldbox zone 4	E-01 coldbox zone 5	E-01 coldbo zone
Type :	Plate-fin heat exchanger					
stream nr :	<20> <23>	<12> <20> <23>	<7> <12> <20> <23>	<3> <7> <12> <20>	<3> <7> <20>	<3> <7>
Duty [MW] :	0.5	7.0	1.55	4.7	2.8	0.2
Heat Exchange area [max] [m^2] :	1750	964	348	547	419	83
Number						
- series :	6	6	6	6	6	6
- parallel :	4	4	4	4	4	4
Pressure [bara]						
stream <3> :				73.5	73.5	73.5
stream <7> :			31	31	31	31
stream <12> :		31	31	31		
stream <20> :	31	31	31	31	31	
stream <23> :	31	31	31			
Temperature In [K]						
stream <3> :				251.1	256.07	257.05
stream <7> :			223.73	226.8	235	242.5
stream <12> :		207.5	223.73	226.8		
stream <20> :	206.14	207.5	223.73	226.8	235	
stream <23> :	221.5	238.9	241.7			
Out [K]						
stream <3> :				241.7	251.1	256.07
stream <7> :			226.8	235	242.5	250
stream <12> :		223.73	226.8	235		
stream <20> :	207.5	223.73	226.8	235	242.5	
stream <23> :	220	221.5	238.9			
Special Materials of Construction :	All zones constructed of Aluminium					

Remarks :six zones in series, four units in parallel

Designers : R.F. Eilers K.M.de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Technical University Delft,
Process Systems Engineering

HEAT EXCHANGERS & FURNACES - SUMMARY

EQUIPMENT NR. : NAME :	E-02 Reboiler C-02	E-03 Condensor C-03	E-04 Reboiler C-03	E-05 Sales gas cooler
Type :	Thermo-syphon	Floating head	Thermo-syphon	Thermo-syphon
Substance - Tubes : - Shell :	 Bottom C-02 LP Steam	 Cool water Top C-03	 Bottom C-03 LP Steam	 Sales gas Cool Water
Duty [kW] :	16,707	3,314	8,087	12,896
Heat Exchange area [m^2] :	441	497	213	479
Number - series : - parallel :	 1 1	 1 3	 1 1	 1 2
Pressure [bara] - Tubes : - Shell :	 30 5	 1 16	 16 5	 190 1
Temperature In/Out [K] - Tubes : - Shell :	 377.9/377.9 423.2/423.2	 303.2/311.2 318.8/318.8	 389.7/389.7 423.3/423.2	 394.3/322.2 303.2/311.2
Special Materials of Construction :	CS	CS	CS	CS
Other :				

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

HEAT EXCHANGERS & FURNACES - SUMMARY

EQUIPMENT NR. : NAME :	E-06 Condensor C-04	E-07 Reboiler C-04	E-08 C5+ cooler	E-09 C3 cooler	E-10 C4 coole
Type :	Floating head	Thermo-syphon	Floating head	Floating head	Floating head
Substance **- Tubes** : **- Shell** :	Cool water Top C-04	Bottom C-04 LP Steam	C5+ product Cool water	C3 product Cool water	C4 product Cool water
Duty [kW] :	6,176	5,373	1,338	325	625
Heat Exchange area [m^2] :	301	142	75	77	66
Number **- series** : **- parallel** :	1 1	1 1	1 1	1 1	1 1
Pressure [bara] **- Tubes** : **- Shell** :	1 9	9 5	8 1	21 1	9 1
Temperature In/Out [K] **- Tubes** : **- Shell** :	303.2/311.2 343.0/343.0	400.8/400.8 423.2/423.2	400.8/310.2 303.2/311.2	319.5/310.2 303.2/311.2	400.8/310. 303.2/311.
Special Materials of Construction :	CS	CS	CS	CS	CS
Other :					

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Technical University Delft,
Process Systems Engineering

PUMPS, BLOWERS & COMPRESSORS - SUMMARY

EQUIPMENT NR. : **NAME :**	**P-01** **Bottom V-02**	**P-02** **Bottom V-03**	**P-03** **Bottom C-01**	**P-04** **Reflux C-03**	**P-05** **C3 Pump**
Type : **Number :**	Centrifugal 1	Centrifugal 1	Centrifugal 1	Centrifugal 1	Centrifugal 1
Medium transferred :	<12>	<16>	<22>	<51>	<53>
Capacity **[kg/s] :** **[m^3/s] :**	 0.16	 0.06	 0.01	 0.05	 0.03
Density [kg/m^3] :	127	364	348	452	452
Pressure [bara] **Suct./Disch. :**	30.0/30.6	29.7/30.8	30.2/30.7	16.0/17.5	16.0/21.0
Temperature **In/Out [K] :**	207.5/207.5	220.0/220.0	208.7/208.7	318.8/318.8	318.8/319.5
Power [kW] **- Theor. :** **- Actual :**	 9.8 11.8	 6.1 7.6	 0.7 1.0	 7.4 9.2	 12.5 16.4
Number **- Series :** **- Parallel :**	 1	 1	 1	 1	 1
Special Construction Materials :	MS casing	MS casing	MS casing	MS casing	MS casing
Other :					

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

PUMPS, BLOWERS & COMPRESSORS - SUMMARY

EQUIPMENT NR. : **NAME** :	**P-06** **Reflux C-04**	**K-01** **Turbo expander 1**	**K-02** **Sales gas compressor**	**K-03** **Turbo expander**
Type : **Number** :	Centrifugal 1	Turbo-expander 1	Compressor 1	Turbo-expande 1
Medium transferred :	<60>,<62>	^{1} Sales gas ^{2} Feed	Sales gas	^{1} Sales gas ^{2} Vapor V-01
Capacity : **$[m^3/s]$** :	0.04	^{1} 1.41 ^{2} 0.90	2.50	^{1} 2.20 ^{2} 1.00
Density : **$[kg/m^3]$** :	504.9	^{1} 47.7 ^{2} 111.8	112.7	^{1} 47.7 ^{2} 48.8
Pressure [bara] : **Suct./Disch.** :	9.0/10.2	^{1} 30.0/52.4 ^{2} 109.8/72.5	52.2/190.0	^{1} 30.0/52.4 ^{2} 72.5/30.0
Temperature : **In/Out [K]** :	343.0/343.0	^{1} 240.2/284.5 ^{2} 278.2/257.1	284.8/394.3	^{1} 240.2/284.5 ^{2} 245.0/207.5
Power [kW] **- Theor.** : **- Actual** :	 4.9 6.3	^{1} ^{2} 2,522 3,957 3,363 3,363	 20,475 24,088	^{1} ^{2} 3,939 6,179 5,252 5,252
Number **- Series** **- Parallel** :	 1	^{1} ^{2} 1 1	 1	^{1} ^{2} 1 1
Special Construction Materials :	MS casing	-	-	-
Other :				

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Appendix 4

Equipment Specification Sheets

Technical University Delft
Process Systems Engineering

DISTILLATION COLUMN - SPECIFICATION SHEET

EQUIPMENT NUMBER : C-01
NAME : Absorber

General Data

Function : - ~~distillation~~ / ~~extraction~~ / absorption / ________

Column Type : - ~~packed~~ / tray / ~~spray~~ / ________

Tray Type *(1)* : - ~~cap~~ / ~~sieve~~ / valve / ________
Tray Number
- **Theoretical** : 10
- **Actual** : 24
- **Feed Top (actual)** : 1
- **Feed Bottom (actual)** : 24

Tray Distance (HETP) [m] : 0.45
Column Diameter [m] : 5.70
Column Height [m] : 16.40
Tray Material : CS *(2)*
Column Material : CS

Heating : - none / ~~open steam~~ / ~~reboiler~~ / ________

Process Conditions

Stream Details	Feed (top)	Feed (bottom)	Top	Bottom	~~Reflux~~ / ~~Absorbent~~
Temp. [°C]	: 220.0	: 209.5	: 206.1	: 208.7	
Pressure [bara]	: 30.0	: 30.0	: 30.0	: 30.0	
Density $[kg/m^3]$	: 363.6	: 41.8	: 44.4	: 348.4	
Mass Flow [kg/s]	: 9.68	: 113.80	: 118.28	: 5.20	

Components	mol%	wt%	mol%	wt%	mol%	wt%	mol%	wt%	mol%	wt%
Methane	34.9	21.2	88.9	78.6	87.1	76.2	44.3	26.0		
Ethane	56.3	64.1	6.7	11.1	8.7	14.3	32.9	36.2		
Propane	0.2	0.4	0.5	1.3	0.1	0.2	14.7	23.8		
Butane	0.0	0.0	0.0	0.1	0.0	0.0	1.5	3.1		
Pentane*	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.2		
Rest	8.6	14.4	3.8	8.9	4.0	9.3	6.6	10.6		

Column Internals

Trays
Number of ~~Caps~~ / ~~sieve holes~~ / valves : 1503
Active Tray Area $[m^2]$: 18.88
Weir Height [mm]: 75.0
Diameter of
- ~~chute pipe~~ / ~~hole~~ / valve [mm]: 40.0

Packing *Not applicable*
- **Type** :
- **Material** :
- **Volume** $[m^3]$:
- **Length** [m] :
- **Width** [m] :
- **Height** [m] :

Remarks:
(1) Tray numbering from top to bottom.
(2) CS = Carbon Steel.

Designers: R.F. Eilers J.R. Law	K.M.de Lathouder B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Sheet 1 of 35

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DISTILLATION COLUMN - SPECIFICATION SHEET

EQUIPMENT NUMBER : C-02
NAME : Deethaniser

General Data

Function : - distillation / ~~extraction~~ / ~~absorption~~ / ________

Column Type : - ~~packed~~ / tray / ~~spray~~ / ________

Tray Type : - ~~cap~~ / ~~sieve~~ / valve / ________
Tray Number *(1)*
- Theoretical : 27
- Actual : 39
- Feed (actual) : 1/12/17/23
Tray Distance (HETP) [m] : 0.45
Tray Material : CS *(2)*
Column Material : CS
Column Diameter (trays 1 to 11) [m] : 3.50
Column Diameter (trays 12 to 39) [m] : 6.50
Column Height [m] : 23.10
(top to tray 11) [m] : 6.20
Column Heigt (tray 12 to bottom) [m] : 16.90
Heating : - ~~none~~ / ~~open steam~~ / reboiler / __________ *(3)*

Process Conditions *(4)*

Stream Details	**Feed 1**		**Feed 2**		**Feed 3**		**~~Reflux~~ / ~~Absorbent~~**		**Extractant**	
Temp. [K]	: 220.0		: 208.7		: 235.0		:			
Pressure [bara]	: 30.0		: 30.0		: 30.0		:			
Density [kg/m^3]	: 363.6		: 348.4		: 127.1		:			
Mass Flow [kg/s]	: 10.91		: 5.19		: 19.83		:			
Components	**Mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Methane	34.9	21.2	44.3	26.1	44.2	23.9				
Ethane	56.3	64.1	32.9	36.2	25.1	25.5				
Propane	0.2	0.4	14.7	23.8	16.8	25.0				
Butane	0.0	0.0	1.5	3.1	6.3	12.4				
Pentane	0.0	0.0	0.1	0.2	1.8	4.4				
Rest	8.6	14.3	6.5	10.6	5.8	8.8				

Column Internals *(5)*

Trays (1 to 11)
Number of ~~Caps~~ / ~~sieve holes~~ / valves : 588
Active Tray Area [m^2] : 7.38
Weir Height [mm] : 75.0
Diameter of - ~~chute pipe~~ / ~~hole~~ / valve [mm] : 40.0

Packing *Not applicable*
- Type :
- Material :
- Volume [m^3] :
- Length [m] :
- Width [m] :
- Height [m] :

Remarks:
(1) Tray numbering from top to bottom.
(2) CS = Carbon Steel.
(3) Reboiler is E02; operates with LP steam.
(4) Streams are continued on next sheet.
(5) Description of other tray on next sheet.

Designers: R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Technical University Delft
Process Systems Engineering

DISTILLATION COLUMN - SPECIFICATION SHEET

EQUIPMENT NUMBER : C-02
NAME : Deethaniser (continued)

General Data

Function : - distillation / ~~extraction~~ / ~~absorption~~ / _______

Column Type : - ~~packed~~ / tray / ~~spray~~ / ________

Tray Type : - ~~cap~~ / ~~sieve~~ / valve / _________
Tray Number *(1)*
- Theoretical :
- Actual :
- Feed (actual) :
Tray Distance (HETP) [m] : **Tray Material : CS**
Column Diameter [m] : **Column Material : CS**
Column Height [m] :
Heating : - ~~none~~ / ~~open steam~~ / reboiler / __________

Process Conditions

Stream Details	**Feed 4**		**Top**		**Bottom**		~~**Reflux**~~ / ~~**Absorbent**~~		~~**Extractant**~~	
Temp. [°C]	: 250.0		: 240.2		: 377.9		:		:	
Pressure [bara]	: 30.0		: 30.0		: 30.0		:		:	
Density [kg/m³]	: 85.9		: 46.9		: 311.5		:		:	
Mass Flow [kg/s]	: 27.90		: 39.77		: 24.06		:		:	
Components	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Methane	52.9	27.7	58.0	40.6	0.0	0.0				
Ethane	15.4	15.1	34.6	45.3	0.8	0.5				
Propane	12.4	17.8	0.1	0.2	56.1	46.2				
Butane	8.4	15.9	0.0	0.0	27.0	29.3				
Pentane	6.7	17.5	0.0	0.0	16.1	24.0				
Rest	4.2	6.0	7.4	13.9	0.0	0.0				

Column Internals

Trays (12 to 39)
Number of Caps / sieve holes / valves : 1994
Active Tray Area [m²]: **25.06**
Weir Length [mm]: **75.0**
Diameter of
- chute pipe / hole / valve [mm]: **40.0**

Packing *Not applicable*
- Type :
- Material :
- Volume [m³] :
- Length [m] :
- Width [m] :
- Height [m] :

remarks:

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken

Project ID-Number : **CPD**
Date : 23 April 1999

Appendix 4

Technical University Delft
Process Systems Engineering

DISTILLATION COLUMN - SPECIFICATION SHEET

EQUIPMENT NUMBER : C-03
NAME : Depropaniser

General Data

Function : - distillation / ~~extraction~~ / ~~absorption~~ / _______

Column Type : - ~~packed~~ / tray / ~~spray~~ / ________

Tray Type : - ~~cap~~ / ~~sieve~~ / valve / _________
Tray Number *(1)*
- Theoretical : 44
- Actual : 61
- Feed (actual) : 25
Tray Distance (HETP) [m] : 0.45 **Tray Material** : CS *(2)*
Column Diameter [m] : 2.90 **Column Material** : CS
Column Height [m] : 33.50
Heating : - ~~none~~ / ~~open steam~~ / reboiler / __________ *(3)*

Process Conditions *(1)*

Stream Details	**Feed**	**Top**	**Bottom**	**Reflux / ~~Absorbent~~**	**~~Extractant~~**
Temp. [K]	: 349.0	: 318.8	: 389.7	: 318.8	
Pressure [bara]	: 16.0	: 16.0	: 16.0	: 16.0	
Density [kg/m^3]	: 85.9	: 452.0	: 462.2	: 452.0	
Mass Flow [kg/s]	: 24.06	: 11.30	: 12.76	: 22.46	

Components *(2)*	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Methane	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		
Ethane	0.8	0.5	1.4	1.0	0.0	0.0	1.4	1.0		
Propane	56.1	46.2	98.0	98.3	0.1	0.1	98.0	98.3		
Butane	27.0	29.3	0.6	0.7	62.3	54.6	5.1	0.7		
Pentane	16.1	24.0	0.0	0.0	37.6	45.3	0.0	0.0		
Rest	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		

Column Internals *(3)*

Trays *(4)*
Number of ~~caps~~ / ~~sieve holes~~ / valves : 390
Active Tray Area [m^2] : 4.89
Weir Height [mm] : 75.0
Diameter of
- ~~chute pipe~~ / ~~hole~~ / valve [mm] : 40.0

Packing *Not applicable*
- Type :
- Material :
- Volume [m^3] :
- Length [m] :
- Width [m] :
- Height [m] :

Remarks:
(1) Tray numbering from top to bottom.
(2) CS = Carbon Steel.
(3) Reboiler is E04; operates with LP steam.

Designers: R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Technical University Delft
Process Systems Engineering

DISTILLATION COLUMN - SPECIFICATION SHEET

EQUIPMENT NUMBER : C-04
NAME : Debutaniser

General Data

Function : - distillation / ~~extraction~~ / ~~absorption~~ / ________

Column Type : - ~~packed~~ / tray / ~~spray~~ / ________

Tray Type : - ~~cap~~ / ~~sieve~~ / valve / ________
Tray Number *(1)*
- Theoretical : 30
- Actual : 40
- Feed (actual) : 17
Tray Distance (HETP) [m] : 0.45 **Tray Material** : CS *(2)*
Column Diameter [m] : 2.50 **Column Material** : CS
Column Height [m] : 24.00
Heating : - ~~none~~ / ~~open steam~~ / reboiler / ________ *(3)*

Process Conditions

Stream Details	**Feed**	**Top**	**Bottom**	**Reflux / ~~Absorbent~~**	**~~Extractant~~**
Temp. [K]	: 363.5	: 343.0	: 400.8	: 343.0	
Pressure [bara]	: 9.0	: 9.0	: 9.0	: 9.0	
Density [kg/m^3]	: 72.5	: 504.9	: 519.3	: 504.9	
Mass Flow [kg/s]	: 12.76	: 7.02	: 5.74	: 13.72	

Components	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Methane	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		
Ethane	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0		
Propane	0.1	0.1	0.2	98.3	0.0	0.1	0.2	0.2		
Butane	62.3	54.6	98.7	0.7	1.3	54.6	98.7	98.5		
Pentane	37.6	45.3	1.1	0.0	98.7	45.3	1.1	1.3		
Rest	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		

Column Internals

Trays
Number of ~~caps~~ / ~~sieve holes~~
/ valves : 305
Active Tray Area [m^2]: 3.82
Weir Height [mm]: 75.0
Diameter of
- ~~chute pipe~~ / ~~hole~~
/ valve [mm]: 40.0

Packing *Not applicable*
- Type :
- Material :

- Volume [m^3] :
- Length [m] :
- Width [m] :
- Height [m] :

Remarks:
(1) Tray numbering from top to bottom.
(2) CS = Carbon Steel.
(3) Reboiler is E07; operates with LP steam.

Designers: R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-01	**In Series :6**
NAME : Coldbox (zone 1)	**In Parallel :4**

General Data

Type : - **Heat Exchanger** - **~~Vaporizer~~**
- **~~Cooler~~**
- **~~Reboiler~~**
- **~~Condensor~~**

Execution : - **~~Fixed Tube Sheets~~** - **Plate-fin Heat Exchanger**
- **~~Floating Head~~**
- **~~Finned Tubes~~**
- **~~Hair Pin~~**
- **~~Thermosiphon~~**
- **~~Double Tube~~**

Position : - **~~Horizontal~~**
- **Vertical**

Capacity	**[MW]**	**:**	0.5	
			<20>	**<23>**
Heat Exchange Area	**[m^2]**	**:**	1750	155
Overall Heat transfer Coëff.	**[$W/m^2.K$]**	**:**	2593	5430
Log. Mean Temp.Diff. (LMTD)	**[K]**	**:**	13.93	
Passes Tubeside		**:**	-	
Passes Shell Side		**:**	-	
Correction Factor LMTD (> 0.75)		**:**	-	
Corrected LMTD	**[K]**	**:**	-	

Process Conditions

			<20>	**<23>**
Medium		**:**	vapour	liquid/vapour
Mass Stream	**[kg/s]**	**:**	118.9134	39.9892
Average Specific Heat	**[kJ/kg.K]**	**:**	2.803	2.843
Heat of Evap./ Condens.	**[kJ/kg]**	**:**	-	-
Temperature IN	**[K]**	**:**	206.14	221.5
Temperature OUT	**[K]**	**:**	207.5	220
Pressure	**[bara]**	**:**	30	30
Materials of Construction		**:**	Aluminium	Aluminium

Remarks :Data for zone 1 only

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-01 **In Series :**
NAME : Coldbox (zone 2) **In Parallel :**

General Data

Type : - Heat Exchanger - ~~Vaporizer~~
- ~~Cooler~~ - ~~Reboiler~~
- ~~Condensor~~

Execution : - ~~Fixed Tube Sheets~~ - Plate-fin Heat Exchang
- ~~Floating Head~~ - ~~Finned Tubes~~
- ~~Hair Pin~~ - ~~Thermosiphon~~
- ~~Double Tube~~

Position : - ~~Horizontal~~
- Vertical

Capacity	**[MW]**	:	7		
			<20>	**<12>**	**<23>**
Heat Exchange Area	**[m^2]**	:	**964**	**38**	**133**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	:	**769**	**3203**	**2321**

Log. Mean Temp.Diff. (LMTD) [K] : **14.58**

Passes Tubeside : -
Passes Shell Side : -

Correction Factor LMTD (> 0.75) : -
Corrected LMTD [K] : -

Process Conditions

			<20>	<12>	<23>
Medium		:	vapour	liquid/ vapour	liquid/ vapour
Mass Stream	**[kg/s]**	:	118.9134	19.933	39.9892
Average Specific Heat	**[kJ/kg.K]**	:	2.635	2.497	2.67
Heat of Evap./ Condens.	**[kJ/kg]**	:	-	-	-
Temperature IN	**[K]**	:	207.5	207.5	238.9
Temperature OUT	**[K]**	:	223.73	223.73	221.5
Pressure	**[bara]**	:	30	30	72.5
Materials of Construction		:	Alumina	Alumina	Alumina

Remarks :Data for zone 2 only

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-01
NAME : Coldbox (zone 3)
In Series :6
In Parallel :4

General Data

Type : - Heat Exchanger - ~~Vaporizer~~
- ~~Cooler~~ - ~~Reboiler~~
- ~~Condensor~~

Execution : - ~~Fixed Tube Sheets~~ - Plate-fin Heat Exchanger
- ~~Floating Head~~ - ~~Finned Tubes~~
- ~~Hair Pin~~ - ~~Thermosiphon~~
- ~~Double Tube~~

Position : - ~~Horizontal~~
- Vertical

Capacity	**[MW]**	1.55			
		<20>	<12>	<7>	<23>
Heat Exchange Area	**[m^2]**	348	14	27	77.8
Overall Heat transfer Coëff	**[$W/m^2.K$]**	2088	6418	4717	2897
Log. Mean Temp.Diff. (LMTD)	**[K]**	15.03			
Passes Tubeside		-			
Passes Shell Side		-			
Correction Factor LMTD (> 0.75)		-			
Corrected LMTD	**[K]**	-			

Process Conditions

		<20>	<12>	<7>	<23>
Medium		vapour	liquid/ vapour	liquid/ vapour	liquid/ vapour
Mass Stream	**[kg/s]**	118.9134	19.933	28.056	39.9892
Average Specific Heat	**[kJ/kg.K]**	2.476	2.549	2.393	2.416
Heat of Evap./ Condens.	**[kJ/kg]**	-	-	-	-
Temperature IN	**[K]**	223.73	223.73	223.73	241.7
Temperature OUT	**[K]**	226.8	226.8	226.8	238.9
Pressure	**[bara]**	30	30	30	30
Materials of Construction		Alumina	Alumina	Alumina	Alumina

Remarks : Data for zone 3 only

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER	**: E-02**	**In Series**	**:1**
NAME	**: Reboiler C-02**	**In Parallel**	**:1**

General Data

Type : - ~~Heat Exchanger~~ - ~~Vaporizer~~
- ~~Cooler~~
- Reboiler
- ~~Condensor~~

Execution : - Fixed Tube Sheets - ~~Plate Heat Exchanger~~
- ~~Floating Head~~
- ~~Finned Tubes~~
- ~~Hair Pin~~
- Thermosiphon
- ~~Double Tube~~

Position : - ~~Horizontal~~
- Vertical

Capacity	**[kW]**	:	16,707	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	441	**(Calc.)**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	:	837	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	:	45.3	
Passes Tubeside		:	1	
Passes Shell Side		:	1	
Correction Factor LMTD (> 0.75)		:	1	
Corrected LMTD	**[°C]**	:	45.3	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	LP steam	Bottom C-02
Mass Stream	**[kg/s]**	:	7.59	74.95
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	74.95
- Condense	**[kg/s]**	:	7.59	-
Average Specific Heat	**[kJ/kg.°C]**	:	-	-
Heat of Evap./ Condens.	**[kJ/kg]**	:	2,200	222.9
Temperature IN	**[K]**	:	423.2	377.9
Temperature OUT	**[K]**	:	423.2	377.9
Pressure	**[bara]**	:	5	30
Materials of Construction		:	CS	CS

Remarks :

Designers : R.F.Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-03 | **In Series :1**
NAME : Condensor C-03 | **In Parallel :3**

General Data

Type : - ~~Heat Exchanger~~ - ~~Vaporizer~~
- ~~Cooler~~ - ~~Reboiler~~
- Condensor

Execution : - ~~Fixed Tube Sheets~~ - ~~Plate Heat Exchanger~~
- Floating Head - ~~Finned Tubes~~
- Hair Pin - ~~Thermosiphon~~
- ~~Double Tube~~

Position : - ~~Horizontal~~
- Vertical

Capacity	**[kW]**	:	3,314	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	497	**(Calc.)**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	:	600	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	:	11.12	
Passes Tubeside		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (> 0.75)		:	1	
Corrected LMTD	**[°C]**	:	11.12	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Top C-03	Cooling water
Mass Stream	**[kg/s]**	:	33.77	297.3
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	-
- Condense	**[kg/s]**	:	33.77	-
Average Specific Heat	**[kJ/kg.°C]**	:	-	4.200
Heat of Evap./ Condens.	**[kJ/kg]**	:	294.4	-
Temperature IN	**[K]**	:	318.8	303.2
Temperature OUT	**[K]**	:	318.8	311.2
Pressure	**[bara]**	:	16	1
Materials of Construction		:	CS	CS

Remarks :

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-04 **In Series :1**
NAME : Reboiler C-03 **In Parallel :1**

General Data

Type :
- ~~Heat Exchanger~~
- ~~Cooler~~
- ~~Condensor~~
- ~~Vaporizer~~
- Reboiler

Execution :
- Fixed Tube Sheets
- ~~Floating Head~~
- ~~Hair Pin~~
- ~~Double Tube~~
- ~~Plate Heat Exchanger~~
- ~~Finned Tubes~~
- Thermosiphon

Position :
- ~~Horizontal~~
- Vertical

Capacity	**[kW]**	:	8,087	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	213	**(Calc.)**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	:	1133	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	:	33.45	
Passes Tubeside		:	1	
Passes Shell Side		:	1	
Correction Factor LMTD (> 0.75)		:	1	
Corrected LMTD	**[°C]**	:	33.45	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	LP steam	Bottom C-03
Mass Stream	**[kg/s]**	:	3.68	25.57
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	25.57
- Condense	**[kg/s]**	:	3.68	-
Average Specific Heat	**[kJ/kg.°C]**	:	-	-
Heat of Evap./ Condens.	**[kJ/kg]**	:	2,200	316.2
Temperature IN	**[K]**	:	423.2	389.7
Temperature OUT	**[K]**	:	423.2	389.7
Pressure	**[bara]**	:	5	16
Materials of Construction		:	CS	CS

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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Technical University Delft,
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER	**: E-05**	**In Series**	**:1**
NAME	**: Sales gas cooler**	**In Parallel**	**:2**

General Data

Type :
- ~~Heat Exchanger~~
- Cooler
- ~~Condensor~~
- ~~Vaporizer~~
- ~~Reboiler~~

Execution :
- ~~Fixed Tube Sheets~~
- Floating Head
- ~~Hair Pin~~
- ~~Double Tube~~
- ~~Plate Heat Exchanger~~
- ~~Finned Tubes~~
- ~~Thermosiphon~~

Position :
- Horizontal
- ~~Vertical~~

Capacity	**[kW]**	**:**	12,896	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	479	**(Calc.)**
Overall Heat transfer Coëff.	**[$W/m^2.K$]**	**:**	620	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	**:**	43.4	
Passes Tubeside		**:**	1	
Passes Shell Side		**:**	1	
Correction Factor LMTD (> 0.75)		**:**	1	
Corrected LMTD	**[°C]**	**:**	43.4	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Cooling water	Sales gas
Mass Stream	**[kg/s]**	**:**	383.8	58.6
Mass Stream to				
- Evaporize	**[kg/s]**	**:**	-	-
- Condense	**[kg/s]**	**:**	-	-
Average Specific Heat	**[kJ/kg.°C]**	**:**	4.200	3.063
Heat of Evap./ Condens.	**[kJ/kg]**	**:**	-	-
Temperature IN	**[K]**	**:**	303.2	394.3
Temperature OUT	**[K]**	**:**	311.2	322.2
Pressure	**[bara]**	**:**	1	190
Materials of Construction		**:**	CS	CS

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-06
NAME : Condensor C-04
In Series :1
In Parallel :1

General Data

Type : - ~~Heat Exchanger~~ - ~~Vaporizer~~
- ~~Cooler~~ - ~~Reboiler~~
- Condensor

Execution : - ~~Fixed Tube Sheets~~ - ~~Plate Heat Exchanger~~
- Floating Head - ~~Finned Tubes~~
- Hair Pin - ~~Thermosiphon~~
- ~~Double Tube~~

Position : - ~~Horizontal~~
- Vertical

Capacity	**[kW]**	: 6,176	**(Calc.)**
Heat Exchange Area	**[m^2]**	: 301	**(Calc.)**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	: 575	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	: 35.71	
Passes Tubeside		: 2	
Passes Shell Side		: 1	
Correction Factor LMTD (> 0.75)		: 1	
Corrected LMTD	**[°C]**	: 35.71	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Top C-04	Cooling water
Mass Stream	**[kg/s]**	:	20.24	184.7
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	-
- Condense	**[kg/s]**	:	20.24	-
Average Specific Heat	**[kJ/kg.°C]**	:	-	4.200
Heat of Evap./ Condens.	**[kJ/kg]**	:	315.2	-
Temperature IN	**[K]**	:	343.0	303.2
Temperature OUT	**[K]**	:	343.0	311.2
Pressure	**[bara]**	:	9	1
Materials of Construction		:	CS	CS

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD** 3228 Date : 23 April 1999

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER	**: E-07**	**In Series**	**:1**
NAME	**: Reboiler C-04**	**In Parallel**	**:1**

General Data

Type : - ~~Heat Exchanger~~ - ~~Cooler~~ - ~~Condensor~~ - ~~Vaporizer~~ - Reboiler

Execution : - Fixed Tube Sheets - ~~Floating Head~~ - ~~Hair Pin~~ - ~~Double Tube~~ - ~~Plate Heat Exchanger~~ - ~~Finned Tubes~~ - Thermosiphon

Position : - ~~Horizontal~~ - Vertical

Capacity	**[kW]**	:	5,373	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	142	**(Calc.)**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	:	1,697	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	:	22.34	
Passes Tubeside		:	1	
Passes Shell Side		:	1	
Correction Factor LMTD (> 0.75)		:	1	
Corrected LMTD	**[°C]**	:	22.34	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	LP steam	Bottom C-04
Mass Stream	**[kg/s]**	:	2.44	19.47
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	19.47
- Condense	**[kg/s]**	:	2.44	-
Average Specific Heat	**[kJ/kg.°C]**	:	-	-
Heat of Evap./ Condens.	**[kJ/kg]**	:	2,200	276
Temperature IN	**[K]**	:	423.2	400.8
Temperature OUT	**[K]**	:	423.2	400.8
Pressure	**[bara]**	:	5	9
Materials of Construction		:	CS	CS

Remarks :

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-08	**In Series :1**
NAME : C5+ cooler	**In Parallel :1**

General Data

Type	:	- ~~Heat Exchanger~~ - Cooler - ~~Condensor~~	- ~~Vaporizer~~ - ~~Reboiler~~
Execution	:	- ~~Fixed Tube Sheets~~ - Floating Head - ~~Hair Pin~~ - ~~Double Tube~~	- ~~Plate Heat Exchanger~~ - ~~Finned Tubes~~ - ~~Thermosiphon~~
Position	:	- Horizontal - ~~Vertical~~	

Capacity	**[kW]**	:	1,338.3	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	75.0	**(Calc.)**
Overall Heat transfer Coëff.	**[W/m^2 .K]**	:	550	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	:	32.4	
Passes Tubeside		:	1	
Passes Shell Side		:	1	
Correction Factor LMTD (> 0.75)		:	1	
Corrected LMTD	**[°C]**	:	32.4	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Cooling water	C5+ product
Mass Stream	**[kg/s]**	:	39.8	5.74
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	-
- Condense	**[kg/s]**	:	-	-
Average Specific Heat	**[kJ/kg.°C]**	:	4.200	2.606
Heat of Evap./ Condens.	**[kJ/kg]**	:	-	-
Temperature IN	**[K]**	:	303.2	400.8
Temperature OUT	**[K]**	:	311.2	310.2
Pressure	**[bara]**	:	1	8
Materials of Construction		:	CS	CS

Remarks :

Designers :R.F.Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Technical University Delft,
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-09		**In Series**	**:1**
NAME : C3 cooler		**In Parallel**	**:1**

General Data

Type : - ~~Heat Exchanger~~ - ~~Vaporizer~~
- Cooler - ~~Reboiler~~
- ~~Condensor~~

Execution : - ~~Fixed Tube Sheets~~ - ~~Plate Heat Exchanger~~
- Floating Head - ~~Finned Tubes~~
- ~~Hair Pin~~ - ~~Thermosiphon~~
- ~~Double Tube~~

Position : - Horizontal
- ~~Vertical~~

Capacity	**[kW]**	:	324.7	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	77.3	**(Calc.)**
Overall Heat transfer Coëff.	**[$W/m^2.K$]**	:	550	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	:	7.64	
Passes Tubeside		:	1	
Passes Shell Side		:	1	
Correction Factor LMTD (> 0.75)		:	1	
Corrected LMTD	**[°C]**	:	7.64	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Cooling water	C3 product
Mass Stream	**[kg/s]**	:	9.66	11.3
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	-
- Condense	**[kg/s]**	:	-	-
Average Specific Heat	**[kJ/kg.°C]**	:	4.200	3.086
Heat of Evap./ Condens.	**[kJ/kg]**	:	-	-
Temperature IN	**[K]**	:	303.2	319.5
Temperature OUT	**[K]**	:	311.2	310.2
Pressure	**[bara]**	:	1	21
Materials of Construction		:	CS	CS

Remarks :

Designers : R.F. Eilers	K.M. de Lathouder	Project ID-Number	: **CPD 3228**
J.R. Law	B.A. Tulleken	Date	: 23 April 1999

Technical University Delft,
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E-10
NAME : C4 cooler

In Series :1
In Parallel :1

General Data

Type : - ~~Heat Exchanger~~ - ~~Vaporizer~~
- Cooler
- ~~Reboiler~~
- ~~Condensor~~

Execution : - ~~Fixed Tube Sheets~~ - ~~Plate Heat Exchanger~~
- Floating Head
- ~~Finned Tubes~~
- ~~Hair Pin~~
- ~~Thermosiphon~~
- ~~Double Tube~~

Position : - Horizontal
- ~~Vertical~~

Capacity	**[kW]**	**:**	625.0	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	66.2	**(Calc.)**
Overall Heat transfer Coëff.	**[$W/m^2.K$]**	**:**	570	**(Appr.)**
Log. Mean Temp.Diff. (LMTD)	**[°C]**	**:**	16.4	
Passes Tubeside		**:**	1	
Passes Shell Side		**:**	1	
Correction Factor LMTD (> 0.75)		**:**	1	
Corrected LMTD	**[°C]**	**:**	16.4	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Cooling water	C4 product
Mass Stream	**[kg/s]**	**:**	18.60	7.06
Mass Stream to				
- Evaporize	**[kg/s]**	**:**	-	-
- Condense	**[kg/s]**	**:**	-	-
Average Specific Heat	**[kJ/kg.°C]**	**:**	4.200	2.695
Heat of Evap./ Condens.	**[kJ/kg]**	**:**	-	-
Temperature IN	**[K]**	**:**	303.2	343.0
Temperature OUT	**[K]**	**:**	311.2	310.2
Pressure	**[bara]**	**:**	1	9
Materials of Construction		**:**	CS	CS

Remarks :

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken

Project ID-Number : **CPD 3228**
Date : 23 April 1999

Technical University Delft,
Process Systems Engineering

TURBO-EXPANDER - SPECIFICATION SHEET

EQUIPMENT NUMBER : **K-01**		**Operating**	: 1
NAME : Turbo expander 1		**Installed Spare**	:

Service : Expansion feed / compression Sales gas
Type : Turbo-expander
Number : 1

Operating Conditions & Physical Data

Gas			:	Sales gas <29>	Feed <02>
Compressed/Expanded			:	Compressed	Expanded
Temperature	(T)	**[K]**	:	284.5	257.1
Density	(ρ)	**[kg/m^3]**	:	47.68	111.80
Viscosity	(η)	**[N.s/m^2]**	:	1.22E-05	1.36E-05
Vapour Pressure	(p_v)	**[bara]**	:	52.4	28.2
At Temperature		**[K]**	:	284.8	257.1

Power

Capacity	(Φ_v)	**[m^3/s]**	:	1.41	0.90
Suction Pressure	(p_s)	**[bara]**	:	30.0	109.8
Discharge Pressure	(p_d)	**[bara]**	:	52.4	72.5
Compressibility	(κ)	**[-]**	:	1.5426	2.4044
Theoretical Power		**[kW]**	:	2522	3957
Compressor Effic.	(η)	**[-]**	:	0.75	0.85
Power at Shaft		**[kW]**	:	3,363	3,363

Construction Details (1)

RPM	: 10,000	**Nominal diameter Suction Nozzle**	: m
Drive	: -	**Nominal diameter Discharge Nozzle**	: m
Type electrical motor	: -	**Cooled bearings**	: Yes/No
Tension [V]	: -	**Cooled Stuffing Box**	: Yes/No
Rotational direction	: -	**Smothering gland**	: Yes/No
Foundation Plate	: -	**If yes:**	
Flexible Coupling	: Yes/No	**- Seal Liquid**	: Yes/No
Pressure Gauge Suction	: Yes/No	**- Splash Rings**	: Yes/No
Pressure Gauge Discharge	: Yes/No	**- Packing Type**	
		- Mechanical Seal	: Yes/No

Construction Materials (2)

Compressor Housing	: Carbon Steel
Compressor/ Expander Wheel	: Aluminum alloy 6061-T6
Shaft	: Stainless Steel
Special provisions	: -
Operating Pressure [bara]	:

Remarks :
(1) Further details to be specified by Rotating Equipment Specialist
(2) Further details are given in Appendix XXX

Designers: R.F. Eilers K.M. de Lathouder	Project ID-Number	: **CPD 3228**
J.R. Law B.A. Tulleken	Date	: 23 April 1999

Technical University Delft,
Process Systems Engineering

CENTRIFUGAL COMPRESSOR - SPECIFICATION SHEET

EQUIPMENT NUMBER : K-02
NAME : Sales gas compressor
Operating : 1
Installed Spare :

Service : Compression sales gas
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Compressed Gas : Sales gas (methane, ethane,..)
Temperature (T) [K] : 394.3
Density (ρ) [kg/m^3] : 112.67
Viscosity (η) [N.s/m^2] : 1.88E-05
Vapour Pressure (p_v) [bara] : 190 **at temperature [K]** : 394.3

Power

Capacity (Φ_v) [m^3 /s] : 2.50
Suction Pressure (p_s) [bara] : 52.2
Discharge Pressure (p_d) [bara] : 190
Compressibility (κ) [-] : 1.5297 **(C_p/ C_v)**
Theoretical Power [kW] : 20,475
Compressor Effic. (η) [-] : 0.85
Power at Shaft [kW] : 24,088

Construction Details

RPM :
Drive : Electrical
Type electrical motor : -
Tension [V] : 3300
Rotational direction : -
Foundation Plate : -
Flexible Coupling : Yes/No
Pressure Gauge Suction : Yes/No
Pressure Gauge Discharge : Yes/No

Nominal diameter
Suction Nozzle [m] : -
Nominal diameter
Discharge Nozzle [m] : -
Cooled bearings : Yes/No
Cooled Stuffing Box : Yes/No
Smothering gland : Yes/No
If yes :
- Seal Liquid : Yes/No
- Splash Rings : Yes/No
- Packing Type :
- Mechanical Seal : Yes/No

Construction Materials

Pump House : steel
Pump Rotor : alloy steel
Shaft : alloy steel
Special provisions : -

Operating Pressure [bara] :

Wear Rings :
Shaft Box :

Test
Pressure [bara] :

Remarks :

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken
Project ID-Number : **CPD 3228**
Date : 23 April 1999

Technical University Delft,
Process Systems Engineering

TURBO-EXPANDER - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**K-03**	**Operating**	:	1
NAME	:	Turbo expander 2	**Installed Spare**	:	

Service : Expansion Vapor V-01 <05> / compression Sales gas
Type : Turbo-expander
Number : 1

Operating Conditions & Physical Data

Gas			:	Sales gas <27>	Vapor V-01 <05>
Compressed/Expanded			:	Compressed	Expanded
Temperature	(T)	[K]	:	284.5	207.5
Density	(ρ)	$[kg/m^3]$	:	47.66	48.83
Viscosity	(η_v)	$[N.s/m^2]$	:	1.22E-05	0.93E-05
Vapour Pressure	(p_v)	[bara]	:	52.4	21.4
At Temperature		[K]	:	284.5	207.5

Power

Capacity	(Φ_v)	$[m^3/s]$	:	2.20	1.00
Suction Pressure	(p_s)	[bara]	:	30.0	72.5
Discharge Pressure	(p_d)	[bara]	:	52.4	30.0
Compressibility	(κ)	[-]	:	1.5424	2.7428
Theoretical Power		[kW]	:	3939	6179
Compressor Effic.	(η)	[-]	:	0.75	0.85
Power at Shaft		[kW]	:	5252	5252

Construction Details (1)

RPM	:	10,100	**Nominal diameter Suction Nozzle**	:	m
Drive	:	-	**Nominal diameter Discharge Nozzle**	:	m
Type electrical motor	:	-	**Cooled bearings**	:	Yes/No
Tension [V]	:	-	**Cooled Stuffing Box**	:	Yes/No
Rotational direction	:	-	**Smothering gland**	:	Yes/No
Foundation Plate	:	-	**If yes:**		
Flexible Coupling	:	Yes/No	**- Seal Liquid**	:	Yes/No
Pressure Gauge Suction	:	Yes/No	**- Splash Rings**	:	Yes/No
Pressure Gauge Discharge	:	Yes/No	**- Packing Type**		
			- Mechanical Seal	:	Yes/No

Construction Materials (2)

Compressor Housing	:	Carbon Steel
Compressor/ Expander Wheel	:	Aluminum alloy 6061-T6
Shaft	:	Stainless Steel
Special provisions	:	-
Operating Pressure	:	

Remarks :
(1) Further details to be specified by Rotating Equipment Specialist
(2) Further details are given in Appendix XXX

Designers: R.F. Eilers K.M. de Lathouder	Project ID-Number	:	**CPD 3228**
J.R. Law B.A. Tulleken	Date	:	23 April 1999

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER	**: P-01**	**Operating**	**: 1**
NAME	**: Bottom V-02**	**Installed Spare**	**: 0**

Service : Liquid transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid			:	<12>
Temperature	**(T)**	**[K]**	:	207.5
Density	**(ρ)**	**[kg/m^3]**	:	127.06
Viscosity	**(η)**	**[$N.s/m^2$]**	:	7.52E-05
Vapour Pressure	**(p_v)**	**[bara]**	:	14.3 **at temperature [K]** : 207.5

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.16
Suction Pressure	**(p_s)**	**[bara]**	:	30.0
Discharge Pressure	**(p_d)**	**[bara]**	:	30.6
Theoretical Power		**[kW]**	:	9.8 [$= \Phi_v.(p_d - p_s).10^2$]
Pump Efficiency		**[-]**	:	0.83
Power at Shaft		**[kW]**	:	11.8

Construction Details (1)

RPM	**: 1450**	**Nominal diameter Suction Nozzle [m]**	**: -**
Drive	**: Electrical**	**Nominal diameter Discharge Nozzle [m]**	**: -**
Type electrical motor	**:**	**Cooled bearings**	**: Yes/No**
Tension [V]	**: 380**	**Cooled Stuffing Box**	**: Yes/No**
Rotational direction	**: -**	**Smothering gland**	**: Yes/No**
Foundation Plate	**: -**	**If yes :**	
Flexible Coupling	**: Yes**	**- Seal Liquid**	**: Yes/No**
Pressure Gauge, Suction	**: No**	**- Splash Rings**	**: Yes/No**
Pressure Gauge, Discharge	**: Yes**	**- Packing Type**	**:**
Min. Over Pressure above p_v, $\Delta p_b/\Delta p_m$ [bar]/[m liq.]	**: -**	**- Mechanical Seal**	**: Yes/No**
		N.P.S.H. ($\Delta p_m.\rho.g$) [m liq.]	**: 1263**

Construction Materials (2)

Pump House	**: MS**	**Wear Rings**	**:**
Pump Rotor	**: HT Steel**	**Shaft Box**	**:**
Shaft	**: HT Steel**		
Special provisions	**: none**		
Operating Pressure [bara]	**:**	**Test Pressure [bara]**	**:**

Remarks :
(1) Further details to be specified by Rotating Equipment specialist.
(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER : P-02	**Operating : 1**
NAME : Bottom V-03	**Installed Spare : 0**

Service : Liquid transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid			: <16>
Temperature	**(T)**	**[K]**	: 220
Density	(ρ)	$[\mathrm{kg/m^3}]$	: 363.64
Viscosity	(η)	$[\mathrm{N.s/m^2}]$	: 6.67E-05
Vapour Pressure	(p_v)	**[bara]**	: 13.8 **at temperature [K]** : 220

Power

Capacity	(Φ_v)	$[\mathrm{m^3/s}]$	: 0.06
Suction Pressure	(p_s)	**[bara]**	: 29.7
Discharge Pressure	(p_d)	**[bara]**	: 30.8
Theoretical Power		**[kW]**	: 6.1 $[= \Phi_v.(p_d - p_s).10^2]$
Pump Efficiency		**[-]**	: 0.80
Power at Shaft		**[kW]**	: 7.6

Construction Details (1)

RPM	**: 1450**	**Nominal diameter Suction Nozzle [m]**	**: -**
Drive	**: Electrical**	**Nominal diameter Discharge Nozzle [m]**	**: -**
Type electrical motor	**: -**	**Cooled bearings**	**: Yes/No**
Tension [V]	**: 380**	**Cooled Stuffing Box**	**: Yes/No**
Rotational direction	**: -**	**Smothering gland**	**: Yes/No**
Foundation Plate	**: -**	**If yes :**	
Flexible Coupling	**: Yes**	**- Seal Liquid**	**: Yes/No**
Pressure Gauge, Suction	**: No**	**- Splash Rings**	**: Yes/No**
Pressure Gauge, Discharge	**: Yes**	**- Packing Type**	**:**
Min. Over Pressure above p_v, $\Delta p_b/\Delta p_m$ **[bar]/[m liq.]**	**: -**	**- Mechanical Seal**	**: Yes/No**
		N.P.S.H. $(\Delta p_m.\rho.g)$ **[m liq.]**	**: 455**

Construction Materials (2)

Pump House	**: MS**	**Wear Rings**	**:**
Pump Rotor	**: HT Steel**	**Shaft Box**	**:**
Shaft	**: HT Steel**		
Special provisions	**: none**		
Operating Pressure [bara]	**:**	**Test Pressure [bara]**	**:**

Remarks :
(1) Further details to be specified by Rotating Equipment specialist.
(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER : P-03
NAME : Bottom C-01
Operating : 1
Installed Spare : 0

Service : Liquid transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid : <22>
Temperature (T) [K] : 208.7
Density (ρ) $[kg/m^3]$: 348.4
Viscosity (η) $[N.s/m^2]$: 6.73E-05
Vapour Pressure (p_v) [bara] : 21.7 **at temperature [K]** : 208.7

Power

Capacity (Φ_v) $[m^3/s]$: 0.01
Suction Pressure (p_s) [bara] : 30.2
Discharge Pressure (p_d) [bara] : 30.7

Theoretical Power [kW] : 0.7 $[= \Phi_v.(p_d - p_s).10^2]$
Pump Efficiency [-] : 0.75
Power at Shaft [kW] : 1.0

Construction Details (1)

RPM : 1450
Drive : Electrical
Type electrical motor : -
Tension [V] : 380
Rotational direction : -
Foundation Plate : -
Flexible Coupling : Yes
Pressure Gauge, Suction : No
Pressure Gauge, Discharge : Yes
Min. Over Pressure above p_v,
$\Delta p_b/\Delta p_m$ [bar]/[m liq.]: -

Nominal diameter
Suction Nozzle [m] : -
Nominal diameter
Discharge Nozzle [m] : -
Cooled bearings : Yes/No
Cooled Stuffing Box : Yes/No
Smothering gland : Yes/No
If yes :
- Seal Liquid : Yes/No
- Splash Rings : Yes/No
- Packing Type :
- Mechanical Seal : Yes/No
N.P.S.H. $(\Delta p_m.\rho.g)$ [m liq.]: 250

Construction Materials (2)

Pump House : MS
Pump Rotor : HT Steel
Shaft : HT Steel
Special provisions : none

Operating Pressure [bara] :

Wear Rings :
Shaft Box :

Test
Pressure [bara] :

Remarks :
(1) Further details to be specified by Rotating Equipment specialist.
(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken

Project ID-Number : **CPD 3228**
Date : 23 April 1999

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CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER	: P-04	**Operating**	: 1
NAME	: Reflux C-03	**Installed Spare**	: 0

Service : Liquid transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid			: <51>
Temperature	(T)	[K]	: 318.8
Density	(ρ)	[kg/m^3]	: 452.0
Viscosity	(η)	[N.s/m^2]	: 7.90E-05
Vapour Pressure	(p_v)	[bara]	: 16 **at temperature [K]** : 318.8

Power

Capacity	(Φ_v)	[m^3/s]	: 0.05	
Suction Pressure	(p_s)	[bara]	: 16.0	
Discharge Pressure	(p_d)	[bara]	: 17.5	
Theoretical Power		[kW]	: 7.4	[= $\Phi_v.(p_d - p_s).10^2$]
Pump Efficiency		[-]	: 0.80	
Power at Shaft		[kW]	: 9.2	

Construction Details (1)

RPM	: 1450	**Nominal diameter Suction Nozzle [m]**	: -
Drive	: Electrical	**Nominal diameter Discharge Nozzle [m]**	: -
Type electrical motor	: -	**Cooled bearings**	: Yes/No
Tension [V]	: 380	**Cooled Stuffing Box**	: Yes/No
Rotational direction	: -	**Smothering gland**	: Yes/No
Foundation Plate	: -	**If yes :**	
Flexible Coupling	: Yes	**- Seal Liquid**	: Yes/No
Pressure Gauge, Suction	: No	**- Splash Rings**	: Yes/No
Pressure Gauge, Discharge	: Yes	**- Packing Type**	:
Min. Over Pressure above p_v, $\Delta p_b/\Delta p_m$ [bar]/[m liq.]	: -	**- Mechanical Seal**	: Yes/No
		N.P.S.H. ($\Delta p_m.\rho.g$) [m liq.]	: 3

Construction Materials (2)

Pump House	: MS	**Wear Rings**	:
Pump Rotor	: HT Steel	**Shaft Box**	:
Shaft	: HT Steel		
Special provisions	: none		
Operating Pressure [bara]	:	**Test Pressure [bara]**	:

Remarks :
(1) Further details to be specified by Rotating Equipment specialist.
(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER : P-05
NAME : C3 pump
Operating : 1
Installed Spare : 0

Service : Liquid transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid : <53>
Temperature (T) [K] : 318.8
Density (ρ) [kg/m^3] : 452.0
Viscosity (η) [N.s/m^2] : 7.91E-05
Vapour Pressure (p_v) [bara] : 16 **at temperature [K]** : 318.8

Power

Capacity (Φ_v) [m^3 /s] : 0.03
Suction Pressure (p_s) [bara] : 16.0
Discharge Pressure (p_d) [bara] : 21.0

Theoretical Power [kW] : 12.5 **[= $\Phi_v.(p_d - p_s).10^2$]**
Pump Efficiency [-] : 0.76
Power at Shaft [kW] : 16.4

Construction Details (1)

RPM : 2900
Drive : Electrical
Type electrical motor : -
Tension [V] : 380
Rotational direction : -
Foundation Plate : -
Flexible Coupling : Yes
Pressure Gauge, Suction : No
Pressure Gauge, Discharge : Yes
Min. Over Pressure above p_v,
$\Delta p_b / \Delta p_m$ [bar]/[m liq.]: -

Nominal diameter
Suction Nozzle [m] : -
Nominal diameter
Discharge Nozzle [m] : -
Cooled bearings : Yes/No
Cooled Stuffing Box : Yes/No
Smothering gland : Yes/No
If yes :
- Seal Liquid : Yes/No
- Splash Rings : Yes/No
- Packing Type :
- Mechanical Seal : Yes/No
N.P.S.H. ($\Delta p_m . \rho . g$) [m liq.]: 3

Construction Materials (2)

Pump House : MS
Pump Rotor : HT Steel
Shaft : HT Steel
Special provisions : none

Operating Pressure [bara] :

Wear Rings :
Shaft Box :

Test
Pressure [bara] :

Remarks :
(1) Further details to be specified by Rotating Equipment specialist.
(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken
Project ID-Number : **CPD 3228**
Date : 23 April 1999

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CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER	**: P-06**	**Operating**	**: 1**
NAME	**: Reflux C-04**	**Installed Spare**	**: 0**

Service : Liquid transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid			: <60>,<62>
Temperature	**(T)**	**[K]**	: 343.0
Density	**(ρ)**	**[kg/m^3]**	: 504.9
Viscosity	**(η)**	**[N.s/m^2]**	: 1.09E-05
Vapour Pressure	**(p_v)**	**[bara]**	: 9 **at temperature [K]** : 343

Power

Capacity	**(Φ_v)**	**[m^3/s]**	: 0.04	
Suction Pressure	**(p_s)**	**[bara]**	: 9.0	
Discharge Pressure	**(p_d)**	**[bara]**	: 10.2	
Theoretical Power		**[kW]**	: 4.9	**[= $\Phi_v . (p_d - p_s) . 10^2$]**
Pump Efficiency		**[-]**	: 0.78	
Power at Shaft		**[kW]**	: 6.3	

Construction Details (1)

RPM	**: 1450**	**Nominal diameter Suction Nozzle [m]**	**: -**
Drive	**: Electrical**	**Nominal diameter Discharge Nozzle [m]**	**: -**
Type electrical motor	**: -**	**Cooled bearings**	**: Yes/No**
Tension [V]	**: 380**	**Cooled Stuffing Box**	**: Yes/No**
Rotational direction	**: -**	**Smothering gland**	**: Yes/No**
Foundation Plate	**: -**	**If yes :**	
Flexible Coupling	**: Yes**	**- Seal Liquid**	**: Yes/No**
Pressure Gauge, Suction	**: No**	**- Splash Rings**	**: Yes/No**
Pressure Gauge, Discharge	**: Yes**	**- Packing Type**	**:**
Min. Over Pressure above p_v, $\Delta p_b / \Delta p_m$ [bar]/[m liq.]	**: -**	**- Mechanical Seal**	**: Yes/No**
		N.P.S.H. ($\Delta p_m . \rho . g$) [m liq.]	**: 3**

Construction Materials (2)

Pump House	**: MS**	**Wear Rings**	**:**
Pump Rotor	**: HT Steel**	**Shaft Box**	**:**
Shaft	**: HT Steel**		
Special provisions	**: none**		
Operating Pressure [bara]	**:**	**Test Pressure [bara]**	**:**

Remarks :
(1) Further details to be specified by Rotating Equipment specialist.
(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : R.F. Eilers J.R. Law	K.M. de Lathouder B.A. Tulleken	Project ID-Number Date	: **CPD 3228** : 23 April 1999

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Process Systems Engineering

MOLSIEVE BED - SPECIFICATION SHEET

EQUIPMENT NUMBER : S-01/S-02/S-03
NAME : Molsieve Bed

General Data

Function : - Dehydration

Vessel Type : - Vertical

Vessel Contents : - Zeolite 3A
Heating/Cooling medium *(1)* : - None
Vessel Diameter (ID) [m] : 2.53 **Vessel Material : CS**
Vessel Height (T-T) [m] : 6.65
Vessel Tot. Volume [m^3] : 33.5

Process Conditions

Stream Data	Feed <1>		Feed, heating <42>		Feed, cooling <25>					
Temp. [K]	:278		:700		:240		:		:	
Pressure [bara]	:110		:110		:30		:		:	
Density [kg/m^3]	:157.8		:112.7		:32.78		:		:	
Mass Flow [kg/s]	:143.2		:2.43		:0.96		:		:	
Components *(2)*	mol%	wt%	mol%	wt%	mol%	wt%	mol%	wt%	mol%	wt
CH4	81.43	63.28	87.14	76.18	87.14	76.18				
C2H6	8.23	11.99	8.75	14.34	8.75	14.34				
C3H8	3.73	7.97	0.08	0.19	0.08	0.19				
C4	1.76	4.95	0.00	0.00	0.00	0.00				
C5+	1.04	4.06	0.00	0.00	0.00	0.00				
H2O	0.03	0.03	0.00	0.00	0.00	0.00				
Other	3.77	7.72	4.03	9.29	4.03	9.29				

Vessel Contents **Sketch**

Zeolite 3A:

Adsorption capacity	: 0.113	kg/kg
Porosity of bed	: 0.3	-
Density of bed	: 688.79	kg/m3
Cost of molecular sieve	: 3.18	$/kg
Heat capacity	: 1.0467	kJ/kgK

Remarks: *(1)* **During regeneration cycle heating and cooling is provided by the stream entering the vessel. The feed enters the vessel at the top, the regeneration stream at the bottom.**
(2) **C4 includes I- and N-C4H10, C5+ includes I- and N-C5H12 and all heavier components. Other includes N2, CO2, H2S.**

Designers :R.F. Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

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VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : V-01
NAME : L/V Separator 1

General Data

Function : - ~~Buffer~~ / ~~Storage~~ / Separation

Vessel Type : - ~~Horizontal~~ / Vertical

Internals : - Demister/~~Plate~~/~~Coil~~_________

Heating/Cooling medium : - none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~

Vessel Diameter (ID) [m] : 2.84 **Vessel Material:** CS
Vessel Height (T-T) [m] : 8.16
Vessel Tot. Volume [m^3] : 51.6

Process Conditions

Stream Data	**Feed** **<04>**		**Top** **<05>**		**Bottom** **<06>**					
Temp. [K]	: 245.0		: 245.0		: 245.0					
Pressure [bara]	: 72.5		: 72.5		: 72.5					
Density [kg/m^3]	: 131.9		: 114.8		: 339.3					
Mass Flow [kg/s]	: 142.34		: 114.45		: 27.90					
Components	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
CH_4	81.5	63.3	85.8	72.0	52.9	27.7				
C_2H_6	8.2	12.0	7.1	11.2	15.4	15.1				
C_3H_8	3.7	8.0	2.4	5.6	12.4	17.8				
C_4H_{10}	1.8	5.0	0.8	2.3	8.4	15.9				
$C_5H_{12}^+$	1.1	4.1	0.2	0.8	6.7	17.5				
Other	3.7	7.7	3.7	8.1	4.2	6.0				

Vessel Internals

Vane-type demister
- baffle spacing [m]: 0.02
- angle [°] : 45

Feed Insert
- Type : Tube/Ring/______
- Position :
- Diameter [m] :
- Length [m] :
- Material :

Remarks:

Designers : R.F.Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : V-02
NAME : L/V Separator 2

General Data

Function : - ~~Buffer~~ / ~~Storage~~ / Separation

Vessel Type : - ~~Horizontal~~ / Vertical

Internals : - Demister/~~Plate~~/~~Coil~~_________

Heating/Cooling medium : - none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~

Vessel Diameter (ID) [m] : 2.97 **Vessel Material: CS**
Vessel Height (T-T) [m] : 6.80
Vessel Tot. Volume [m^3] : 47.0

Process Conditions

Stream Data	**Feed <09>**	**Top <10>**	**Bottom <11>**		
Temp. [K]	: 207.5	: 207.5	: 207.5		
Pressure [bara]	: 30.0	: 30.0	: 30.0		
Density [kg/m^3]	: 48.8	: 41.3	: 366.3		
Mass Flow [kg/s]	: 114.45	: 94.62	: 19.83		

Components	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**w**
CH4	85.8	72.0	91.0	82.1	34.9	21.2				
C2H6	7.1	11.2	4.9	8.2	56.3	64.0				
C3H8	2.4	5.6	0.6	1.5	0.2	0.4				
C4H10	0.8	2.3	0.1	0.2	0.0	0.0				
C5H12^{+}	0.2	0.8	0.0	0.0	0.0	0.0				
Other	3.7	8.1	3.4	8.0	8.6	14.4				

Vessel Internals

Vane-type demister
- baffle spacing [m]: 0.02
- angle [°] : 45

Feed Insert
- Type : Tube/Ring/______
- Position :
- Diameter [m] :
- Length [m] :
- Material :

Remarks:

Designers : R.F.Eilers K.M. de Lathouder J.R. Law B.A. Tulleken	Project ID-Number : **CPD 3228** Date : 23 April 1999

Technical University Delft
Process Systems Engineering

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : **V-03**
NAME : **L/V Separator 3**

General Data

Function : - ~~Buffer~~ / ~~Storage~~ / Separation

Vessel Type : - ~~Horizontal~~ / Vertical

Internals : - Demister/~~Plate~~/~~Coil~~________

Heating/Cooling medium : - none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~

Vessel Diameter (ID) [m] : 1.65 **Vessel Material:** CS
Vessel Height (T-T) [m] : 10.40
Vessel Tot. Volume [m^3] : 22.32

Process Conditions

Stream Data	**Feed <26>**	**Top <14>**	**Bottom <15>**		
Temp. [K]	: 220.0	: 220.0	: 220.0		
Pressure [bara]	: 30.0	: 30.0	: 30.0		
Density [kg/m^3]	: 81.8	: 44.7	: 363.6		
Mass Flow [kg/s]	: 39.77	: 19.17	: 20.60		

Components	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
CH_4	58.0	40.6	76.9	61.4	34.9	21.2				
C_2H_6	34.6	45.3	16.9	25.2	56.3	64.1				
C_3H_8	0.1	0.2	0.0	0.1	0.2	0.3				
C_4H_{10}	0.0	0.0	0.0	0.0	0.0	0.0				
$C_5H_{12}^+$	0.0	0.0	0.0	0.0	0.0	0.0				
Other	7.3	13.9	6.2	13.4	8.6	14.4				

Vessel Internals

Vane-type demister
- baffle spacing [m]: 0.02
- angle [°] : 45

Feed Insert
- Type : **Tube/Ring/**_____
- Position :
- Diameter [m] :
- Length [m] :
- Material :

Remarks:

Designers : R.F.Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken
Project ID-Number : **CPD 3228**
Date : 23 April 1999

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : V-04
NAME : Reflux accumulator C-03

General Data

Function : - Buffer / ~~Storage~~ / ~~Separation~~

Vessel Type : - ~~Horizontal~~ / Vertical

Internals : - ~~Demister/Plate/Coil~~/None

Heating/Cooling medium : - none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~

Vessel Diameter (ID) [m] : 3.0 **Vessel Material** : CS
Vessel Height (T-T) [m] : 4.6
Vessel Tot. Volume [m^3] : 35.0

Process Conditions

Stream Data	Feed <48>		Bottom <49>							
Temp. [K]	: 318.8		: 318.8							
Pressure [bara]	: 16		: 16							
Density [kg/m^3]	: 452		: 452							
Mass Flow [kg/s]	: 33.77		: 33.77							
Components	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt**
CH4	0.0	0.0	0.0	0.0						
C2H6	1.4	1.0	1.4	1.0						
C3H8	98.0	98.3	98.0	98.3						
C4H10	0.6	0.7	0.6	0.7						
C5H12$^+$	0.0	0.0	0.0	0.0						

Vessel Internals

Coils
Number of tubes :
Tube shape : single/spiral/______
Tube diam. [m^2]:
Tube length [m] :
Tube surf. [m^2]:

Feed Insert
- **Type** : Tube/Ring/_____
- **Position** :
- **Diameter** [m] :
- **Length** [m] :
- **Material** :

Remarks:

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken

Project ID-Number : **CPD 3228**
Date : 23 April 1999

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VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : V-05
NAME : Reflux accumulator C-04

General Data

Function : - Buffer / ~~Storage~~ / ~~Separation~~

Vessel Type : - ~~Horizontal~~ / Vertical

Internals : - ~~Demister/Plate/Coil~~/None

Heating/Cooling medium : - none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~

Vessel Diameter (ID) [m] : 2.3 **Vessel Material** : CS
Vessel Height (T-T) [m] : 4.5
Vessel Tot. Volume [m^3] : 20.0

Process Conditions

Stream Data	Feed <58>		Bottom <59>							
Temp. [K]	: 343.0		: 343.0							
Pressure [bara]	: 9		: 9							
Density [kg/m^3]	: 505		: 505							
Mass Flow [kg/s]	: 20.75		: 20.75							
Components	mol%	wt%	mol%	wt%	mol%	wt%	mol%	wt%	mol%	wt%
CH_4	0.0	0.0	0.0	0.0						
C_2H_6	0.0	0.0	0.0	0.0						
C_3H_8	0.2	0.1	0.2	0.1						
C_4H_{10}	98.7	98.5	98.7	98.5						
$C_5H_{12}^+$	1.1	1.4	1.1	1.4						
Rest	0.0	0.0	0.0	0.0						

Vessel Internals

Coils
Number of tubes :
Tube shape : single/spiral/______
Tube diam. [m^2]:
Tube length [m] :
Tube surf. [m^2]:

Feed Insert
- Type : Tube/Ring/_____
- Position :
- Diameter [m] :
- Length [m] :
- Material :

Remarks:

Designers : R.F. Eilers K.M. de Lathouder
J.R. Law B.A. Tulleken

Project ID-Number : **CPD 3228**
Date : 23 April 1999

Appendix 5-1

PURE COMPONENT PROPERTIES

Component Name		Technological Data					Medical Data		Notes
Design	Systematic	Formula	Mol. Weight	Boiling Point (1)	Melting Point (1)	Density of Liquid (2)	MAC value	LC50 (3)	
			g/mol	oC	oC	kg/m3	mg/m3	mg/m3	
N2	Nitrogen	N2	28.01	-195.8	-209.9	808.1	n.a.		(7)
CH4	Methane	CH4	16.04	-164.0	-182.0	466.0	n.a.		(8)
C2H6	Ethane	C2H6	30.07	-88.6	-183.3	572.0	n.a.		(9)
CO2	Carbondioxide	CO2	44.01	-78.5	-56.6	1101.0	9100		(5) (6) (10)
H2S	Hydrogensulfide	H2S	34.08	-60.7	-85.5	1.539	14	987	(13)
C3H8	Propane	C3H8	44.10	-42.1	-189.7	585.3	n.a.		(11)
I-C4	I-Butane	C4H10	58.12	-11.7	-159.4	549.0	2400		
N-C4	N-Butane	C4H10	58.12	-0.5	-138.4	579.0	2400		(4)
I-C5	I-Pentane	C5H12	72.15	27.8		0.625	3000		(12)
N-C5	N-Pentane	C5H12	72.15	36.1	-130.0	626.2	3000		(4)
N-C6	N-Hexane	C6H14	86.18	69.0	-95.0	660.3	180		(4)
N-C7	N-Heptane	C7H16	100.20	98.4	-90.6	683.7	2100		(4)
H2O	Water	H2O	18.02	100.0	0.0	1000.0	n.a.		
N-C8	N-Octane	C8H18	114.23	125.7	-56.8	702.5	2400		(4)

Notes :

(1)	At 101.3 kPa
(2)	Density at 25 °C, unless specified otherwise
(3)	For 30 min exposure
(4)	Density at 20 °C from the density of H2O at 4 °C
(5)	Boiling temperature is the sublimation temperature
(6)	Melting point at 5.2 atm.
(7)	Density at -195.8 °C
(8)	Density at -164 °C
(9)	Density at -100 °C from the density of H2O at 4 °C
(10)	Density at -37 °C
(11)	Density at -45 °C from the density of H2O at 4 °C
(12)	Density at 16 °C relative to the density of H2O at 16 °C
(13)	Density of gas at 0 °C

Project ID Number	:	**CPD 3228**
Completion Date	:	23 April 1999

PROCESS STREAM SUMMARY

STREAM Nr.	:	1	IN	2		3		4		5	
	Name :	Feed		Dehydrated feed		Expanded K01		Cooled feed		Vapour V01	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.78	0.0277	0.77	0.0276	0.77	0.0276	0.77	0.0276	0.74	0.0265
CH4	16.04	90.59	5.6479	90.11	5.6179	90.11	5.6179	90.11	5.6179	82.39	5.1366
C2H6	30.07	17.16	0.5707	17.07	0.5676	17.07	0.5676	17.07	0.5676	12.85	0.4275
CO2	44.01	10.27	0.2333	10.21	0.2321	10.21	0.2321	10.21	0.2321	8.56	0.1946
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	11.41	0.2586	11.34	0.2572	11.34	0.2572	11.34	0.2572	6.38	0.1447
I-C4H10	58.12	2.54	0.0437	2.53	0.0434	2.53	0.0434	2.53	0.0434	1.03	0.0176
N-C4H10	58.12	4.55	0.0784	4.53	0.0779	4.53	0.0779	4.53	0.0779	1.59	0.0274
I-C5H12	72.15	1.63	0.0225	1.62	0.0224	1.62	0.0224	1.62	0.0224	0.37	0.0051
N-C5H12	72.15	1.65	0.0229	1.64	0.0228	1.64	0.0228	1.64	0.0228	0.34	0.0048
N-C6H14	86.18	1.46	0.0170	1.46	0.0169	1.46	0.0169	1.46	0.0169	0.14	0.0016
N-C7H16	100.20	0.83	0.0083	0.83	0.0083	0.83	0.0083	0.83	0.0083	0.04	0.0004
H2O	18.02	0.04	0.0024	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.24	0.0021	0.24	0.0021	0.24	0.0021	0.24	0.0021	0.01	0.0000
Total		143.16	6.9357	142.34	6.8962	142.34	6.8962	142.34	6.8962	114.45	5.9869
Enthalpy	MW	-648		-644		-648		-656		-549	
Phase		V		V		L/V		L/V		V	
Press.	Bara	110.0		109.6		72.5		72.2		72.2	
Temp	K	278.2		278.2		257.1		245.0		245.0	

STREAM Nr.	:	6		7		8		9		10	
	Name :	Liquids V01		Expanded Liq. V01		Feed C02-lower		Expanded K03		Vapour V02	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.03	0.0011	0.03	0.0011	0.03	0.0011	0.74	0.0265	0.73	0.0262
CH4	16.04	7.72	0.4813	7.72	0.4813	7.72	0.4813	82.39	5.1366	77.65	4.8408
C2H6	30.07	4.21	0.1402	4.21	0.1402	4.21	0.1402	12.85	0.4275	7.80	0.2595
CO2	44.01	1.65	0.0375	1.65	0.0375	1.65	0.0375	8.56	0.1946	6.85	0.1556
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	4.96	0.1125	4.96	0.1125	4.96	0.1125	6.38	0.1447	1.42	0.0322
I-C4H10	58.12	1.50	0.0258	1.50	0.0258	1.50	0.0258	1.03	0.0176	0.08	0.0014
N-C4H10	58.12	2.93	0.0505	2.93	0.0505	2.93	0.0505	1.59	0.0274	0.08	0.0014
I-C5H12	72.15	1.25	0.0173	1.25	0.0173	1.25	0.0173	0.37	0.0051	0.01	0.0001
N-C5H12	72.15	1.30	0.0180	1.30	0.0180	1.30	0.0180	0.34	0.0048	0.00	0.0001
N-C6H14	86.18	1.32	0.0153	1.32	0.0153	1.32	0.0153	0.14	0.0016	0.00	0.0000
N-C7H16	100.20	0.79	0.0079	0.79	0.0079	0.79	0.0079	0.04	0.0004	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.23	0.0020	0.23	0.0020	0.23	0.0020	0.01	0.0000	0.00	0.0000
Total		27.90	0.9094	27.90	0.9094	27.90	0.9094	114.45	5.9869	94.62	5.3172
Enthalpy	MW	-106		-106		-104		-554		-473	
Phase		L		L/V		L/V		L/V		V	
Press.	Bara	72.2		30.3		30.0		30.0		30.0	
Temp	K	245.0		223.7		250.0		207.5		207.5	

STREAM Nr. :		11		12		13		14		15	
Name :		Liquid V02		Liquid V02		Feed C02-second		Vapour V03		Liquid V03	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.01	0.0004	0.01	0.0004	0.01	0.0004	0.04	0.0014	0.00	0.0002
CH4	16.04	4.74	0.2958	4.74	0.2958	4.74	0.2958	11.77	0.7336	4.36	0.2720
C2H6	30.07	5.05	0.1680	5.05	0.1680	5.05	0.1680	4.83	0.1608	13.20	0.4389
CO2	44.01	1.71	0.0390	1.71	0.0390	1.71	0.0390	2.53	0.0574	2.96	0.0672
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	4.96	0.1125	4.96	0.1125	4.96	0.1125	0.01	0.0001	0.07	0.0017
I-C4H10	58.12	0.95	0.0163	0.95	0.0163	0.95	0.0163	0.00	0.0000	0.00	0.0000
N-C4H10	58.12	1.51	0.0260	1.51	0.0260	1.51	0.0260	0.00	0.0000	0.00	0.0000
I-C5H12	72.15	0.36	0.0050	0.36	0.0050	0.36	0.0050	0.00	0.0000	0.00	0.0000
N-C5H12	72.15	0.34	0.0047	0.34	0.0047	0.34	0.0047	0.00	0.0000	0.00	0.0000
N-C6H14	86.18	0.14	0.0016	0.14	0.0016	0.14	0.0016	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.04	0.0004	0.04	0.0004	0.04	0.0004	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.01	0.0000	0.01	0.0000	0.01	0.0000	0.00	0.0000	0.00	0.0000
Total		19.82	0.6697	19.82	0.6697	19.82	0.6697	19.17	0.9534	20.60	0.7799
Enthalpy	MW	-82		-82		-79		-95		-94	
Phase		L		L		L/V		V		L	
Press.	Bara	30.0		30.6		30.3		29.7		29.7	
Temp	K	207.5		207.5		235.0		220.0		220.0	

STREAM Nr. :		16		17		18		19		20	
Name :		Liquid V03		Vapour Feed C01		Liquid Feed C01		Reflux C02		Vapour C01	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.00	0.0002	0.77	0.0276	0.00	0.0001	0.00	0.0001	0.77	0.0276
CH4	16.04	4.36	0.2720	89.41	5.5744	2.05	0.1278	2.31	0.1442	90.11	5.6179
C2H6	30.07	13.20	0.4389	12.64	0.4202	6.20	0.2063	6.99	0.2326	16.96	0.5640
CO2	44.01	2.96	0.0672	9.37	0.2130	1.39	0.0316	1.57	0.0356	10.21	0.2321
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	0.07	0.0017	1.42	0.0323	0.04	0.0008	0.04	0.0009	0.22	0.0051
I-C4H10	58.12	0.00	0.0000	0.08	0.0014	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C4H10	58.12	0.00	0.0000	0.08	0.0014	0.00	0.0000	0.00	0.0000	0.00	0.0000
I-C5H12	72.15	0.00	0.0000	0.01	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C5H12	72.15	0.00	0.0000	0.00	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C6H14	86.18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		20.60	0.7799	113.80	6.2705	9.68	0.3666	10.92	0.4134	118.28	6.4467
Enthalpy	MW	-94		-568		-44		-50		-590	
Phase		L		V		L		L		V	
Press.	Bara	30.8		29.7		30.8		30.8		30.0	
Temp	K	220.0		209.5		220.0		220.0		206.1	

STREAM Nr. :		21		22		23		24		25	
Name :		Liquid C01		Liquid C01		Vapour C02		Hot SG, Pre Comp.		Cooling Molsieves*	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.00	0.0001	0.00	0.0001	0.04	0.0016	0.77	0.0276	0.01	0.0002
CH4	16.04	1.35	0.0844	1.35	0.0844	16.13	1.0056	90.11	5.6179	0.73	0.0457
C2H6	30.07	1.88	0.0626	1.88	0.0626	18.03	0.5997	16.96	0.5640	0.14	0.0046
CO2	44.01	0.55	0.0125	0.55	0.0125	5.48	0.1246	10.21	0.2321	0.08	0.0019
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	1.24	0.0280	1.24	0.0280	0.08	0.0018	0.22	0.0051	0.00	0.0000
I-C4H10	58.12	0.08	0.0014	0.08	0.0014	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C4H10	58.12	0.08	0.0014	0.08	0.0014	0.00	0.0000	0.00	0.0000	0.00	0.0000
I-C5H12	72.15	0.01	0.0001	0.01	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C5H12	72.15	0.00	0.0001	0.00	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C6H14	86.18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		5.20	0.1905	5.20	0.1905	39.77	1.7333	118.28	6.4467	0.96	0.0524
Enthalpy	MW	-23		-23		-182		-579		-5	
Phase		L		L		V		V		V	
Press.	Bara	30.2		30.7		30.0		29.7		29.7	
Temp	K	208.7		208.7		240.2		240.2		240.2	

* if valve is open for cooling cycle, otherwise no flow

STREAM Nr. :		26		27		28		29		30	
Name :		Feed V03		Comp. K03 Feed		Compressed K03		Comp. K01 Feed**		Compressed K01**	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.04	0.0016	0.47	0.0168	0.47	0.0168	0.30	0.0108	0.30	0.0108
CH4	16.04	16.13	1.0056	54.97	3.4269	54.97	3.4269	35.14	2.1910	35.14	2.1910
C2H6	30.07	18.03	0.5997	10.34	0.3440	10.34	0.3440	6.61	0.2199	6.61	0.2199
CO2	44.01	5.48	0.1246	6.23	0.1416	6.23	0.1416	3.98	0.0905	3.98	0.0905
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	0.08	0.0018	0.14	0.0031	0.14	0.0031	0.09	0.0020	0.09	0.0020
I-C4H10	58.12	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C4H10	58.12	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
I-C5H12	72.15	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C5H12	72.15	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C6H14	86.18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		39.77	1.7333	72.15	3.9325	72.15	3.9325	46.13	2.5142	46.13	2.5142
Enthalpy	MW	-190		-353		-348		-226		-223	
Phase		L/V		V		V		V		V	
Press.	Bara	29.7		29.7		52.4		29.7		52.4	
Temp	K	220.0		240.2		284.7		240.2		284.8	

** if valve for cooling is closed

STREAM Nr. :		31		32		33		34	OUT	35	
Name :		K02 Feed		Hot Sales Gas		Split Sales Gas***		Sales Gas***		Regen. Sales Gas***	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.77	0.0276	0.77	0.0276	0.76	0.0270	0.76	0.0270	0.02	0.0006
CH4	16.04	90.11	5.6179	90.11	5.6179	88.26	5.5025	88.26	5.5025	1.85	0.1154
C2H6	30.07	16.96	0.5640	16.96	0.5640	16.61	0.5524	16.61	0.5524	0.35	0.0116
CO2	44.01	10.21	0.2321	10.21	0.2321	10.01	0.2273	10.01	0.2273	0.21	0.0048
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	0.22	0.0051	0.22	0.0051	0.22	0.0050	0.22	0.0050	0.00	0.0001
I-C4H10	58.12	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C4H10	58.12	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
I-C5H12	72.15	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C5H12	72.15	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C6H14	86.18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		118.28	6.4467	118.28	6.4467	115.85	6.3142	115.85	6.3142	2.43	0.1324
Enthalpy	MW	-571		-547		-535		-561		-11	
Phase		V		V		V		V		V	
Press.	Bara	52.4		190.1		190.1		190.0		190.1	
Temp	K	284.7		394.3		394.3		322.2		394.3	

*** if valve for regeneration (heating) is open

STREAM Nr. :		36		37		38		39		40	OUT
Name :		Fuel Gas		Split Regen. Gas		Fuel Gas, expanded		Regen. Gas, Exp.		Fuel/Flare Gas, mixed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	N.N.F.		0.02	0.0006	N.N.F.		0.02	0.0006	0.02	0.0007
CH4	16.04			1.85	0.1154			1.85	0.1154	2.33	0.1454
C2H6	30.07			0.35	0.0116			0.35	0.0116	0.44	0.0147
CO2	44.01			0.21	0.0048			0.21	0.0048	0.26	0.0060
H2S	34.08			0.00	0.0000			0.00	0.0000	0.00	0.0000
C3H8	44.10			0.00	0.0001			0.00	0.0001	0.07	0.0015
I-C4H10	58.12			0.00	0.0000			0.00	0.0000	0.01	0.0002
N-C4H10	58.12			0.00	0.0000			0.00	0.0000	0.03	0.0004
I-C5H12	72.15			0.00	0.0000			0.00	0.0000	0.01	0.0001
N-C5H12	72.15			0.00	0.0000			0.00	0.0000	0.01	0.0001
N-C6H14	86.18			0.00	0.0000			0.00	0.0000	0.01	0.0001
N-C7H16	100.20			0.00	0.0000			0.00	0.0000	0.01	0.0001
H2O	18.02			0.00	0.0000			0.00	0.0000	0.04	0.0024
N-C8H18	114.23			0.00	0.0000			0.00	0.0000	0.00	0.0000
Total		N.N.F.		2.43	0.1324	N.N.F.		2.43	0.1324	3.24	0.1719
Enthalpy	MW			-11				-11		-12	
Phase		V		V		V		V		V	
Press.	Bara	190.1		190.1		110.0		110.0		110.0	
Temp	K	390.2		394.3		354.7		381.4		370.0	

*** if valve for regeneration (heating) is open

**** held up water and other hydrocarbons are assumed to flow out of the bed at a constant rate

STREAM Nr. :		41		42		43		44		45	
Name :		Fuel/Flare Gas****		Hot Regen. Gas		Dry Regen. Gas***		Bottoms C02		Feed C03	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.02	0.0007	0.02	0.0006	0.02	0.0006	0.00	0.0000	0.00	0.0000
CH4	16.04	2.33	0.1454	1.85	0.1154	1.85	0.1154	0.00	0.0000	0.00	0.0000
C2H6	30.07	0.44	0.0147	0.35	0.0116	0.35	0.0116	0.11	0.0037	0.11	0.0037
CO2	44.01	0.26	0.0060	0.21	0.0048	0.21	0.0048	0.00	0.0000	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	0.07	0.0015	0.00	0.0001	0.00	0.0001	11.12	0.2521	11.12	0.2521
I-C4H10	58.12	0.01	0.0002	0.00	0.0000	0.00	0.0000	2.53	0.0434	2.53	0.0434
N-C4H10	58.12	0.03	0.0004	0.00	0.0000	0.00	0.0000	4.53	0.0779	4.53	0.0779
I-C5H12	72.15	0.01	0.0001	0.00	0.0000	0.00	0.0000	1.62	0.0224	1.62	0.0224
N-C5H12	72.15	0.01	0.0001	0.00	0.0000	0.00	0.0000	1.64	0.0228	1.64	0.0228
N-C6H14	86.18	0.01	0.0001	0.00	0.0000	0.00	0.0000	1.46	0.0169	1.46	0.0169
N-C7H16	100.20	0.01	0.0001	0.00	0.0000	0.00	0.0000	0.83	0.0083	0.83	0.0083
H2O	18.02	0.04	0.0024	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.24	0.0021	0.24	0.0021
Total		3.24	0.1719	2.43	0.1324	2.43	0.1324	24.06	0.4496	24.06	0.4496
Enthalpy	MW	-12		-9		-9		-57		-57	
Phase		V		V		V		L		L/V	
Press.	Bara	109.6		110.0		110.0		30.3		16.0	
Temp	K	370.0		700.0		700.0		377.9		349.0	

STREAM Nr. :		46		47		48=49		50		51	
Name :		Vapour C03		Bottoms C03		Cond. Vap. C03		Reflux C03		Reflux C03	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
CH4	16.04	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C2H6	30.07	0.33	0.0110	0.00	0.0000	0.33	0.0110	0.22	0.0074	0.22	0.0074
CO2	44.01	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	33.19	0.7526	0.01	0.0003	33.19	0.7526	22.08	0.5007	22.08	0.5007
I-C4H10	58.12	0.23	0.0039	2.45	0.0421	0.23	0.0039	0.15	0.0026	0.15	0.0026
N-C4H10	58.12	0.02	0.0004	4.52	0.0778	0.02	0.0004	0.02	0.0003	0.02	0.0003
I-C5H12	72.15	0.00	0.0000	1.62	0.0224	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C5H12	72.15	0.00	0.0000	1.64	0.0228	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C6H14	86.18	0.00	0.0000	1.46	0.0169	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.00	0.0000	0.83	0.0083	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.24	0.0021	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		33.77	0.7680	12.76	0.1926	33.77	0.7680	22.47	0.5110	22.47	0.5110
Enthalpy	MW	-81		-28		-91		-60		-60	
Phase		V		L		L		L		L	
Press.	Bara	16.0		16.5		16.0		16.0		17.5	
Temp	K	318.8		389.7		318.8		318.8		318.8	

STREAM Nr.	:	52		53		54	OUT	55		56	
Name :		LP Hot C3 Product		HP Hot C3 Product		C3 Product		Feed C04		Vapour C04	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
CH4	16.04	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C2H6	30.07	0.11	0.0037	0.11	0.0037	0.11	0.0037	0.00	0.0000	0.00	0.0000
CO2	44.01	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	11.11	0.2519	11.11	0.2519	11.11	0.2519	0.01	0.0003	0.03	0.0007
I-C4H10	58.12	0.08	0.0013	0.08	0.0013	0.08	0.0013	2.45	0.0421	7.24	0.1245
N-C4H10	58.12	0.01	0.0001	0.01	0.0001	0.01	0.0001	4.52	0.0778	13.20	0.2272
I-C5H12	72.15	0.00	0.0000	0.00	0.0000	0.00	0.0000	1.62	0.0224	0.19	0.0027
N-C5H12	72.15	0.00	0.0000	0.00	0.0000	0.00	0.0000	1.64	0.0228	0.08	0.0011
N-C6H14	86.18	0.00	0.0000	0.00	0.0000	0.00	0.0000	1.46	0.0169	0.00	0.0000
N-C7H16	100.20	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.83	0.0083	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.24	0.0021	0.00	0.0000
Total		11.30	0.2570	11.30	0.2570	11.30	0.2570	12.76	0.1926	20.75	0.3562
Enthalpy	MW	-30		-30		-31		-28		-45	
Phase		L		L		L		L/V		V	
Press.	Bara	16.0		21.0		21.0		9.0		9.0	
Temp	K	318.8		319.5		310.2		363.5		343.0	

STREAM Nr.	:	57		58=59		60		61	OUT	62	
Name :		Bottoms C04		Cond. Vap. C04		Hot C4 Product		C4 Product*****		Reflux C04	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
CH4	16.04	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C2H6	30.07	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
CO2	44.01	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	0.00	0.0000	0.03	0.0007	0.01	0.0003	0.01	0.0003	0.02	0.0005
I-C4H10	58.12	0.00	0.0000	7.24	0.1245	2.45	0.0421	2.45	0.0421	4.79	0.0824
N-C4H10	58.12	0.05	0.0009	13.20	0.2272	4.47	0.0769	4.47	0.0769	8.74	0.1503
I-C5H12	72.15	1.55	0.0215	0.19	0.0027	0.06	0.0009	0.06	0.0009	0.13	0.0018
N-C5H12	72.15	1.61	0.0224	0.08	0.0011	0.03	0.0004	0.03	0.0004	0.05	0.0007
N-C6H14	86.18	1.46	0.0169	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C7H16	100.20	0.83	0.0083	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
H2O	18.02	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.24	0.0021	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		5.74	0.0721	20.75	0.3562	7.02	0.1205	7.02	0.1205	13.73	0.2357
Enthalpy	MW	-12		-51		-17		-18		-34	
Phase		L		L		L		L		L	
Press.	Bara	9.3		9.0		10.2		9.0		10.2	
Temp	K	400.8		343.0		343.0		310.2		343.0	

***** with use of valve (not shown)

STREAM Nr. :		63		64	OUT
	Name :	Cooled C5+ Prod.		C5+ Product	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.00	0.0000	0.00	0.0000
CH4	16.04	0.00	0.0000	0.00	0.0000
C2H6	30.07	0.00	0.0000	0.00	0.0000
CO2	44.01	0.00	0.0000	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000
C3H8	44.10	0.00	0.0000	0.00	0.0000
I-C4H10	58.12	0.00	0.0000	0.00	0.0000
N-C4H10	58.12	0.05	0.0009	0.05	0.0009
I-C5H12	72.15	1.55	0.0215	1.55	0.0215
N-C5H12	72.15	1.61	0.0224	1.61	0.0224
N-C6H14	86.18	1.46	0.0169	1.46	0.0169
N-C7H16	100.20	0.83	0.0083	0.83	0.0083
H2O	18.02	0.00	0.0000	0.00	0.0000
N-C8H18	114.23	0.24	0.0021	0.24	0.0021
Total		5.74	0.0721	5.74	0.0721
Enthalpy	MW	-13		-13	
Phase		L		L	
Press.	Bara	9.3		8.0	
Temp	K	310.2		310.2	

Overall Component Mass Balance & Stream Heat balance							
STREAM Nr. :		1		34+40+54+61+64			
	Name :	Total Plant, In		Total Plant, Out		Total Plant, Out-In	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
N2	28.01	0.78	0.0277	0.78	0.0277	0.00	0.0000
CH4	16.04	90.59	5.6479	90.59	5.6479	0.00	0.0000
C2H6	30.07	17.16	0.5707	17.16	0.5707	0.00	0.0000
CO2	44.01	10.27	0.2333	10.27	0.2334	0.00	0.0000
H2S	34.08	0.00	0.0000	0.00	0.0000	0.00	0.0000
C3H8	44.10	11.41	0.2586	11.41	0.2586	0.00	0.0000
I-C4H10	58.12	2.54	0.0437	2.54	0.0437	0.00	0.0000
N-C4H10	58.12	4.55	0.0784	4.55	0.0784	0.00	0.0000
I-C5H12	72.15	1.63	0.0225	1.63	0.0225	0.00	0.0000
N-C5H12	72.15	1.65	0.0229	1.65	0.0229	0.00	0.0000
N-C6H14	86.18	1.46	0.0170	1.46	0.0170	0.00	0.0000
N-C7H16	100.20	0.83	0.0083	0.83	0.0083	0.00	0.0000
H2O	18.02	0.04	0.0024	0.04	0.0024	0.00	0.0000
N-C8H18	114.23	0.24	0.0021	0.24	0.0021	0.00	0.0000
Total		143.16	6.9357	143.16	6.9357	0.00	0.0000
Press.	Bara	110.0					
Temp	K	278.2					
Enthalpy	MW	-648		-635		13 ******	

****** For heating cycle. Total utilities-input is 11 MW, the 2 extra come from burning <40>

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HEAT & MASS BALANCE FOR STREAMS TOTAL

IN						OUT				
Plant		EQUIPMENT			EQUIPM. IDENTIF.	EQUIPMENT			Plant	
Mass kg/s	Heat MW	Mass kg/s	Heat MW	Stream Nr.		Stream Nr.	Mass kg/s	Heat MW	Mass kg/s	Hea MW
		113.80	-568	<17>	**C01**	<20>	118.28	-590		
		9.68	-44	<18>		<21>	5.20	-23		
		123.48	-612		Total		123.48	-612		
		10.92	-50	<19>	**C02**	<23>	39.77	-182		
		5.20	-23	<22>		<44>	24.06	-57		
		19.82	-79	<13>						
		27.90	-104	<8>						
	17		17		**E02**					
		63.83	-238		Total		63.83	-238		
		24.06	-57	<45>	**C03**	<46>	33.77	-81		
		22.47	-60	<51>		<47>	12.76	-28		
	8		8		**E04**					
		46.53	-109		Total		46.53	-109		
		12.76	-28	<55>	**C04**	<56>	20.75	-45		
		13.73	-34	<62>		<57>	5.74	-12		
	5		5		**E07**					
		26.49	-57		Total		26.49	-57		
		118.28	-590	<20>	**E01**	<24>	118.28	-579		
		39.77	-182	<23>		<26>	39.77	-190		
		142.34	-648	<3>		<4>	142.34	-656		
		27.90	-106	<7>		<8>	27.90	-104		
		19.82	-82	<12>		<13>	19.82	-79		
		348.12	-1607		Total		348.12	-1607		
		33.77	-81	<46>		<48>	33.77	-91		
					E03			10		
		33.77	-81		Total		33.77	-91		
		115.85	-535	<33>		<34>	115.85	-561	**115.85**	**-5**
					E05			25		
		115.85	-535		Total		115.85	-535		
		20.75	-45	<56>		<58>	20.75	-51		
					E06			6		
		20.75	-45		Total		20.75	-45		
		5.74	-12	<57>		<63,64>	5.74	-13	**5.74**	
					E08			1		
		5.74	-12		Total		5.74	-12		
		11.30	-30	<53>		<54>	11.30	-31	**11.30**	
					E09			0		
		11.30	-30		Total		11.30	-30		
		7.02	-17	<60>		<61>	7.02	-18	**7.02**	
					E10			1		
		7.02	-17		Total		7.02	-17		
		142.34	-644	<2>		<3>	142.34	-648		
		46.13	-226	<29>		<30>	46.13	-223		
					K01					
		188.47	-870		Total		188.47	-870		

		118.28	-571	<31>		<32>	118.28	-547		
	24		24		**K02**					
		118.28	-547		Total		118.28	-547		
		114.45	-549	<5>		<9>	114.45	-554		
		72.15	-353	<27>		<28>	72.15	-348		
					K03					
		186.60	-903		Total		186.60	-903		
		19.82	-82	<11>	**P01**	<12>	19.82	-82		
		19.82	-82		Total		19.82	-82		
		20.60	-94	<15>	**P02**	<16>	20.60	-94		
		20.60	-94		Total		20.60	-94		
		5.20	-23	<21>	**P03**	<22>	5.20	-23		
		5.20	-23		Total		5.20	-23		
		22.47	-60	<50>	**P04**	<51>	22.47	-60		
		22.47	-60		Total		22.47	-60		
		11.30	-30	<52>	**P05**	<53>	11.30	-30		
		11.30	-30		Total		11.30	-30		
		20.75	-51	<59>	**P06**	<60,62>	20.75	-51		
		20.75	-51		Total		20.75	-51		
143.16	**-648**	143.16	-648	<1>	**S01-S03**	<2>	142.34	-644		
		2.43	-9	<43>		<41>	3.24	-12		
		145.59	-657		Total		145.59	-657		
		142.34	-656	<4>	**V01**	<5>	114.45	-549		
						<6>	27.90	-106		
		142.34	-656		Total		142.34	-656		
		114.45	-554	<9>	**V02**	<10>	94.62	-473		
						<11>	19.82	-82		
		114.45	-554		Total		114.45	-554		
		39.77	-190	<26>	**V03**	<14>	19.17	-95		
						<15>	20.60	-94		
		39.77	-190		Total		39.77	-190		
		33.77	-91	<48>	**V04**	<49>	33.77	-91		
		33.77	-91		Total		33.77	-91		
		20.75	-51	<58>	**V05**	<59>	20.75	-51		
		20.75	-51		Total		20.75	-51		
		2.43	-11	<39>	**F01**	<42>	2.43	-9		
		3.24	-12	<40>		<40>*	3.24	-15	3.24	-15
		5.67	-24		Total		5.67	-24		
143.16	**-593**				Total				**143.16**	**-594**
OUT - IN :									0.00	0

* burned, back calculation, air not included

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SUMMARY OF UTILITIES

EQUIPMENT		UTILITIES													REMARKS
		Heating					Cooling				Power				
Nr.	Name	Load MW	Consumption (t/h) Steam LP	MP	HP	Hot Oil	Load MW	Consumption (t/h) Cooling Water	Air	Refrig.	Actual Load MW	Consumption (t/h, kWh/h) Steam (t/h) HP	MP	Electr. kWh/h	
E02	Reboiler C02	16.71	27.34												
E03	Condensor C03						10.01	1077.04							
E04	Reboiler C03	8.15	13.33												
E05	Sales Gas Cooler						25.26	2763.45							For product cooling only.
E06	Condensor C04						6.22	669.85							
E07	Reboiler C04	5.42	8.87												
E08	C5+ Cooler						1.34	143.46							For product cooling only.
E09	C3 Cooler						0.32	34.79							For product cooling only.
E10	C4 Cooler						0.62	66.30							For product cooling only.
K02	Sales Gas Comp.										24.05			2.41E+04	
P01	Bottoms V03										0.01			11.81	
P02	Bottoms V04										0.01			7.58	
P03	Bottoms C01										0.00			1.05	
P04	Reflux C03										0.01			9.23	
P05	C3 Pump										0.02			16.45	
P06	Reflux C04										0.01			6.26	
TOTAL		30.28	**49.54**				43.78	**4.75E+03**			24.10			**2.41E+04**	

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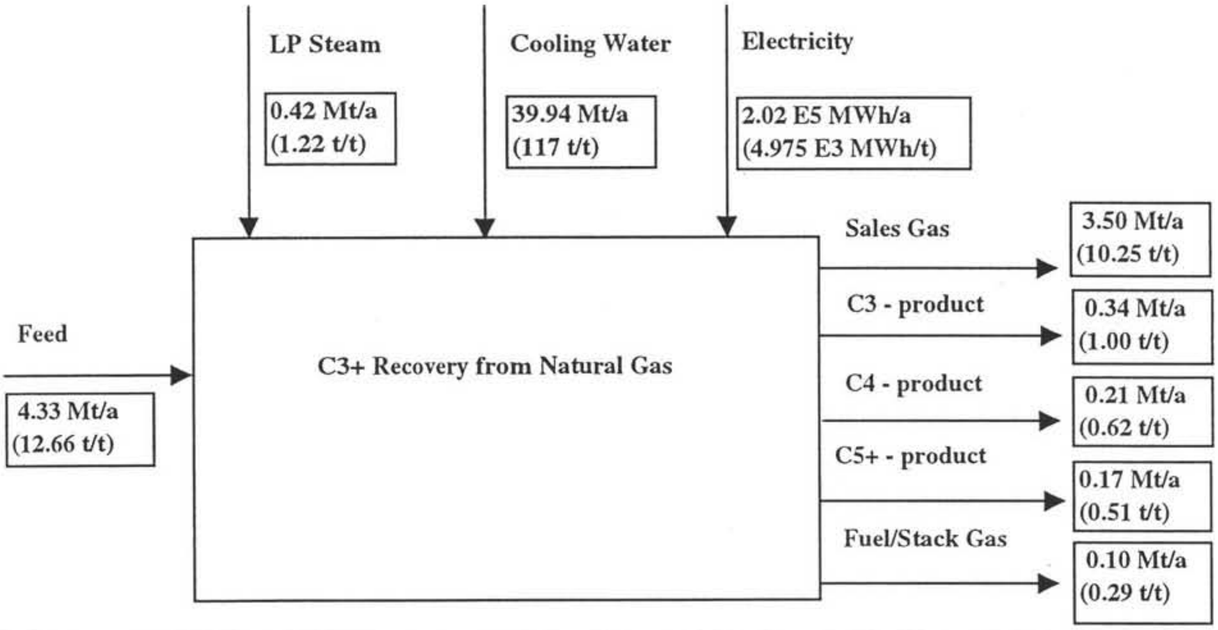
Appendix 5-5: PROCESS YIELDS AND SIMPLIFIED BLOCKSCHEME

Process Streams							
Name	Ref. Stream	kg/s		Mt/a		t/t C3 Product	
		IN	OUT	IN	OUT	IN	OUT
Feed	<1>	143.16		4.329		12.66	
Sales Gas*	<34>		115.85		3.503		10.25
C3 Product	<54>		11.30		0.342		1.00
C4 Product	<61>		7.02		0.212		0.62
C5+ Product	<64>		5.74		0.174		0.51
Fuel/Stack Gas**	<40>		3.24		0.098		0.29
Total		143.16	143.16	4.329	4.329	12.66	12.67

* For heating cycle, more for cooling cycle

** For heating cycle, less for cooling cycle

Utilities							
Name	Ref. Stream	kg/s	MW	Mt/a	MWh/a	t/t C3	MWh/ t C3
LP Steam	-	13.76		0.42		1.22	
CW	-	1.32E+03		39.94		116.85	
Electricity	-		24.10		2.02E+05		4975.42



COMMENTS :

1. The amount of cooling water is enormous. Therefore it may be economical to install an external cooling loop to cool the water to beneath 303 K and be able to return it above 311 K. Even only returning the water at 313 K reduces the amount of cooling water needed by ~20 %. It also may be possible to use stream <45> and <55> to partially cool the sales gas or C5+-product.
2. The amount of electricity required is also enormous. This is due to the need to compress the sales gas to 190 bara. Over 99 % of the electricity is used for this. Most savings in the process will be achieved by reducing the compressor duty.
3. Quantities of C4- and C5+-product only vary with their quantities in the feed, not with the recovery of C3-product.
4. It can be seen that the fuel/stack gas is a large portion of the feed/sales gas, but is necessary to regenerate the molsieve bed. Due to excess heat in the stack the temperature of the regenerating gas could be raised, thereby decreasing the amount of regenerating gas needed.

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Appendix 5-6: Utilities

1. Available Utilities:

a. Steam

Superheated steam is considered to be available at the temperatures given below:

Conditions	Steam Class	
	Medium Pressure	Low Pressure
T(condensation) [°C] :	175	150

Fouling coefficient: 10 kW/m$^{2\circ}$C Fouling factor: 0.1 m$^{2\circ}$C/kW

b. Electricity

Power	Voltage	Current
Low	220	AC
Medium	380	three-phase AC
High	3000 – 10000	three-phase AC

c. Water

Water	T [°C]		H	Fouling factor
	in	Out	[kW/m$^{2\circ}$C]	[m$^{2\circ}$C/kW]
Cooling	30	38$^{(1)}$	2.0	0.5
Remarks:	(1) Maximum allowed			

d. Air

Air-cooling:

Conditions	Value
T(design) [°C]	30

2. Utility Costs:

Utility costs[1]						
Utility	Units		Unit Costs, Dfl/unit			
	Quantity	Energy	Quantity		Energy	
			Min.	Max.	Min.	Max.
Steam LP/MP	Ton		30	35		
Electricity[2]		kWh			0.13	0.22
Cooling Water	m^3		0.05	0.10		
Remarks 1. Ref. Prijzenboekje, WEBSI/DACE, 17^e Ed., May 1994 2. For quantities larger than 10^6 kWh/a						

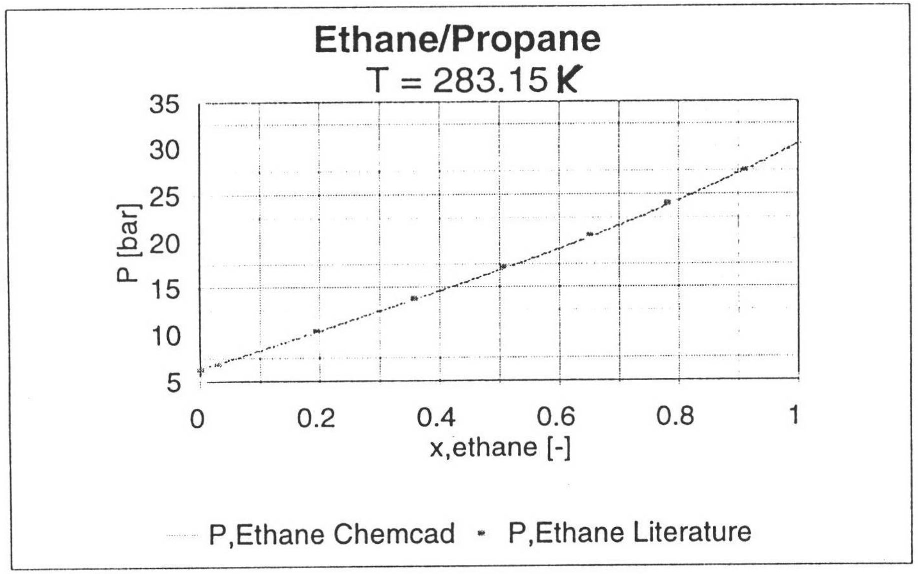


Figure 1: PX-diagram of ethane/propane, T=283.15 K

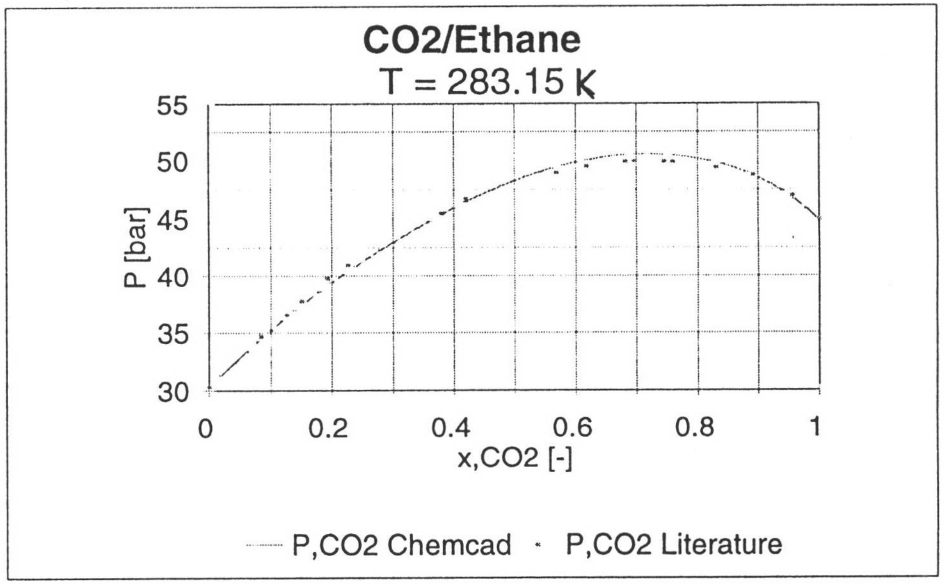


Figure 3: PX-diagram of CO2/ethane, T=283.15 K

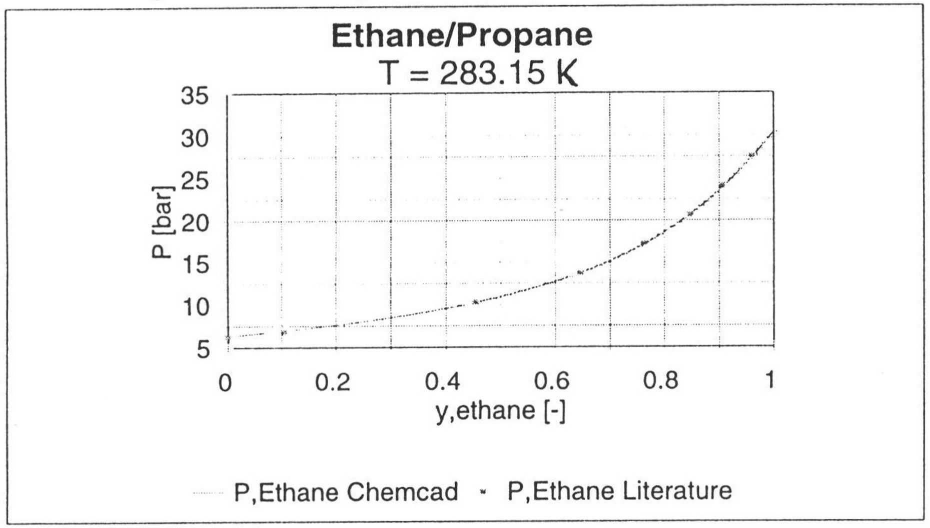


Figure 2: PY-diagram of ethane/propane, T=283.15 K

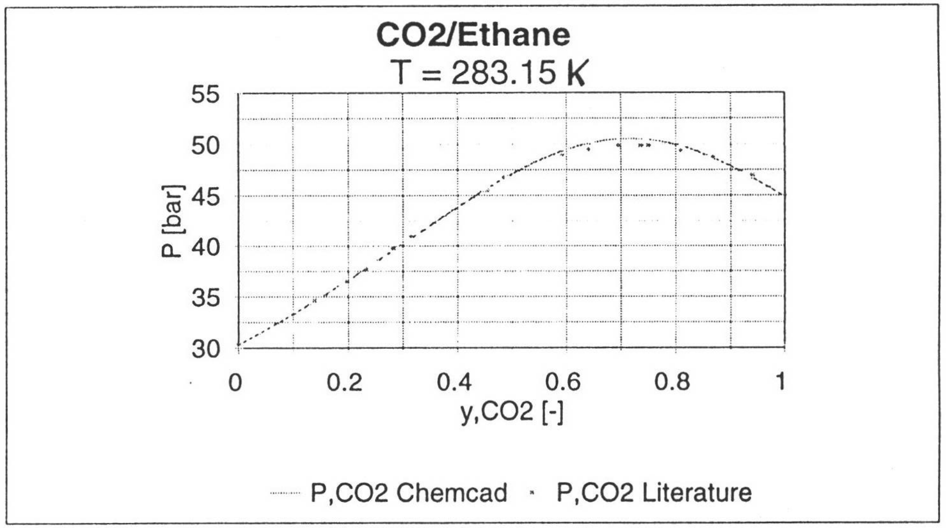


Figure 4: PY-diagram of CO2/ethane, T=283.15 K

Appendix 6-1: The four processes, first comparison, before optimising.

It assumed that the feed does not contain water. The C_3^+- seperation units are not yet included, as they are the same for all options.

Table 6.1.1: Comparing the four options

	Reflux heat	Cryomax	Turbo Expander (2)	Turbo Expander (1)
Recovery	98.9	92.3	98.4	99.3
C2/C3 (wt %)	1.4	1	1	1

Equipment:	Reflux heat	Cryomax	Turbo Expander (2)	Turbo Expander (1)
Reflux heat exch.	1	-	-	-
Turbo expander	-	1	2	1
Expander	1	-	2	1
Compressor	2	1	1	1
Heat exchanger	5 (6)*	5	8***	7***
Separator	1	1	3	2
Contactor	-	1	1	1
Deethaniser	1	1	1	1
Splitter/mixer	2 (1)**	2	2	2
Valve	3	-	1	-

* incl. reflux heat exchanger
** 1 used only for simulation
*** can be grouped in one coldbox

Deethaniser:	Reflux Heat Exchanger	Cryomax	Turbo Expander (2)	Turbo Expander (1)
Reboiler duty (MW)	43.8*	23.4	24.1	17.9
Trays	30	20	30	28
Feedstreams	2	2	4	3
Total feed (kg/s)	129	128.02	66.9	75.7

* A duty of 17.6 MW can be achieved, but recovery decreases to 88 %

Contactor:	Reflux Heat Exchanger	Cryomax	Turbo Expander (2)	Turbo Expander (1)
Trays	-	10	10	10
Total feed (kg/s)	-	176.36	133.19	129.13

Expanders:	Reflux Heat Exchanger	Cryomax	Turbo Expander (2)	Turbo Expander (1)
Pressure drop (bar)	110→50	110→35	110→70+70→30	110→ 30
Duty (MW)	-6.23	-14.11	-3.65-4.1= -7.75	-7.97

Appendix 6-1

Compressor:	**Reflux Heat Exchanger**	**Cryomax**	**Turbo Expander (2)**	**Turbo Expander (1)**
Pressure incr. (bar)	30→50	35→42.59	30→52.69	30→52.3
Duty (MW)	3.92	14.11	7.75	7.97
Sales gas Compressor:	**Reflux Heat Exchanger**	**Cryomax**	**Turbo Expander (2)**	**Turbo Expander (1)**
Pressure incr. (bar)	50→ 190	42.59→190	52.69→190	52.73→190
Duty (MW)	24.90	27.5	24.6	25.8

Heat exchangers:	**Reflux Heat Exchanger**	**Cryomax**	**Turbo Expander (2)**	**Turbo Expander (1)**
Amount	5	5	8	7
Heat	58.69	40.1	22.75	24.52
Cold	32.98	27.24	21.6	22.28
Integration	Not possible	Not complete	Possible	Possible

Appendix 6-2: Full comparison between the Cryomax and turbo-expander processes.

The feed does not contain any water.

Table 6.2.1: Process comparison at optimum conditions and heat integration

	Cryomax	**Turbo-Expander (2)**	**Turbo-expander (1)**
Recovery (%)	92.3	97.9	99.8
C2/C3 (wt%)	1.00	1.00	1.00
Equipment:			
Turbo-expander / compressor	1	2	1
Compressor	2	2	1
Heat exchanger	11	8*	8*
Separator	1	3	2
Contactor	1	1	1
Deethaniser	1	1	1
Depropaniser	1	1	1
Debutaniser	1	1	1
Splitter/mixer	2	2	2
Valve	3	4	3

*five can be grouped in one box

Deethaniser:	**Cryomax**	**Turbo-Expander (2)**	**Turbo-expander (1)**
Reboiler duty (MW)	9.58	16.7	17.9
Trays	20	35	28
Feedstreams	2	4	3
Total feed (kg/s)	52.01	64.19	75.7

Absorber:	**Cryomax**	**Turbo-Expander (2)**	**Turbo-expander (1)**
Trays	10	10	10
Total feed (kg/s)	158.06	124.14	129.13

Expanders:	**Cryomax**	**Turbo-Expander (2)**	**Turbo-expander (1)**
Pressure drop (bar)	110→17	110→72.5	110→30
Duty (MW)	-14.1	-3.4	-7.97
Pressure drop (bar)	-	72.5→30	-
Duty (MW)	-	-5.3	-
Total duty (MW)	-14.1	-8.7	-7.97

Compressor:	**Cryomax**	**Turbo-Expander (2)**	**Turbo-expander (1)**
Pressure incr. (bar)	17→46.2	30→52.4	30→49.2
Duty (MW)	14.1	3.4	7.97
Pressure incr. (bar)	-	30→52.4	-
Duty (MW)	-	5.3	-
Pressure incr. (bar)	46.2→190	52.4→190	49.2→190
Duty (MW)	28.5	24.1	25.9

Pump:*	**Cryomax**	**Turbo-Expander (2)**	**Turbo-expander (1)**
Pressure incr. (bar)	10→21	10.5→21	10.5→21
Duty (MW)	0.0283	0.0289	0.029

*only pumps for pressure increase

Depropaniser:	**Cryomax**	**Turbo-Expander (2)**
Condensor duty (MW)	-8.68	-10.1
Reboiler duty (MW)	7.88	6.7
Trays	42	45
Feedstreams	1	1
Total feed (kg/s)	23.52	24.20
Debutaniser:	**Cryomax**	**Turbo-Expander (2)**
Condensor duty (MW)	-5.88	-5.90
Reboiler duty (MW)	6.04	6.00
Trays	31	31
Feedstreams	1	1
Total feed (kg/s)	12.83	12.83

* depropaniser and debutaniser in the turbo-expander process with one expander, are very similar to the ones in the two expander process, the distillation train in the one expander process has not been optimised.

Appendix 6-3: Comparison of one and two expansion steps

process 1 expander		duty	cost
			$
Energy		MW	0.0068/MW
reboiler		17.9	3680812.8
dT air cooler	70.37		
compressor			0.13/kwh
duty		25.9	14141400
total		43.8	17822213

process 2 expanders		duty	cost
			$
Energy		MW	0.0068/MW
reboiler		16.71	3436110.7
dT air cooler	62.38		
compressor			0.13/kwh
duty		24.1	13158600
total		40.81	16594711

Prices based on chemcad III, with cost estimation

Costs of absorber,deethaniser and destillation train are the same for both processes, and are not considered here.

equipment	duty MW	kosten $
compressor1	25.9	8853000
compressor2	7.969	4304140
expander	-7.969	1046000
total equipment cost		14203140

equipment	duty MW	kosten $
compressor1	24.1	7562200
compressor2		2499040
compressor3		3313990
expander1		522460
expander2		749671
flash		225585
		14872946

propane prices from appendix =1=, 1% more propane recovery = 300000 $

RECOVERY	
11.4042	kg/s
99.8	%

RECOVERY		
11.1863	kg/s	propane
97.9	%	

1.9 %	more		
490000 $	more	5 years	5 years
2450000 $	more in 5 years	103314204 $	97846500 $

---

cost	difference	5467704	$ in 5 years
profits	difference	2450000	$ in 5 years

Conclusion: turbo expander process with one expander has more energy costs due to higher compression costs.

Higher propane recovery can not make up in five years, turbo expander process with two expanders is better.

Appendix 6-4: The Turbo-expander process

Blockscheme of the Turbo-expander process is given in appendix 2-2 and the Chemcad flow scheme is given below.

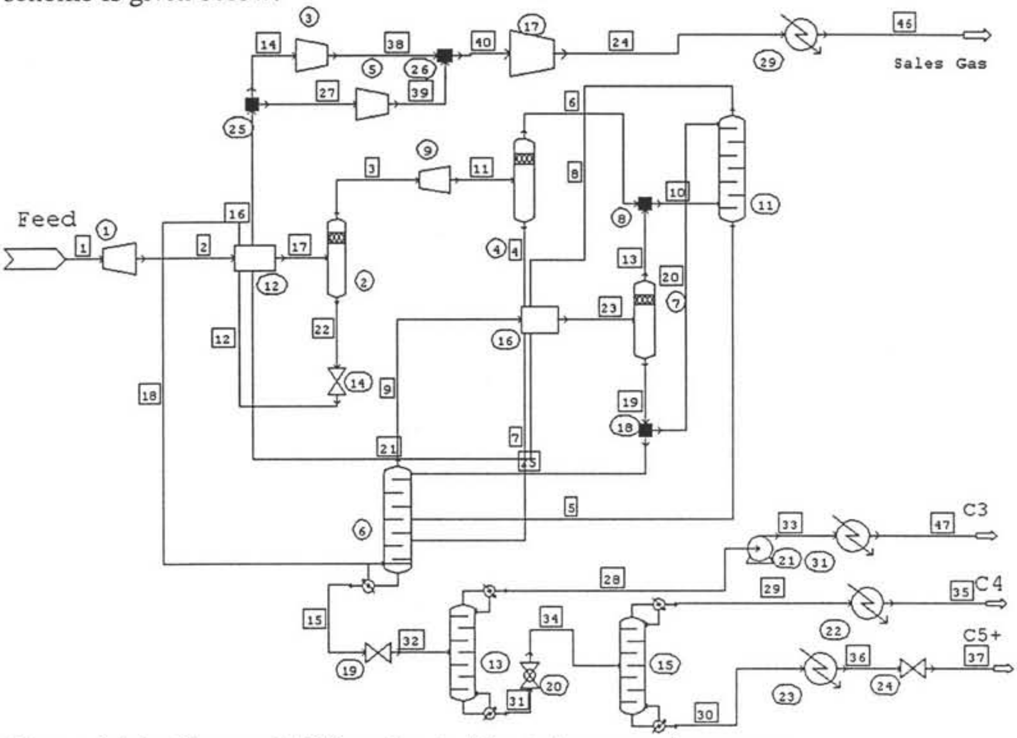


Figure 6.4.1: Chemcad III flowsheet of the turbo expander process.

Turbo-expander process with two expanders.

First the feed is dehydrated, because low temperatures are needed in the process and solids are unwanted. The feed, <1>, is expanded to 72.5 bara. This is done to achieve a lower temperature for separations, multiple phase flow and heat integration. With a valve the pressure from the liquids <22> is decreased to provide cooling. Then it is fed to the deethaniser (6).

The vapour <3> is expanded to process pressure of 30 bara, during which liquids are formed. In liquids <4> from the following separator are fed to the deethaniser after providing cooling for the reflux from the deethaniser. The vapours <6> are mixed with part of the vapour <13> from the deethaniser and fed to the absorber.

The absorber receives its liquid from the deethaniser to wash the remaining propane from the gas <10>. The liquids <5> are returned to the deethaniser. The vapour <8> is used for cooling feed and reflux, after which it is compressed in the compressor part of the turbo-expanders. To bring the sales gas to specification it is compressed and cooled.

The pressure from the liquids <15> from the deethaniser is let down to 16 bara and it is fed to the depropaniser (13). This pressure level is chosen so the depropaniser can be cooled using cooling water. The vapour <28> from the depropaniser is compressed to cooled and sold as C3-product. The pressure from the liquids <31> is let down to 9 bara and fed to the debutaniser (15).

The vapour <29> from the debutaniser is cooled and sold as C4-product. The liquid <30> is cooled and decreased in pressure to 8 bara to be sold as C5+-product.

Turbo-expander process with one expander.

The Turbo-expander process with one expander only differs by the absence of turbo-expander (5) (and compressor section), L/V separator (4) and pressure achieved in the first expander. Heat integration has not been considered for this process.

Appendix 6-5: The Cryomax process

Blockscheme of the Cryomax process is given in appendix 2.3 and the Chemcad flow scheme is given below.

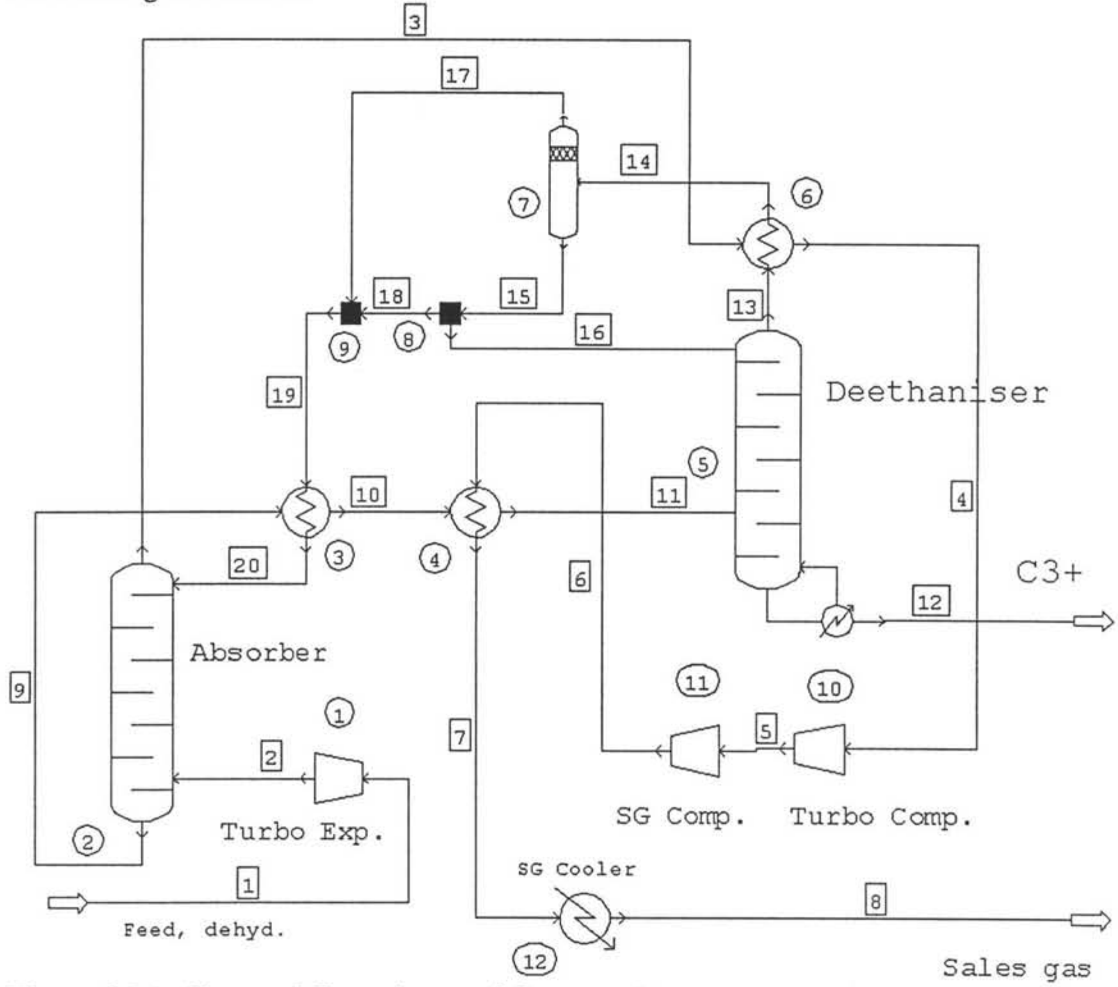


Figure 6.5.1. Chemcad flowscheme of Cryomax Process

The Cryomax process differs from the Turbo Expander process in the order in which the product passes through the absorber and deethaniser and by only using one expansion step.

First the feed is dehydrated, because low temperatures are needed in the process and solids are unwanted. Here the feed, <1>, is fed to the absorber, after being expanded to 17 bara. This is done to achieve a low temperature to use in the process and eliminate the need for external refrigeration. The vapour from the absorber, <3>, is used to condens the top product of the deethaniser in <13>. It is subsequently compressed in the turbo compressor, <5>, and brought to pipeline pressure, <6>. The temperature of the gas is above the maximum export temperature of 232 K. The gas is therefore cooled using the deethaniser feed, <10>, and cooling water. The liquid from the absorber, <9>, is heated to 270 K and fed to the deethaniser. This is done in combination with cooling of the absorber liquid feed, <20>, to 220 K and the sales gas, <6>.

The liquid from the deethaniser, <12>, is fed to the distillation train at the same conditions as described for the Turbo Expander process. The vapour from the deethaniser, <13>, is partially condensed by the sales gas, <3>, at 230 K and separated. The liquid from this separation step, <15>, is split and partially returned to the deethaniser as reflux, <16>. The rest is cooled further and added to the vapour stream of the separator, <17>. This mixture is completely condensed and cooled, <20>, at 220 K and fed to the absorber as liquid.

The following table gives the operating conditions in the Cryomax process.

Table 6.5.1: Operating conditions in the Cryomax process

Deethaniser:	Unit	Amount
Reboiler duty	MW	9.58
Trays	-	20
Feedstreams	-	2
Feedtrays	-	1 10
Reflux	kg/s	13.53
Feed	kg/s	38.48
Bottoms rec. C2	-	0.004

Absorber:		
Trays	-	10
Feedtrays	-	1 10
Liquid feed	kg/s	14.95
Vapour feed	kg/s	143.11

Turbo Exp./Comp.:		
Exp. Duty	MW	-14.1
Exp. Efficiency	%	85
Comp. Duty	MW	14.1
Comp. efficiency*	%	85

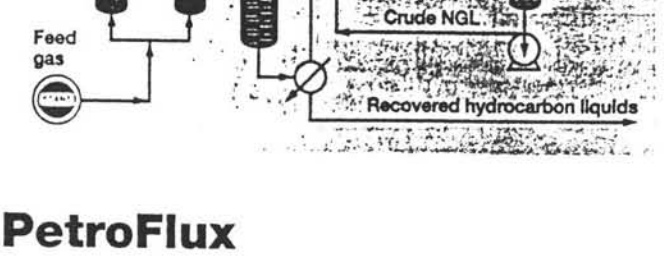
* not corrected to 75 % as used in the final design

Compressor:		
Duty	MW	28.5
Efficiency	%	85

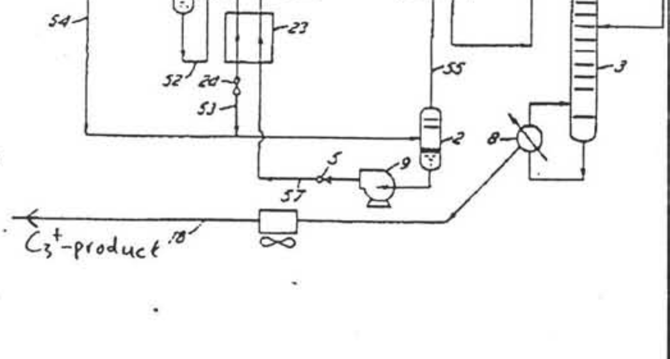
Appendix 6-6: The Reflux Heat Exchanger process

The block scheme of the Reflux Heat Exchanger process is given in appendix 2-4.

In the Reflux Heat Exchanger process the feed needs to be dehydrated as in the other processes, because low temperatures are used. The dehydrated feed is expanded to 50 bara and fed to the reflux heat exchanger.
This piece of equipment can be described as a separator vessel with a vertically placed heat exchanger in the top. In this heat exchanger the vapour is cooled in one or more steps. During this cooling the heavier part of the vapour condenses and flows back into the separator. By using several cooling stages at different temperatures the vapour can be stripped of heavier components. The vapour is the sales gas stream. The pressure of the liquid is let down and it is fed to a deethaniser. The bottoms of the deethaniser are fed to the distillation train as previously described. The vapour from the deethaniser is compressed and partially condensed. This mixture is separated and the vapour is added to the sales gas. The pressure of the liquid is let down again and fed to the deethaniser as reflux. All operations in the reflux loop for the deethaniser are performed to generate cold for the reflux heat exchanger. This is done in the patent on which the process is based. For the given feed this does not work as well and external refrigeration is needed.
Another variation of the process is to feed the vapours from the deethaniser back to the reflux heat exchanger after recompression. The block for compression/cooling/ separation/expansion can then be left out. This process scheme is very difficult to simulate without more knowledge of how a reflux heat exchanger can be modelled. It is also gives an extremely unstable simulation due to three totally interacting recycle loops and gives very large recycle streams in the reflux heat exchanger.
One note that has to be made is the trade off that would still have to be optimised between reboiler duty, C2/C3-ratio, compression costs and sizes of recycle streams. If the reflux heat exchanger can be modelled with greater accuracy it is a option that has to be considered in more detail.



PetroFlux



Appendix 7-1: Optimisation and sensitivity analysis

sensitivity/optimisation

expander 1 Pressure bar	 propane kg/s	reboiler duty MW	compressor duty MW	 recovery %	
72.5	11.1118	16.7	24.076	97.47193	ref.
80	11.3194	18.5	27.9	99.292982	
75	11.2467	17.1	25.2	98.655263	
65	10.4196	15.4	23.6	91.4	
60	9.7052	14.5	23.3	85.133333	* pinch

expander 2 Pressure bar	 propane kg/s	reboiler duty MW	compressor duty MW	 recovery %
30	11.118	17.1	24.076	97.526316
35	10.7188	18.0118	23.143	94.024561
25	11.2689	15.0856	27.191	98.85

splitter reflux deethaniser absorber	 top deeth	 propane kg/s	reboiler duty MW	compressor duty MW	 recovery %	
0.47	0.53	11.118	17.1	24.076	97.52632	optimised before using sg
0.42	0.58	11.1038	17.49	24.594	97.40175	compressor
0.52	0.48	11.1054	16.3974	23.865	97.41579	
0.54	0.46	11.1	16.3636	23.8547	97.36842	
0.4	0.6	11.1028	17.91	24.879	97.39298	
0.495	0.505	11.1078	16.4907	23.955	97.43684	

splitter compressors parallel 1	 2	pressure out bar K03	 K01	compressor duty MW
0.61	0.39	52.2459	52.2868	24.076
0.66	0.34	50.2311	56.3776	24.899
0.56	0.44	54.6977	49.2802	25.303
in series				24.192

tray configuration deethaniser <22> tray nr.	 <13> tray nr.	 <8> tray nr.	 total trays	 propane kg/s	reboiler duty MW	compressor duty MW	 recovery %
8	12	16	28	11.1118	16.7075	24.09876	97.47193
6	12	16	28	11.1039	16.7527	24.11455	97.40263
10	12	16	28	11.1091	16.7246	24.11919	97.44825
8	**10**	16	28	11.1062	16.7764	24.11663	97.42281
8	**14**	16	28	11.1082	16.708	24.10515	97.44035
8	12	**14**	28	11.1063	16.7193	24.08477	97.42368
8	12	**18**	28	11.1085	16.7103	24.10056	97.44298
8	12	16	**32**	11.1085	16.683	24.1089	97.44298
8	12	16	**30**	11.1072	16.6839	24.10943	97.43158
8	12	16	**26**	11.1069	16.8131	24.15503	97.42895
8	12	16	**24**	11.1	17.7096	24.77554	97.36842

temperature

stream,in nr. (base)	T,out K	propane kg/s	duty MW	duty MW	recovery %	
base		11.1118	16.7075	24.09876	97.47193	
<3> (245)	250	10.3598	15.6313	23.0125	90.87544	
	240	11.3231	18.5067	25.646	99.32544	
<7> (250)	255	11.1068	16.1782	23.724	97.42807	*dT of 2°K is very small
	245	11.1087	17.23	24.45	97.44474	
<12> (235)	240	11.1069	16.2845	23.791	97.42895	*pinch, not possible
	230	11.1118	17.1509	24.4	97.47193	
<23> (220)	225	10.4655	15.5566	23.283	91.80263	
	215	10.4655	23.7887	23.283	91.80263	* pinch
	218	11.3187	18.1666	25.1	99.28684	

In this sensitivity analysis, only theoretical trays were considered

absorber

trays	propane kg/s	duty MW	duty MW	recovery %	
10	11.109	16.9	24.125	97.447368	
15	11.1092	16.6986	24.125	97.449123	small difference, 5 trays more expensive
5	11.0767	16.9044	24.193	97.164035	
20	11.1088	16.967	24.082	97.445614	

depropaniser

variable pressure bar	reboiler duty MW	condensor duty MW	cond temp K
16	8.0863	9.9422	318.7
11	6.789	9.978	302.9
21	9.528	10.23	331.5

feed tray

20	8.0863	9.9422	318.7
18	8.148	10.01	318.8
15	8.421	10.28	318.8
29	9.726	11.58	318.8

trays

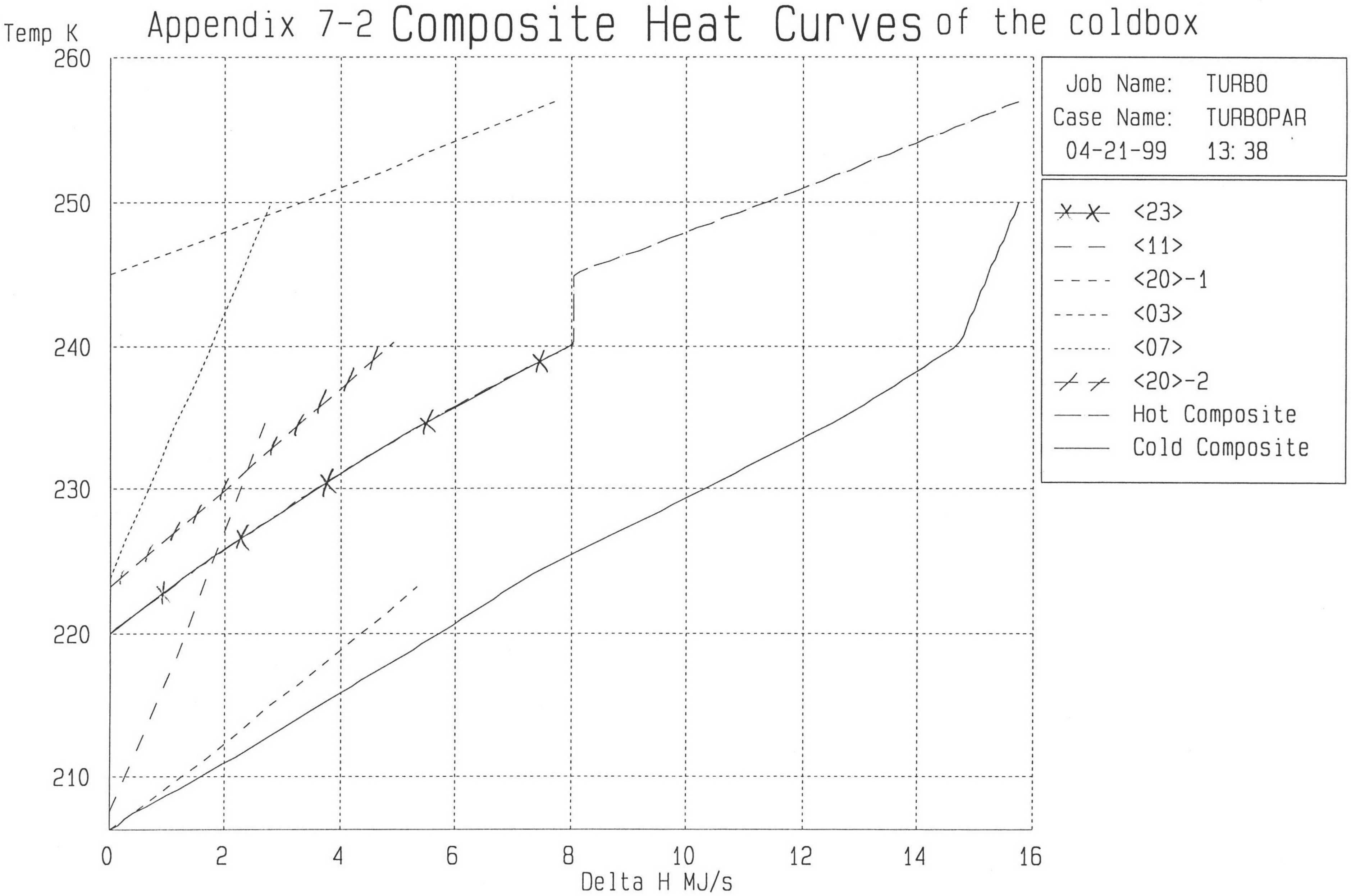
45	8.148	10.01	318.8
50	8.111	9.971	318.8
40	8.371	10.23	318.8

debutaniser

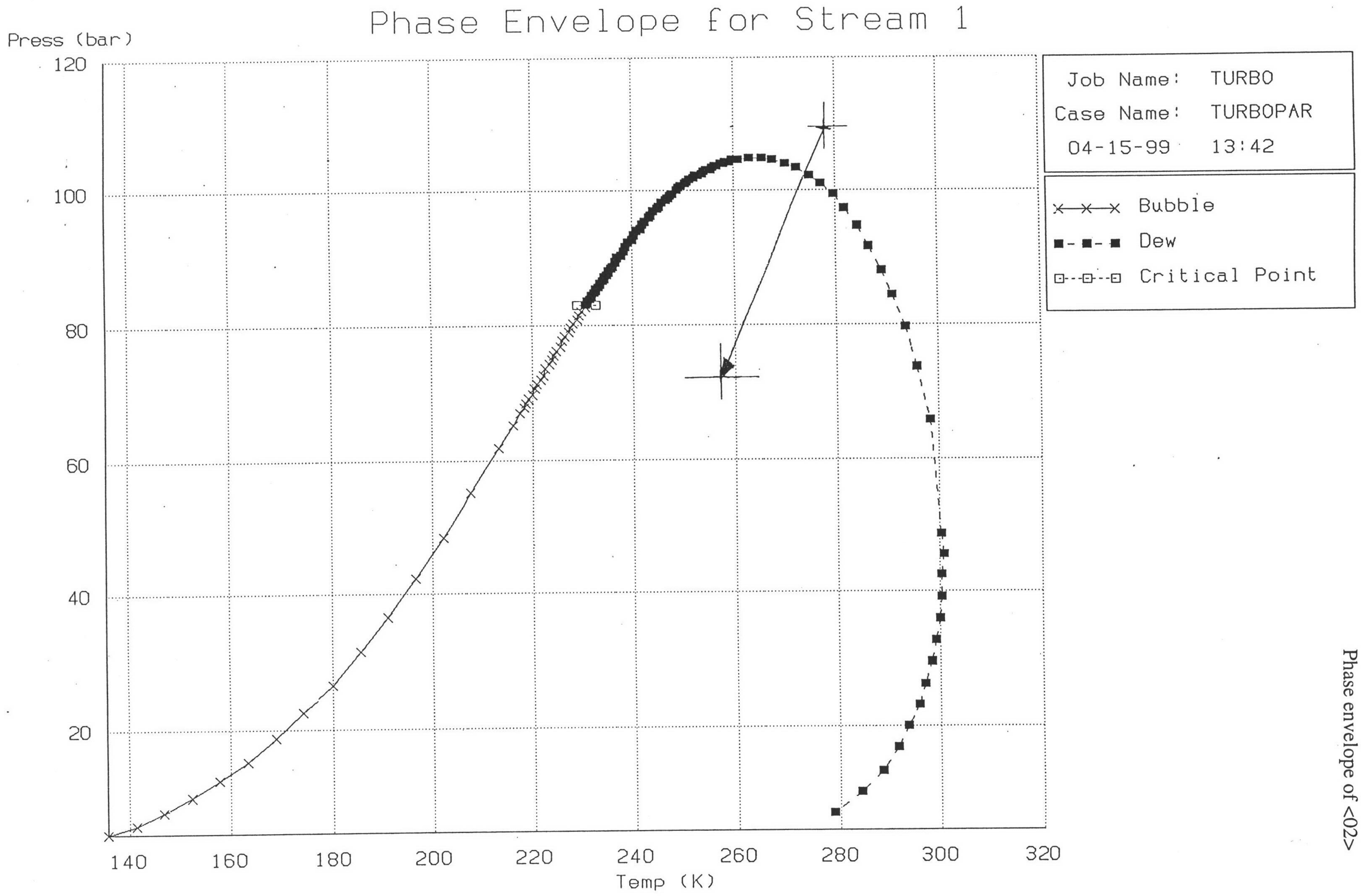
feed tray	reboiler duty MW	condensor duty MW	cond temp K
15	5.3728	6.1763	343
13	5.413	6.216	343
14	5.3762	6.1795	343
16	5.411	6.213	343

trays

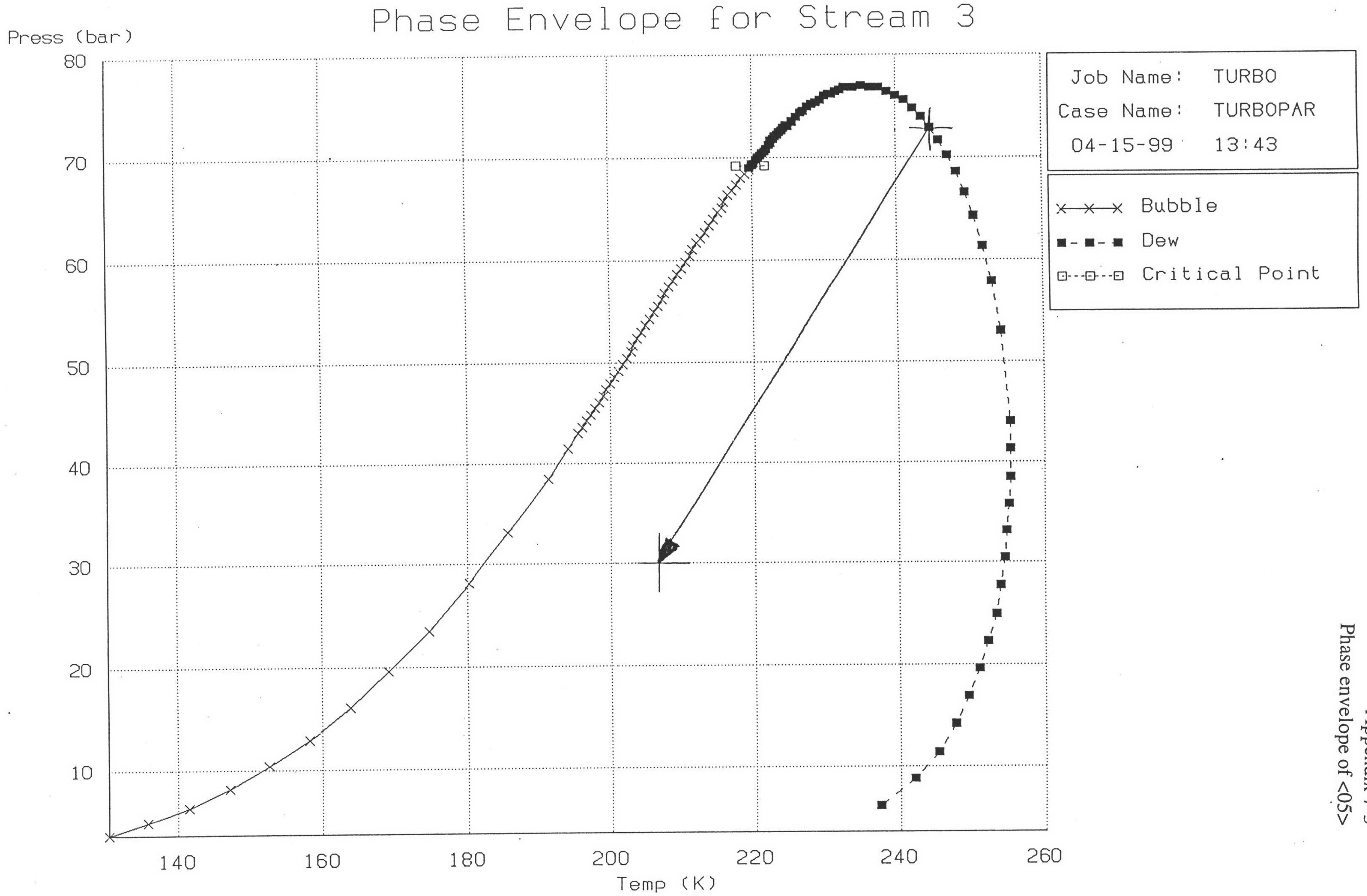
31	5.3728	6.1763	343
36	5.148	5.952	343
26	6.528	7.331	343



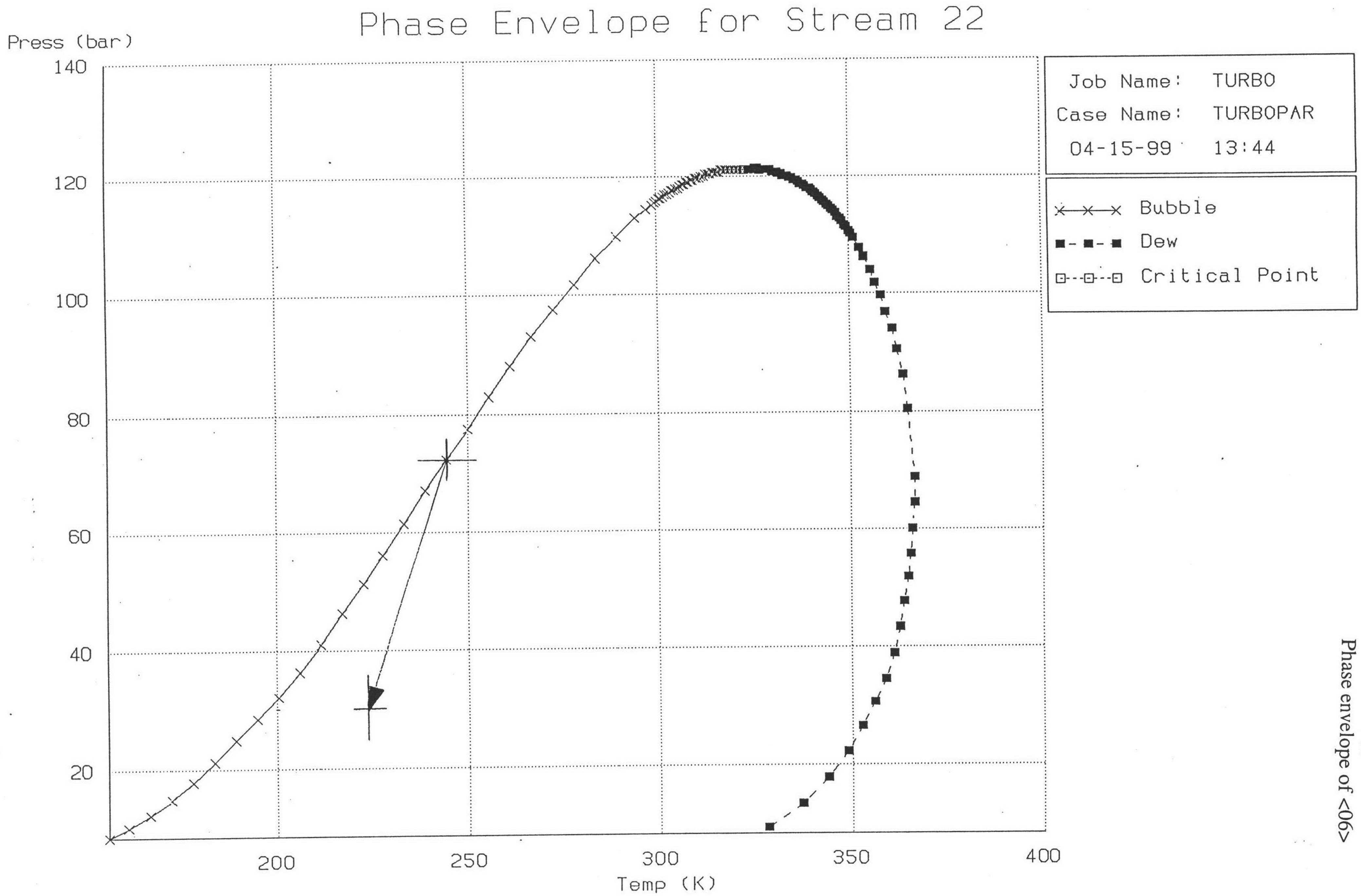
Appendix 7-2
Composite heat curves of the coldbox



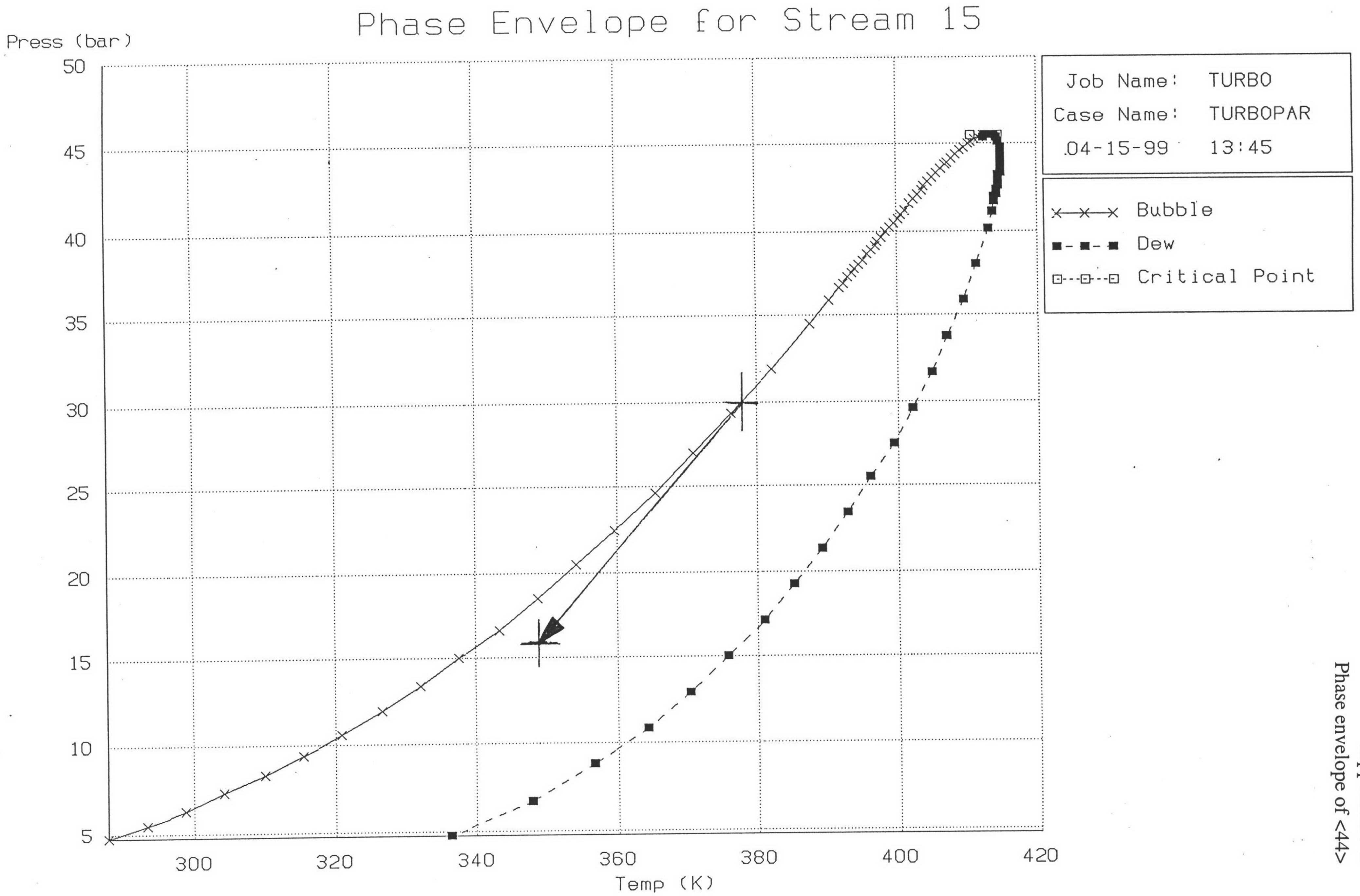
Appendix 7-3
Phase envelope of <02>



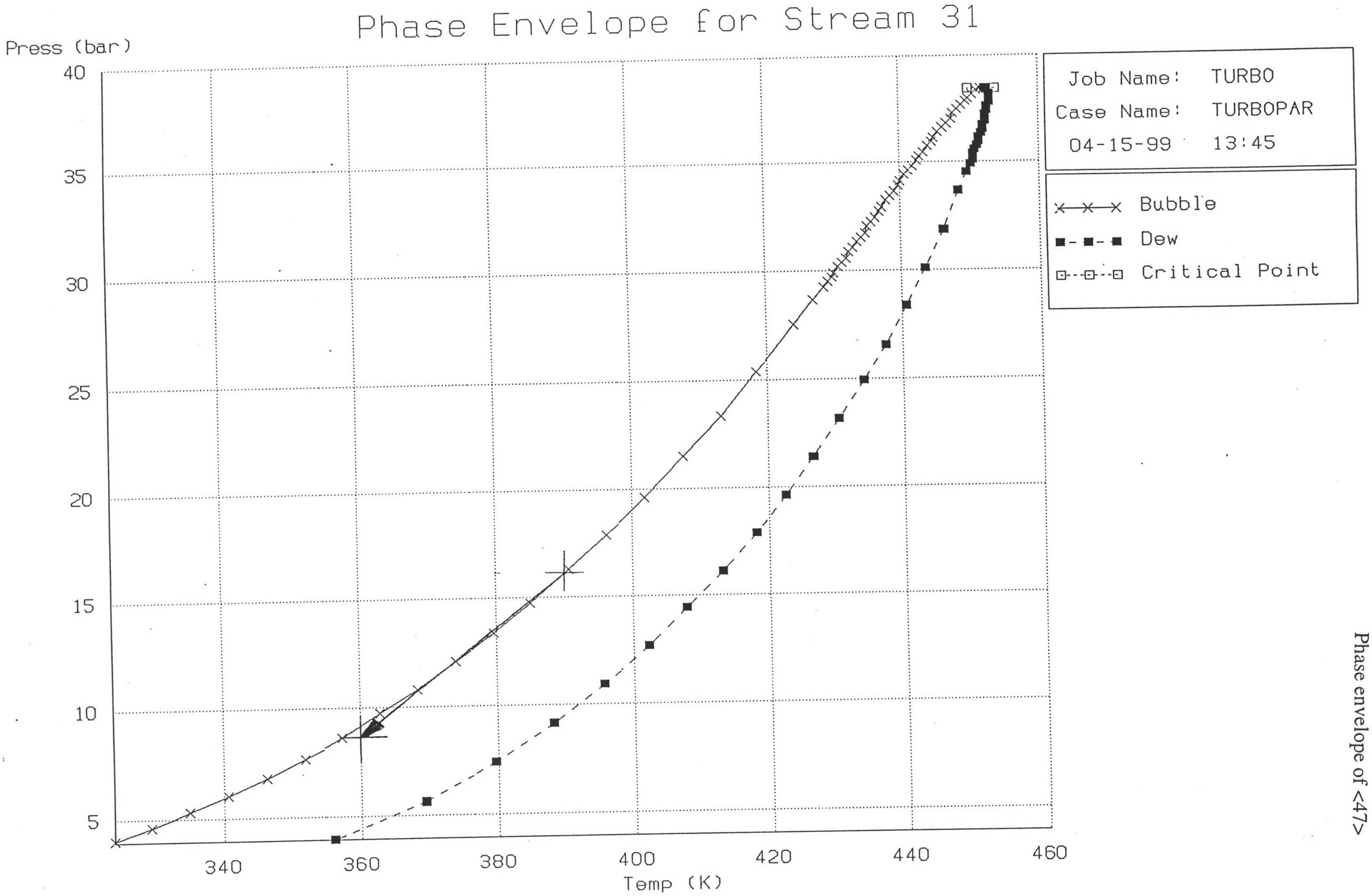
Appendix 7-3
Phase envelope of <05>



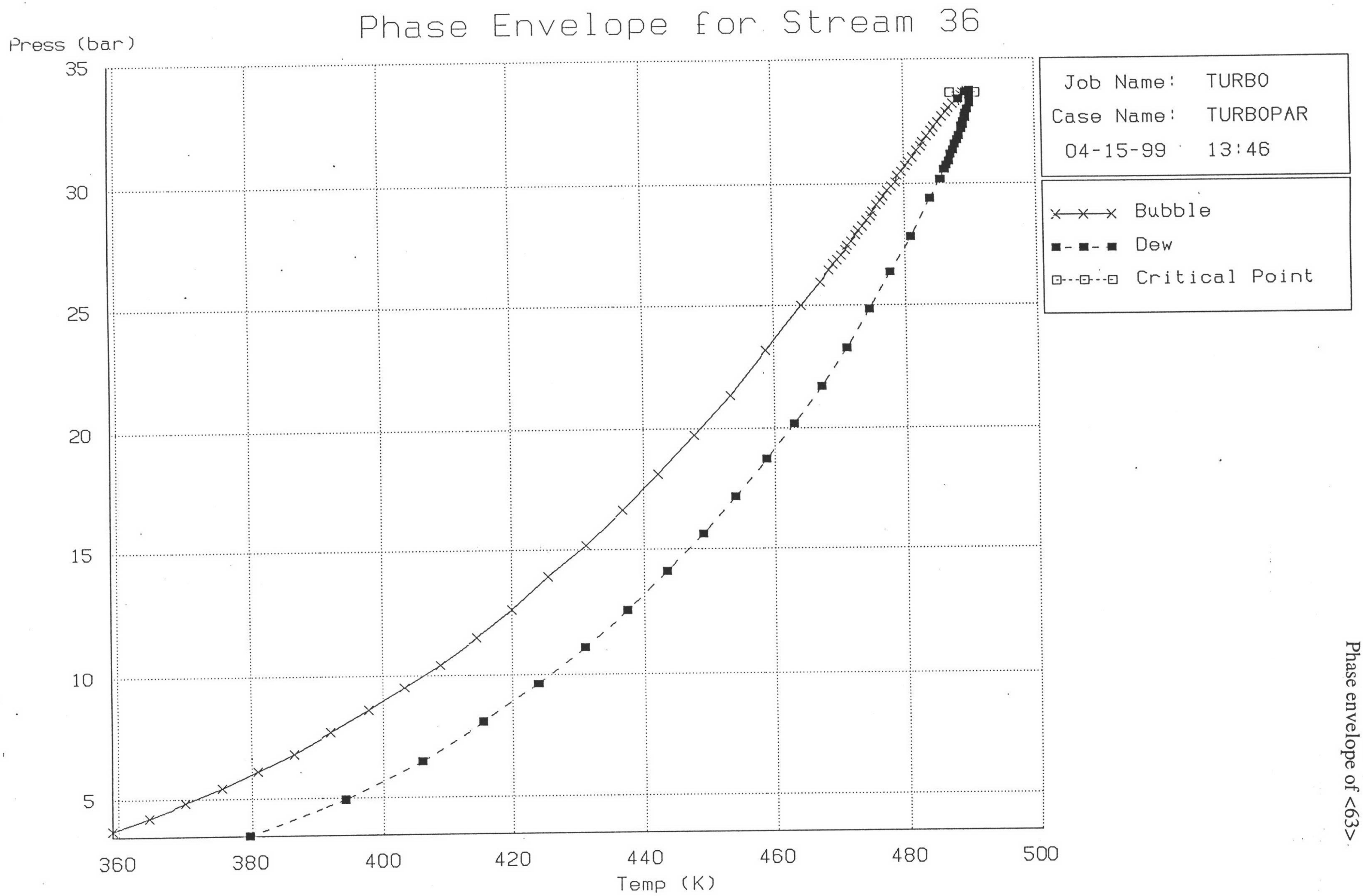
Appendix 7-3
Phase envelope of <06>



Appendix 7-3
Phase envelope of <44>



Appendix 7-3
Phase envelope of <47>



Appendix 7-3
Phase envelope of <63>

Appendix 7-4: Chemcad output for the turboexpander process

Below, the output generated by Chemcad III is given for the turbo-expander process. Where necessary comments have been added. The stream summaries are left out to reduce the length of the output. For stream summaries the reader is referred to appendix 5-2, stream summaries.

```
ChemCAD 3.30-386  License: Univ. of Technology, Delft, NL
Job Code: TURBO     Case Code: TURBOPAR   Date: 04-14-99  Time: 11:46

Date:  04-14-99   Time:  11:46

CALCULATION
Calculation mode : Sequential
Flash algorithm  : Normal

Equipment Calculation Sequence
   1  12   2   9  14   4   6   8  11  16   7  18  19  25   3   5  13  20
  21  26  31  15  17  22  23  24  29

Equipment Recycle Sequence
  12   2   9  14   4   6   8  11  16   7  18

Recycle Cut Streams
  21  12   5  25   7  13  20

Recycle Convergence Method:  Direct Substitution

    Max. loop iterations     40

Recycle Convergence Tolerance

    Flow rate        1.000E-003
    Temperature      1.000E-003
    Pressure         1.000E-003
    Enthalpy         1.000E-003
    Vapor frac.      1.000E-003

Recycle calculation has converged.
```

The calculation sequence is generated by Chemcad. The cut streams were estimated manually. All convergences were chosen for sufficient accuracy (0.1 %) at relatively short calculation times.

COMPONENTS

```
        ID #         Name

  1       46       Nitrogen
  2        2       Methane
  3        3       Ethane
  4       49       Carbon Dioxide
  5       50       Hydrogen Sulfide
  6        4       Propane
  7        5       I-Butane
  8        6       N-Butane
  9        7       I-Pentane
 10        8       N-Pentane
 11       10       N-Hexane
 12       11       N-Heptane
 13       62       Water
 14       12       N-Octane
```

The components are listed in order of increasing boiling points. Water was kept at zero throughout the process.

THERMODYNAMICS

```
 K-value model      :  Peng-Robinson
                       Water/Hydrocarbon immiscible
                       Boston-Mathias Alpha function.
 Enthalpy model     :  Peng-Robinson
 Liquid density     :  Library
```

The choice of thermodynamic model is given in chapter 4. According to the Chemcad manual the Peng-Robinson enthalpy model is used if the Peng-Robinson model is used for the K-value. The immiscibility of water/hydrocarbon option is chosen, because the Peng-Robinson model is used. As many of the components are gases above their critical temperature the Boston-Mathias Alpha function is used for better results.

Overall Mass Balance

```
Overall Mass Balance        kmol/s                     kg/s
                        Input        Output       Input        Output
Nitrogen                0.028        0.028        0.773        0.773
Methane                 5.617        5.617       90.111       90.112
Ethane                  0.568        0.568       17.069       17.070
Carbon Dioxide          0.232        0.232       10.215       10.215
Hydrogen Sulfide        0.000        0.000        0.001        0.001
Propane                 0.257        0.257       11.343       11.343
I-Butane                0.043        0.043        2.525        2.525
N-Butane                0.078        0.078        4.529        4.529
I-Pentane               0.022        0.022        1.617        1.617
N-Pentane               0.023        0.023        1.642        1.642
N-Hexane                0.017        0.017        1.455        1.456
N-Heptane               0.008        0.008        0.829        0.829
Water                   0.000        0.000        0.000        0.000
N-Octane                0.002        0.002        0.236        0.236

Total                   6.895        6.895      142.344      142.346
```

The overall mass balance shows that the in- and outflows are equal, within the chosen tolerance.

EQUIPMENT SUMMARIES

The equipment summary shows all options chosen for the various pieces of equipment. For the meaning of the numbers used in e.g. modes of operation or equipment types the reader is referred to the Chemcad manual.

```
                         Expander Summary

Equip. No.                      1            9
       Name
Type of Expander:               1            1
Pressure out  bar         72.5000      30.0000
Efficiency                 0.8500       0.8500
Actual power  kW       -3363.1091   -5252.4199
Cp/Cv                      2.4044       2.7428
Ideal Cp/Cv                1.1152       1.1122
Calc Pout  bar            72.5000      30.0000
Theoretical power  kW  -3956.5986   -6179.3174
```

Efficiencies of 85% are given by a manufacturer in appendix XXX.

```
                         Flash Summary

Equip. No.                      2            4            7
       Name
Type                            0            0            0

 K values:
 Nitrogen                   3.821        9.138        7.775
 Methane                    1.621        2.061        2.206
 Ethane                     0.463        0.194        0.300
 Carbon Dioxide             0.787        0.503        0.699
 Hydrogen Sulfid            0.422        0.187        0.303
 Propane                    0.195        0.036        0.069
 I-Butane                   0.104        0.011        0.023
 N-Butane                   0.083    6.926E-003       0.017
 I-Pentane                  0.045    1.898E-003   4.942E-003
 N-Pentane                  0.040    1.610E-003   4.100E-003
 N-Hexane                   0.016    2.693E-004   8.601E-004
 N-Heptane             7.394E-003    6.032E-005   2.540E-004
 Water                      0.414        0.072        0.252
 N-Octane              3.528E-003    1.475E-005   8.580E-005
```

```
                         Compressor Summary

Equip. No.                      3            5           17
       Name
Mode of Operation:              1            1            0
Type of Compressor:             1            1            1
Pressure out  bar         52.2660      52.2251     190.0000
Efficiency                 0.7500       0.7500       0.8500
Actual power  kW        3363.1689    5252.3643   24087.7656
Cp/Cv                      1.5424       1.5424       1.5297
Ideal Cp/Cv                1.2363       1.2363       1.2188
Calc Pout  bar            52.2660      52.2251     190.0000
Theoretical power  kW   2522.6138    3939.6379   20474.6016
```

Efficiencies of 75% are given by a manufacturer in appendix XXX. For equipment numbers 3 and 5 (turbo compressors) the actual power taken from the expanders was used as input.

```
                      Towr Rigorous Distillation Summary

Equip. No.                 6            11            13            15
       Name
No. of stages             28            10            45            31
1st feed stage             1             1            20            15
2nd feed stage             8            10             0             0
3rd feed stage            12             0             0             0
4th feed stage            16             0             0             0
Select condenser mode:     0             0             7             7
Condenser spec.                                   0.9990        0.9880
Cond. comp i               0             0             6             8
Select reboiler mode:      9             0             7             7
Reboiler spec.        0.0147                      0.9700        0.9600
Rebl. comp i               3             0             7             9
Rebl. comp j               6             0             0             0
Est. dist. rate       1.0000        1.0000        0.2425        0.1200
 (kmol/s)
Est. reflux rate                                  0.5000        0.2000
 (kmol/s)
Est. stage 1 T  K   230.0000      230.0000      319.0000      338.0000
Est. bottom T  K    377.0000      230.0000      390.0000      395.0000
Initial flag               1             1             1             1
Calc cond duty  MJ/s                             -9.9455       -6.1735
Calc rebr duty  MJ/s 16.7094                      8.0884        5.3701
Reflux ratio                                      1.9709        1.9342
Reflux mole  kmol/s                               0.5066        0.2331
Reflux mass  kg/s                                22.2741       13.5728
```

The feedstage locations were optimised for recovery, reboiler and compressor duty. For the deethaniser (6) the reboiler specification implies a C2/C3-ration of 1 %. Condenser and reboiler specifications for the depropaniser and debutaniser (13 and 15) are recoveries.

```
                      Mixer Summary
Equip. No.                   8            26
       Name

                      LNGH Summary

Equip. No.                   12           16
       Name
1st strm mode                 1            1
1st strm spec          245.0000     220.0000
Overall Q  MJ/s          7.7231       8.0409
Q1  MJ/s                -7.7231      -8.0409
2nd strm mode                 1            1
2nd strm spec          250.0000     235.0000
Q2  MJ/s                 2.8023       2.7190
Q3  MJ/s                 4.9208       5.3219
```

Though there were two cold boxes used in the simulation this was only done to prevent Chemcad from giving a pinch warning. In all cold boxes one stream could not be specified.

```
                      Valve Summary

Equip. No.                14            19            20            24
       Name
Pressure out  bar    30.0000       16.0000        9.0000        8.0000
```

```
                    Divider Summary

Equip. No.                    18            25
       Name
Output stream #1            0.4700        0.6100
Output stream #2            0.5300        0.3900
```

```
                     Pump Summary

Equip. No.                    21
       Name
Output pressure  bar        21.0000
Efficiency                   0.8500
Calculated power  kW        14.7196
Calculated Pout  bar        21.0000
Head  m                    112.8086
Vol. flow rate  m3/h        90.0196
```

For the actual calculation an efficiency of 76 % is used.

```
                     Heat Exchanger Summary

Equip. No.               22            23            29            31
       Name
T Out Str 1  K       310.1500      310.1500      322.1500      310.1500
Calc Ht Duty  MJ/s    -0.6187       -1.3395      -25.7865       -0.3258
LMTD (End points)  K 14.3386        0.0006
LMTD Corr Factor       1.0000        1.0000
Str1 Pout  bar         9.0000        9.0000      190.0000       21.0000
Str2 Pout  bar        21.0000       21.0000
```

DISTILLATION PROFILE

```
Unit type : TOWR    Unit name:                 Eqp #  6

                       *   Net Flows   *
      Temp    Pres     Liquid     Vapor      Feeds      Product     Duties
Stg     K     bar      kmol/s     kmol/s     kmol/s     kmol/s      MJ/s
  1  240.2    30.00       0.41                  0.41       1.73
  2  243.7    30.00       0.41       1.73
  3  244.6    30.00       0.40       1.73
  4  245.4    30.00       0.39       1.72
  5  246.3    30.00       0.37       1.71
  6  247.4    30.00       0.36       1.69
  7  248.8    30.00       0.34       1.68
  8  250.3    30.00       0.52       1.66       0.19
  9  254.5    30.00       0.51       1.65
 10  256.5    30.00       0.49       1.64
 11  258.5    30.00       0.46       1.62
 12  261.6    30.00       0.93       1.59       0.67
 13  270.9    30.00       0.96       1.39
 14  275.2    30.00       0.95       1.42
 15  278.7    30.00       0.91       1.41
 16  284.9    30.00       1.41       1.37       0.91
 17  309.9    30.00       1.64       0.96
 18  322.5    30.00       1.76       1.19
 19  331.2    30.00       1.83       1.31
 20  338.2    30.00       1.89       1.38
 21  344.1    30.00       1.95       1.44
 22  348.8    30.00       2.01       1.51
 23  352.6    30.00       2.06       1.56
 24  355.8    30.00       2.09       1.61
 25  358.8    30.00       2.10       1.64
 26  362.3    30.00       2.07       1.65
 27  367.6    30.00       1.97       1.62
 28  377.9    30.00                  1.52        0.45       16.71

Unit type : TOWR    Unit name:                 Eqp #  11

                       *   Net Flows   *
      Temp    Pres     Liquid     Vapor      Feeds      Product     Duties
Stg     K     bar      kmol/s     kmol/s     kmol/s     kmol/s      MJ/s
  1  206.1    30.00       0.37                  0.37       6.45
  2  205.6    30.00       0.34       6.45
  3  206.3    30.00       0.30       6.42
  4  207.0    30.00       0.27       6.38
  5  207.6    30.00       0.25       6.35
  6  208.0    30.00       0.24       6.33
  7  208.2    30.00       0.24       6.32
  8  208.3    30.00       0.23       6.31
  9  208.5    30.00       0.22       6.31
 10  208.7    30.00                  6.30       6.27       0.19
```

```
Unit type : TOWR    Unit name:                 Eqp #  13

                              *   Net Flows    *
       Temp .     Pres     Liquid       Vapor     Feeds    Product      Duties
Stg     K         bar      kmol/s       kmol/s    kmol/s   kmol/s       MJ/s
  1   318.7      16.00        0.51                          0.26        -9.946
  2   319.8      16.00        0.51         0.76
  3   320.2      16.00        0.51         0.76
  4   320.6      16.00        0.50         0.76
  5   321.1      16.00        0.50         0.76
  6   321.7      16.00        0.50         0.76
  7   322.4      16.00        0.49         0.76
  8   323.3      16.00        0.49         0.75
  9   324.4      16.00        0.49         0.75
 10   325.6      16.00        0.48         0.74
 11   327.0      16.00        0.48         0.74
 12   328.4      16.00        0.47         0.73
 13   329.8      16.00        0.47         0.73
 14   331.2      16.00        0.46         0.72
 15   332.5      16.00        0.46         0.72
 16   333.9      16.00        0.45         0.71
 17   335.2      16.00        0.45         0.71
 18   336.8      16.00        0.44         0.70
 19   339.0      16.00        0.41         0.69
 20   343.6      16.00        0.67         0.67     0.45
 21   344.0      16.00        0.68         0.48
 22   344.4      16.00        0.67         0.48
 23   344.9      16.00        0.68         0.48
 24   345.6      16.00        0.67         0.48
 25   346.5      16.00        0.67         0.48
 26   347.9      16.00        0.67         0.48
 27   349.6      16.00        0.67         0.48
 28   351.8      16.00        0.67         0.48
 29   354.4      16.00        0.67         0.48
 30   357.3      16.00        0.68         0.48
 31   360.4      16.00        0.68         0.48
 32   363.3      16.00        0.69         0.49
 33   366.0      16.00        0.69         0.49
 34   368.3      16.00        0.70         0.50
 35   370.2      16.00        0.70         0.51
 36   371.8      16.00        0.71         0.51
 37   373.0      16.00        0.71         0.51
 38   373.9      16.00        0.71         0.52
 39   374.7      16.00        0.71         0.52
 40   375.5      16.00        0.71         0.52
 41   376.2      16.00        0.71         0.52
 42   377.2      16.00        0.71         0.52
 43   378.7      16.00        0.70         0.52
 44   381.8      16.00        0.68         0.51
 45   389.6      16.00                     0.49              0.19        8.088

Mole Reflux ratio         1.971
```

```
Unit type : TOWR    Unit name:                    Eqp #  15

                              *   Net Flows   *
       Temp       Pres       Liquid      Vapor      Feeds    Product      Duties
Stg     K         bar        kmol/s      kmol/s     kmol/s   kmol/s        MJ/s
  1   343.0       9.00        0.23                           0.12        -6.174
  2   344.1       9.00        0.23        0.35
  3   345.0       9.00        0.23        0.35
  4   345.8       9.00        0.23        0.35
  5   346.7       9.00        0.23        0.35
  6   347.6       9.00        0.22        0.35
  7   348.6       9.00        0.22        0.35
  8   349.7       9.00        0.22        0.34
  9   350.8       9.00        0.22        0.34
 10   351.8       9.00        0.22        0.34
 11   352.9       9.00        0.22        0.34
 12   353.8       9.00        0.22        0.34
 13   354.9       9.00        0.21        0.34
 14   356.5       9.00        0.20        0.33
 15   360.2       9.00        0.34        0.32       0.19
 16   361.4       9.00        0.34        0.26
 17   362.8       9.00        0.33        0.26
 18   364.5       9.00        0.33        0.26
 19   366.5       9.00        0.33        0.26
 20   368.7       9.00        0.33        0.26
 21   371.1       9.00        0.33        0.26
 22   373.6       9.00        0.33        0.26
 23   375.9       9.00        0.33        0.26
 24   378.1       9.00        0.34        0.26
 25   380.0       9.00        0.34        0.26
 26   381.7       9.00        0.34        0.26
 27   383.2       9.00        0.34        0.27
 28   384.7       9.00        0.34        0.27
 29   386.8       9.00        0.33        0.26
 30   391.1       9.00        0.32        0.26
 31   400.8       9.00                    0.24                   0.07        5.37

Mole Reflux ratio          1.934
```

Appendix 8-1: Calculation of mass amounts in feedstream

Using the Peng-Robinson equation with its accompanying equations for the constants, the molecular volume can be obtained for the feedstream. With this molecular volume and molecular weights the feed stream <01> can be characterised in terms of the massflows of the various components. The data needed for the Peng-Robinson equation can be found in [59], and is shown in the table below. It is assumed that the overall molecular volume is a sum of the product of the individual molecular volumes and their molecular fractions.

Table 8.1.1: Calculation of the molecular weight of the feedstream.

Component	Tc	b	Omega	f_omega	Tr	a
	[K]	[m3/mol]	[-]	[-]	[-]	[J/mol]
N2	126.2	2.41E-05	0.039	0.43	2.28	0.09
C1	190.4	2.68E-05	0.011	0.39	1.51	0.21
C2	305.4	4.05E-05	0.099	0.52	0.94	0.62
CO2	304.1	2.67E-05	0.239	0.73	0.95	0.41
H2S	373.2	2.70E-05	0.097	0.52	0.77	0.56
C3	369.8	5.63E-05	0.153	0.60	0.78	1.17
iC4	408.2	7.23E-05	0.183	0.65	0.71	1.76
nC4	425.2	7.24E-05	0.199	0.67	0.68	1.88
iC5	469.7	8.96E-05	0.227	0.71	0.61	2.74
nC5	460.4	8.84E-05	0.251	0.74	0.63	2.65
nC6	507.5	1.09E-04	0.299	0.81	0.57	3.90
nC7	540.3	1.28E-04	0.349	0.88	0.53	5.16
H2O	647.3	1.89E-05	0.344	0.87	0.45	1.00
nC8	568.8	1.48E-04	0.398	0.95	0.51	6.65

	RT/(V-b)	a/(V^2+2bV-b^2)	V	V*molfraction
	[Pa]	[Pa]	[m3/mol]	[m3/mol]
N2	100156	156	2.39E-02	9.58E-05
C1	100360	360	2.39E-02	1.95E-02
C2	101101	1101	2.37E-02	1.95E-03
CO2	100724	724	2.38E-02	8.01E-04
H2S	100985	985	2.38E-02	5.94E-08
C3	102098	2098	2.35E-02	8.77E-04
iC4	103222	3222	2.33E-02	1.47E-04
nC4	103466	3466	2.32E-02	2.62E-04
iC5	105201	5201	2.29E-02	7.43E-05
nC5	105024	5024	2.29E-02	7.56E-05
nC6	107723	7723	2.23E-02	5.48E-05
nC7	110767	10767	2.18E-02	2.61E-05
H2O	101795	1795	2.36E-02	8.24E-06
nC8	114868	14868	2.10E-02	6.30E-06
Total				2.38E-02

This means a feedstream of 5 Gsm^3/a equals $2.1*10^{11}$ mol/a and $6.9*10^3$ mol/s. Given the molecular weights and the molecular fractions of the components the massflows can be determined, as given in the table below.

Table 8.1.2: Calculation of feedstream

	Feed	Feed	Feed
	[mol%]	[mol/s]	[kg/s]
N2	0.400	27.74	0.78
CO2	3.365	233.35	10.27
C1	81.430	5646.86	90.59
C2	8.230	570.72	17.16
C3	3.730	258.66	11.41
iC4	0.630	43.69	2.54
nC4	1.130	78.36	4.55
iC5	0.325	22.54	1.63
nC5	0.330	22.88	1.65
nC6	0.245	16.99	1.46
nC7	0.120	8.32	0.83
nC8	0.030	2.08	0.24
H2S	2.50E-04	0.02	0.00
H2O	0.035	2.43	0.04
Total	100	6934.64	143.16

Appendix 8-2: Molecular sieve design

For the design of the molecular sieve beds the following equations are used (Trent [3]):

$$V = \frac{61.5}{\sqrt{rho\_gas}}$$

$$LUB = 1.25 + \frac{0.0125 * V * M}{18}$$

$$MTZ = 2 * LUB$$

$$\frac{dP}{L} = A * \mu * V + B * rho\_gas * V^2$$

in which: V = maximum superficial velocity, downflow ft/min
rho_gas = gas density lb/ft^3
LUB = length of unused bed ft
MTZ = length of the mass transfer zone ft
dP/L = pressure drop/bed depth psi/ft
A,B = dimensional constants ……
μ = viscosity of fluid cP

The following table gives all the results for the design of the bed. All variables in bold have been chosen, not calculated. No correction is made for amount of decimals. Under the remarks the equations can be found.

Table 8.2.1: Design of molecular sieve bed

Molecular sieve bed design					Remarks
beds total	2	3	4		
beds in parallel	1	2	3		
Molsieve 3A					
Adsorption capacity	**0.113**	**0.113**	**0.113**	kg/kg	from fig. 2, 0.7 * 0.19-0.02 for residual water
Porosity	**0.3**	**0.3**	**0.3**	-	fraction of void area
Bed density	**688.79**	**688.79**	**688.79**	kg/m^3	43 lb/ft3, from page 11-1992
Cost	**3.18**	**3.18**	**3.18**	$/kg	10300*4 / 28600/0.45359, fig 13-1995, for 3A, 4A is 2.53$/kg
Water in	**0.0437**	**0.02185**	**0.0145667**	kg/s	
Gas flow	**3249.54**	**1624.768**	**1083.1787**	m^3/h	
Gas flow	0.903	0.451	0.301	m^3/s	
density gas	**157.696**	**157.696**	**157.696**	kg/m^3	
density gas	9.845	9.845	9.845	lbs/ft^3	
viscosity G	**1.78E-05**	**1.78E-05**	**1.78E-05**	Pas	
viscosity G	1.78E-02	1.78E-02	1.78E-02	cP	
mol wt gas	**20.644**	**20.644**	**20.644**	g/mol	
V,max	19.601	19.601	19.601	ft/min	downflow, maximum, page 11 (1992)
V,max	0.100	0.100	0.100	m/s	

A,min	9.065	4.533	3.022	m^2	
D,min	3.397	2.402	1.961	m	
V,chosen	**0.090**	**0.090**	**0.090**	m/s	90% of V,max chosen
V,chosen	17.641	17.641	17.641	ft/min	
LUB	1.503	1.503	1.503	ft	
LUB	0.458	0.458	0.458	m	
MTZ	3.006	3.006	3.006	ft	2-3 in good unit
MTZ	0.916	0.916	0.916	m	
A	0.056	0.056	0.056	psimin/ cPft^2	4-8 mesh bead [3]
B	0.0000889	0.0000889	0.0000889	psimin/ cPft^2	4-8 mesh bead [3]
DP/L	2.90E-01	2.90E-01	2.90E-01	psi/ft	
DP/L	6.56E-02	6.56E-02	6.56E-02	bar/m	
Area	10.073	5.036	3.358	m^2	
Diameter	3.581	2.532	2.068	m^2	
Re	2.84E+06	2.01E+06	1.64E+06	-	
Cycle time	**20**	**20**	**20**	h	assume complete water removal
Heat	**7.16**	**7.16**	**4.65**	h	for 2 beds heat + cool=0.5*cycle time chosen for 1bed
Cool	12.84	2.84	2.02	h	
Effective ed mass	27844.248	13922.124	9281.4159	kg	Water mass flow*3600*cycle time/ ads. Capacity
Bed mass	32293.606	16146.803	10764.535	kg	Effective bed mass + MTZ * Area*(1-porosity) * rho_bed
Bed volume	46.884545	23.442273	15.628182	m3	
Vessel V	66.977922	33.488961	22.325974	m^3	Bed volume / (1-porosity)
Bed length	6.6495783	6.6495783	6.6495783	m	Vessel volume / Vessel Area
L/D	1.8568219	2.6259428	3.21611	-	>1 good
Delta P	4.36E-01	4.36E-01	4.36E-01	bar	
Total 3A	205387.33	154040.5	136924.89	$	
Costs/bed	102693.67	51346.833	34231.222	$	
Delta costs	51346.833	0	-17115.611	$	

Regeneration					
Regen.time	6.98	6.98	4.65	h	
Cpbed	**1.0467**	**1.0467**	**1.0467**	kJ/kgK	from Hersch [6], page 55
Tregen.	**375**	**375**	**375**	K	estimate/choose from page 14 [3]
dT,min,heat	**5**	**5**	**5**	K	chosen minimal temperature difference
DTln	76.66	76.66	76.66	K	estimate a constant bed temp of 375 K, for which has been corrected in the 2.5x Qwater
Tgas,in	**700**	**700**	**700**	K	estimate/choose, above Tregen
Tgas,uit	380	380	380	K	5K above Tregen
Tcool	**278.15**	**278.15**	**278.15**	K	choose Tbed for water adsorption
Tcool,avg	326.575	326.575	326.575	K	for use in dTln,tijd,cool
Tgas,cool	**240.3**	**240.3**	**240.3**	K	choose cooling gas stream
Tgas,cool,uit	270.15	270.15	270.15	K	8K below Tbed
dT,min,cool	**8**	**8**	**8**	K	chosen minimal temperature difference
dTln,cool	70.30	70.30	70.30	K	with average Thot of bed = Tcool,avg.
Water abs.	3146.4	1573.2	1048.8	kg	Effective bed mass * adsorption capacity
Heat needed	**4186.8**	**4186.8**	**4186.8**	kJ/kg	1800 BTU/lb, page 10 [3]
Qwater	32933368.8	16466684.4	10977789.6	kJ	Heat needed for water adsorption and evaporation, corrected by 2.5x for losses and bed heating
Qcooling	3273696.3	1636848.2	1091232.1	kJ	Heat needed for cooling from Tregen, not corrected for thermal loss.
Cp,gas	**3.4266089**	**3.4266089**	**3.4266089**	kJ/kgK	110 bar
rho,gas	**33.703**	**33.703**	**33.703**	kg/m^3	
flow gas,heat	4.8608803	2.4304402	2.4963605	kg/s	Qtotal/(regen. time (h)*3600 * Cp,gas * (dTln))
flow gas,heat	0.1442269	0.0721135	0.0740694	m^3/s	
v,gas,heat	0.0143189	0.0143189	0.0220609	m/s	Superficial
Space time	0.1289979	0.1289979	0.0837277	h	
Qcool	3273696.3	1636848.2	1091232.1	kJ	
flow gas,cool	0.424926	0.9618078	0.901643	kg/s	
Cp,gas	**2.3718136**	**2.3718136**	**2.3718136**	kJ/kgK	30 bar
rho,gas,cool	**32.7745**	**32.7684**	**32.7684**	kg/m^3	
flow gas,cool	0.0129651	0.0293517	0.0275156	m^3/s	
v,gas,cool	0.142467	0.0629301	0.0671293	m/s	
Space time	0.0129651	0.0293517	0.0275156	h	

Appendix 8-3: Stream properties of the process streams going through the PFHE.

stream 3	Condensing			T [K]	257.05	256.07	251.1	245	
				ρ-tot. [kg.m^3]	111.816	112.17	119.11	129.97	
	liquid properties								
	P	T	Q	MW	ρ	Cp	μ	λ	
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	
	7.25E+06	257.05	7.7	35.0184	379.6	2784.61	6.77E-05	0.098	
	7.25E+06	256.07	7.5	34.792	378.15	2790.08	6.74E-05	0.0977	
	7.25E+06	251.1	4.7	33.011	362.51	2865.62	6.33E-05	0.097	
	7.25E+06	245	0	30.8451	341.37	2989.25	5.78E-05	0.0958	
	gas properties								
	P	T	Q	MW	ρ	Cp	μ	λ	q
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	[kg/kg]
	7.25E+06	257.05	7.7	19.5581	102.071	3630.24	1.36E-05	0.0377	0.881
	7.25E+06	256.07	7.5	19.55	102.26	3641.18	1.36E-05	0.0377	0.87895
	7.25E+06	251.1	4.7	19.33	106.21	3875.1	1.37E-05	0.038	0.8468
	7.25E+06	245	0	19.11	112.97	4259.07	1.39E-05	0.0388	0.8044

stream 12	Evaporating			T [K]	207.5	223.73	226.8	235	
				ρ-tot [kg/m^3]	366.26	165.607	154.29	126.62	
	liquid properties								
	P	T	Q	MW	ρ	Cp	μ	λ	
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	
	3.00E+06	207.5	0	29.6	366.26	2643.78	7.53E-05	0.1051	
	3.00E+06	223.73	1.75	33	419.99	2566.58	9.09E-05	0.1097	
	3.00E+06	226.8	2	33.43	425.78	2564.57	9.23E-05	0.1099	
	3.00E+06	235	2.5	34.72	441.23	2568.81	9.58E-05	0.1131	
	gas properties								
	P	T	Q	MW	Rhoρ	Cp	μ	Lambdaλ	q
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	[kg/kg]
	3.00E+06	207.5	0						0
	3.00E+06	223.73	1.75	19.18	39.47	2485.43	9.76E-06	0.0254	0.159
	3.00E+06	226.8	2	19.42	39.33	2454.99	9.83E-06	0.0258	0.179
	3.00E+06	235	2.5	20.24	39.11	2370.3	1.00E-05	0.0269	0.2417

stream 20		T [K]	206.14	207.5	223.73	226.8	235	24
		ρ-tot [kg/m^3]	44.3527	43.496	36.7806	35.9221	33.6359	31.8

liquid properties

	P	T	Q	MW	ρ	Cp	μ	λ
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]
	3.00E+06	206.14	0					
	3.00E+06	207.5	0.5					
	3.00E+06	223.73	5.5					
	3.00E+06	226.8	6.5					
	3.00E+06	235	9					
	3.00E+06	242.5	11.4					

gas properties

	P	T	Q	MW	ρ	Cp	μ	λ	q
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	[kg/kg]
	3.00E+06	206.14	0	18.351	44.3527	2825.26	9.35E-06	0.0246	
	3.00E+06	207.5	0.5	18.351	43.496	2780.45	9.37E-06	0.0246	
	3.00E+06	223.73	5.5	18.351	36.7806	2489.3	9.72E-06	0.0258	
	3.00E+06	226.8	6.5		35.9221	2462.15	9.78E-06	0.026	
	3.00E+06	235	9		33.6359	2395.58	9.99E-06	0.0268	
	3.00E+06	242.5	11.4		31.8364	2353.66	1.02E-05	0.0275	

stream 23	Condensing	T [K]	240.15	238.9	221.5	220
		ρ-tot [kg/m^3]	46.9059	48.3113	77.4827	80.6504

Liquid properties

	P	T	Q	MW	ρ	Cp	μ	λ
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]
	3.00E+06	240.15	8.7					
	3.00E+06	238.9	7.5	28.0477	395.894	3358.92	7.15E-05	0.1078
	3.00E+06	221.5	0.5	26.5995	368.123	3220.49	6.76E-05	0.1069
	3.00E+06	220	0	26.4508	364.795	3214.5	6.70E-05	0.1066

gas properties

	P	T	Q	MW	ρ	Cp	μ	λ	q
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	[kg/kg]
	3.00E+06	240.15	8.7	22.9469	46.9059	2427.95	1.02E-05	0.0245	
	3.00E+06	238.9	7.5	22.732	46.5952	2427.98	1.02E-05	0.0245	0.967
	3.00E+06	221.5	0.5	20.3226	44.4993	2551.9	9.78E-05	0.0245	0.582
	3.00E+06	220	0	20.1361	44.2933	2567.27	9.73E-06	0.0245	0.554

Stream 7	Evaporating		T [K]	223.73	226.8	235	242.5	250	
			ρ-tot [kg/m^3]	126.64	119.343	104.28	93.5804	8408245	
				liquid properties					
	P	T	Q	MW	ρ	Cp	μ	λ	
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	
	3.00E+06	223.73	0	38.4241	454.919	2330.27	0.00011	0.1126	
	3.00E+06	226.8	0.3	39.1108	462.142	2328.14	0.00011	0.1129	
	3.00E+06	235	1.6	40.7928	477.468	2338	0.00012	0.1129	
	3.00E+06	242.5	2	42.2445	488.021	2342.37	0.00012	0.1122	
	3.00E+06	250	2.2	43.657	495.997	2357.89	0.00012	0.111	
				gas properties					
	P	T	Q	MW	ρ	Cp	μ	λ	q
Units	[Pa]	[K]	[MW]	[kg/kmol]	[kg/m^3]	J/kg.K	[Pa.s]	[W/m.K]	[kg/kg]
	3.00E+06	223.73	0	18.3606	37.2198	2508.26	9.69E-05	0.0259	0.38597
	3.00E+06	226.8	0.3	18.5948	36.7706	2472.42	9.77E-06	0.026	0.41033
	3.00E+06	235	1.6	19.1344	35.9288	2393.41	9.97E-06	0.0264	0.4669
	3.00E+06	242.5	2	19.6905	35.3422	2337.35	1.10E-05	0.0268	0.51274
	3.00E+06	250	2.2	20.2795	34.9029	2291.14	1.03E-05	0.0271	0.5551

Appendix 8-4 calculation plate-fin heat exchanger, example 6.4 [30]

step 1

Block dimensions			max design pressure
W	H	L	assumption, according to example
[m]	[m]	[m]	80 bar
0.6	0.9	6.2	

step 2,3,4 6 zones counterflow heat curves are given in appendix =1=

step 5

zones	zone 1	zone 2	zone 3	zone 4	zone 5	zone 6
T,in,small	207.5	223.73	241.7	235	256.07	257.05
T,out,small	206.14	207.5	238.9	226.8	251.1	256.07
T,in,large	221.5	238.9	226.8	251.1	242.5	250
T,out,large	220	221.5	223.73	241.7	235	242.5

maximum NTU		zone 1	zone 2	zone 3	zone 4	zone 5	zone 6	
R	[-]	0.906667	0.932759	0.912052	0.87234	0.662667	0.130667	(Tin-Tout)small/(Tin-Tout)large
E	[-]	0.097656	0.55414	0.17084	0.386831	0.355956	0.515464	(Tin-Tout)large/(T1,in-T2,in)
NTU	[-]	0.107682	1.193647	0.204196	0.606755	0.506792	0.75326	ln[(1-ER)/(1-E)]/(1-R)
delta	[-]	9.286588	0.837769	4.897263	1.648112	1.973197	1.327563	(1-R)/LN((1-ER)/(1-E))
LMTD	K	13.92988	14.57718	15.0346	15.49225	14.79897	9.956724	(Tin-Tout)large*delta

step 6,7

Fins		
non perforated	p,h	1
	d,hole	0

aluminium selected material
thermal conductivity
168 W/mK

		sigma,f
UTS	9.60E+07	24000000
0.2% proof stress	3.45E+07	23000000

step 8

	stream	2	12	21	9	4	8	
Pdesign	Pa	7.50E+06	3.50E+06	3.50E+06	3.50E+06	3.50E+06	3.50E+06	
sigma,f	Pa	2.40E+07	2.40E+07	2.40E+07	2.40E+07	2.40E+07	2.40E+07	
p,h		1	1	1	1	1	1	
d,hole	m	0	0	0	0	0	0	
(nt)	[-]	0.3125	0.145833	0.145833	0.145833	0.145833	0.145833	P*p,h/(sigma,f(p,h-d,hole))

step 9,10 table 6/3 p 3.23

stream 2		wavy
aantal	n	787 /metre
dikte	t	6.10E-04 m thick
hoogte	b	0.0071 mm
diameter	d,h	0.051714 mm
opp	Ax	0.002575 m2

stream 12,9,4,8		serrated	
aantal	n	787 /metre	
dikte	t	5.08E-04 m thick	
hoogte	b	0.0089 mm	
diameter	d,h	0.10693 mm	2Ax/W(1-nt+n(b-t))
opp	Ax	0.004136 m2	W(b-t)(1-nt)

step 11

Edge bar minimum thickness			t,b	minimum leverbaar is 0.015 m		
b (gap betw. sheets)	sigma		P	t,b		
0.0071		2.40E+07	7.50E+06	0.004438	b*sqr(1.25*P/sigma)	0.0115
0.0089		2.40E+07	3.50E+06	0.0038		0.0115

effective width for heat transfer W				opp/layer	Ax	
0.577	m	0.6-2*t,b		0.002575	m2	W(b-t)(1-nt)
				0.004136	m2	

Step 12: for each zone, the number of layers and the following block height is calculated here. Other important results of this step are the Area per layer (A_x) and the total area ($A_{x,t}$)

zone 1

	9	**8**		
Pr	2.825426	1.066054	[-]	
Cp	2843.419	2802.856	J/kg.K	Stream
μ	5.96E-05	9.36E-06	Pa.s	properties
λ	0.059984	0.0246	W/m.K	appendix

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	units		
size	m	0.6x0.41x6.2	length calculated in step 15
zones	[-]	6	
flow	[-]	counterflow	
λ	W/m.K	168	

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stream		**9**	**8**
dP	Pa	7000	3000
NTU	[-]	1.193647	1.193647
ρ,gas	kg/m3	44.3963	43.92435
Pr	[-]	2.825426	1.066054
j/f	[-]	0.196	0.107
m’	kg/m^2.s	583.1347	186.1834
Re	[-]	18804.63	38247.6
M’	kg/s	39.9892	118.9134
A,x,t	m^2	0.068576	0.63869
A,x	m^2	0.004136	0.004136
d,h	m	0.001922	0.001922
b	m	0.0089	0.0089
N	[-]	16.58021	154.4208
N,no red.	[-]	32	177
Bf	[-]	1.777778	1.777778

calculation block height

block height (5.8) H 0.406565 m

sum N	171.0011	ΣN total layers
sum N*b	0.147564	Σ(N_i*b_i) b= fin height
tp	0.0015	parting sheet thickness
tc	0.002	cap thickness

zone 2		**9**	**4**	**8**	
	Pr	3.556335	1.942725	0.997854	[-]
	Cp	2670.149	2496.864	2634.875	J/kg.K
	μ	5.75E-05	7.68E-05	9.54E-06	Pa.s
	λ	0.043169	0.098764	0.0252	W/m.K

	units		
size	m	0.6x1.89x	length calculated in step 15
zones	[-]	6	
flow	[-]	counterflow	
λ	W/m.K	168	

stream		**9**	**4**	**8**
dP	Pa	6000	6000	3000
NTU	[-]	1.193647	1.193647	1.193647
ρ,gas	kg/m^3	121.36	362.8737	40.15
Pr	[-]	3.556335	1.942725	0.997854
j/f	[-]	0.196	0.18	0.107
m'	kg/m^2 s	124.5681	215.8064	50.95239
Re	[-]	4084.502	5294.403	10065.24
M'	kg/s	39.9892	19.9329	118.9134
A,x,t	m^2	0.321023	0.092365	2.333814
A,x	m^2	0.004136	0.004136	0.004136
d,h	m	0.001885	0.001885	0.001885
b	m	0.0089	0.0089	0.0089
N	[-]	77.61613	22.33172	564.264
N,no red.	[-]	65	10	174
Bf	[-]	1.777778	1.777778	1.777778

block height **H** **1.888354**

sum N	664.2119	ΣN total layers
sum N*b	0.889536	$\Sigma(N_i*b_i)$ b= fin height
tp	0.0015	parting sheet thickness
tc	0.002	cap thickness

zone 3

	12	8	9	4	
Pr	2.730405	0.932124	1.026436	2.077404	[-]
Cp	2393.359	2475.722	2415.633	2549.458	J/kg.K
μ	8.92E-05	9.75E-06	1.06E-05	7.78E-05	Pa.s
λ	0.078191	0.0259	0.024984	0.095453	W/m.K

	units		
size	m	0.6x2.84x	length calculated in step 15
zones	[-]	6	
flow	[-]	counterflow	
λ	W/m.K	168	

stream		**12**	**8**	**9**	**4**
dP	Pa	7000	6000	10000	10000
NTU	[-]	1.193647	1.193647	1.193647	1.193647
ρ,gas	kg/m^3	126.6397	37.3641	46.906	358.076
Pr	[-]	2.730405	0.932124	1.026436	2.077404
j/f	[-]	0.107	0.11	0.11	0.13
m’	kg/m^2 s	433.5555	141.258	212.2906	572.5508
Re	[-]	9163.013	27309.19	37699.13	13877.69
M’	kg/s	28.056	118.9134	39.9892	19.9329
A,x,t	m^2	0.064711	0.841817	0.18837	0.034814
A,x	m^2	0.004136	0.004136	0.004136	0.004136
d,h	m	0.001885	0.001885	0.001885	0.001885
b	m	0.0089	0.0089	0.0089	0.0089
N	[-]	15.64578	203.5325	45.54368	8.417295
N,no red.	[-]	34	183	57	12
Bf	[-]	1.777778	1.777778	1.777778	1.777778

calculation block height

block height (5.8) H 2.843148 m

sum N	273.1393	ΣN total layers
sum N*b	2.430939	$\Sigma(N_i*b_i)$ b= fin height
tp	0.0015	parting sheet thickness
tc	0.002	cap thickness

zone 4	**2**	**12**	**8**	**4**	
Pr	1.749368	2.209765	0.90949	2.10431	[-]
Cp	3868.33	2376.864	2428.865	2534.287	J/kg.K
μ	2.19E-05	6.96E-05	9.89E-06	7.64E-05	Pa.s
λ	0.048515	0.074872	0.0264	9.20E-02	W/m.K

	units		
size	m	0.6x4.08x	length calculated in step 15
zones	[-]	6	
flow	[-]	counterflow	
λ	W/m.K	168	

stream		**2**	**12**	**8**	**4**
dP	Pa	10000	7000	7000	10000
NTU	[-]	1.193647	1.193647	1.193647	1.193647
ρ,gas	kg/m^3	151.8558	126.6397	37.3641	350.5672
Pr	[-]	1.749368	2.209765	0.90949	2.10431
j/f	[-]	0.3	0.16	0.11	0.158
m'	kg/m^2 s	1281.674	304.3087	89.73535	345.4876
Re	[-]	521302.1	8241.758	17113.24	8529.26
M'	kg/s	143.1106	28.056	118.9134	19.9329
A,x,t	m^2	0.111659	0.092196	1.325157	0.057695
A,x	m^2	0.002575	0.004136	0.004136	0.004136
d,h	m	0.008924	0.001885	0.001885	0.001885
b	m	0.0071	0.0089	0.0089	0.0089
N	[-]	43.37116	22.29089	320.3932	13.94935
N,no red.	[-]	116	26	182	11
Bf	[-]	1.777778	1.777778	1.777778	1.777778

calculation block height

block height (5.8) H 4.08448 m

sum N	400.0046	ΣN total layers
sum N*b	3.481973	$\Sigma(N_i*b_i)$ b= fin height
tp	0.0015	parting sheet thickness
tc	0.002	cap thickness

zone5		**2**	**12**	**8**	
	Pr	1.635363	2.210735	0.882502	[-]
	Cp	3630.574	2352.525	2374.622	J/kg.K
	μ	2.07E-05	6.62E-05	1.01E-05	Pa.s
	λ	0.046009	0.07045	0.02715	W/m.K

	units		
size	m	0.6x3.29x	length calculated in step 15
zones	[-]	6	
flow	[-]	counterflow	
λ	W/m.K	168	

stream		**2**	**12**	**8**
dP	Pa	8000	7000	7000
NTU	[-]	1.193647	1.193647	1.193647
ρ,gas	kg/m^3	140.7233	126.6397	37.3641
Pr	[-]	1.635363	2.210735	0.882502
j/f	[-]	0.3	0.16	0.25
m’	kg/m^2 s	925.4739	328.0762	117.126
Re	[-]	398501.9	9342.402	21884.13
M’	kg/s	143.1106	28.056	118.9134
A,x,t	m^2	0.154635	0.085517	1.01526
A,x	m^2	0.002575	0.004136	0.004136
d,h	m	0.008924	0.001885	0.001885
b	m	0.0071	0.0089	0.0089
N	[-]	60.06403	20.67603	245.4672
N,no red.	[-]	50	20	222
Bf	[-]	1.777778	1.777778	1.777778

calculation block height

block height (5.8) H 3.286941 m

sum N	326.2073	ΣN total layers
sum N*b	2.79513	$\Sigma(N_i * b_i)$ b= fin height
tp	0.0015	parting sheet thickness
tc	0.002	cap thickness

zone 6

	2	12	
Pr	1.579358	2.1872	[-]
Cp	3533.885	2330.97	J/kg.K
μ	2.01E-05	6.23E-05	Pa.s
λ	0.04492	0.066404	W/m.K

	units		
size	m	0.6x0.81x	length calculated in step 15
zones	[-]	6	
flow	[-]	counterflow	
λ	W/m.K	168	

stream		**2**	**12**
dP	Pa	8000	7000
NTU	[-]	1.193647	1.193647
ρ,gas	kg/m^3	135.3776	126.6397
Pr	[-]	1.579358	2.1872
j/f	[-]	0.3	0.16
m'	kg/m^2 s	918.336	246.937
Re	[-]	408212.1	7471.53
M'	kg/s	143.1106	28.056
A,x,t	m^2	0.155837	0.113616
A,x	m^2	0.002575	0.004136
d,h	m	0.008924	0.001885
b	m	0.0071	0.0089
N	[-]	60.53088	27.46981
N,no red.	[-]	133	26
Bf	[-]	1.777778	1.777778

calculation block height

block height (5.8) H 0.808752 m

sum N	88.00069	ΣN total layers
sum N*b	0.674251	$\Sigma(N_i*b_i)$ b= fin height
tp	0.0015	parting sheet thickness
tc	0.002	cap thickness

Step 15: The stream properties needed to calculate the dimensions of the plate-fin heat exchanger for each zone. The formula numbers refer to example 6.4 of [30].

zone 1	**units**	**Stream 20**	**Stream 23**	
T,in	K	206.14	221.5	Temperature,in stream
T,uit	K	207.5	220	Temperature,out stream
M'	kg/s	118.9134	39.9892	Total mass flow
m'	kg/s.m^2	186.1834	583.1347	Mass velocity (6.16)
Ax,t	m^2	0.6386896	0.068576	Total flow cross sectional area (6.17)
d,h	m	0.0019221	0.001922	Flow passage hydrolic diameter (5.17)
q	kg/kg	1	0.56858	Vapour quality as mass fraction
dH/dT	J/kgK	...	8336	Enthalpy/temperature gradient
ρ,l	kg/m^3	....	366.4589	Liquid density
ρ,g	kg/m^3	43.92435	44.3963	Vapour density
μ,l	Pa.s	....	6.73E-05	Dynamic viscosity of liquid
μ,g	Pa.s	9.36E-06	5.38E-05	Dynamic viscosity of vapour
Cp,l	J/kg.K	...	3217.492	Liquid specific heat
Cp,g	J/kg.K	2802.8562	2559.584	Vapour specific heat
λ,l	W/m.K	...	0.10675	Liquid thermal conductivity
λ,g	W/m.K	0.0246	0.0245	Vapour themal conductivity
Re,Tl	[-]		16653.22	Reynolds,total flow (5.21)
j,Tl	[-]		0.008	Colburn J factor from figure 5.2
Re,l	[-]		20848.28	Reynolds,liquid flow (5.30)
J	[-]		0.007	Colburn J factor from figure 5.2
Re,g	[-]	38247.601	11853.91	Reynolds,gas flow (5.25/***5.13***^{1})
j,g	[-]	0.007	0.01	Colburn J factor from figure 5.2
f,g	[-]	0.056	0.056	Friction factor from figure 5.2
Pr,l	[-]	....	2.03	Prandtl number,all liquid (5.22)
Pr,g	[-]	1.07	5.62	Prandtl number,all vapour (5.26)
α,gas	W/m^2.K	3500.35	4721.02	Gas heat transfer coefficient (5.27)
α,Tl	W/m^2.K	....	9364.31	Liquid heat transfer coefficient (5.23)
α,cond	W/m^2.K	...	21198.60	α,Tl corrected for the vapour volume (5.24)
φ	[-]		2.26	[(q*ρ,l /ρ,g) + 1-q] page 3.7
Z	[-]	...	0.17	[q*Cp,g/(dH/dT)] page 3.8
α,c	W/m^2.K	...		α,Tl corrected with boiling factor F (5.32)
St	[-]	0.0067	0.0080	[α/(m'*Cp,total)
α	W/m^2.K	3500.35	11883.12	Heat transfer coefficient (5.29/***5.19***)
.rd	M^2.K/W	0.0001	0.0001	Fouling factor stated in BOD
α,eff	W/m^2.K	2592.78	5430.27	Effective heat transfer coefficient (5.20)
.t	M	0.0005	0.0005	Fin thickness,selected in step 9
λ,Al	W/m.K	168.00	168.00	Aluminium thermal conductivity
.hf(finh)	M	0.0089	0.0089	Fin height,selected in step 10
X	[-]	1.10	1.59	[hf*(α,eff/(2*t*λ,Al))$^{0.5}$] page 3.4
Tanhx	[-]	0.80	0.92	

ε,eff	[-]	0.73	0.58	Fin efficiency (5.10)
S,eff	M^2/m^2	1750.29	155.22	Overall effective heat transfer surface (5.7)
α,m*S,m	W/m^2.K	7894433.23	231968.24	Local effective heat flow gradient (5.9)
(αS)eff	W/m^2.K	2881621.08	181905.44	Effective heat flux gradient (5.1a)
.dP/L	Pa/m	7416.97	17431.32	Friction pressure loss (7.1)[2]
(αS)eff,c/w	W/m^2.K	2881621.08	181905.44	Sum (αS)eff for cold/hot streams
(αS)flow	W/m^2.K	171104.29		Overall incremental heat flow (6.19)
.dQ/MTD	W/K	39882.29		incremental duty from heat curves/mean Te perature difference (5.6), corrected for heat loss to environment with factor 0.9
L,2	m	0.40		Zone length (6.20)
.dP,f	Pa	2996.19	7041.64	[dP/L*L]

[1] the formulas in bold italics are used only in the single phase heat transfer calculations of stream 20

[2] Pressure loss is calculated, assuming single phase behaviour.

zone2	**Units**	**23**	**20**	**12**	
T,in	K	238.9	207.5	207.5	Temperature,in stream
T,uit	K	221.5	223.73	223.73	Temperature,out stream
M'	Kg/s	39.9892	118.913	19.9329	Total mass flow
m'	Kg/s.m^2	124.568	50.9524	215.806	Mass velocity (6.16)
Ax,t	m^2	0.32102	2.33381	0.09236	Total flow cross sectional area (6.17)
d,h	M	0.00189	0.00189	0.00189	Flow passage hydrolic diameter (5.17)
q	Kg/kg	0.77467	1	0.0795	Vapour quality as mass fraction
dH/dT	J/kgK	10060		5225.87	Enthalpy/temperature gradient
ρ,l	Kg/m^3	382.009	...	393.125	Liquid density
ρ,g	Kg/m^3	45.5473	40.1383	19.735	Vapour density
μ,l	Pa.s	7E-05	...	8.3E-05	Dynamic viscosity of liquid
μ,g	Pa.s	5.4E-05	9.54E-06	4.9E-06	Dynamic viscosity of vapour
Cp,l	J/kg.K	3289.71	...	2605.18	Liquid specific heat
Cp,g	J/kg.K	2489.94	2634.88	1242.72	Vapour specific heat
λ,l	W/m.K	0.1074	...	0.1051	Liquid thermal conductivity
λ,g	W/m.K	0.0245	0.0252	0.0254	Vapour themal conductivity
Re,Tl	[-]	3375.85		4898.23	Reynolds,total flow (5.21)
j,Tl	[-]	0.01		0.009	Colburn J factor from figure 5.2
Re,l	[-]	4350.12		4508.82	Reynolds,liquid flow (5.30)
j	[-]	0.0095			Colburn J factor from figure 5.2
f,l	[-]			0.056	Friction factor from figure 5.2
Re,g	[-]	3369.9	10065.2	6625.91	Reynolds,gas flow (5.25/***5.13***[1])
j,g	[-]	0.011	0.0062		Colburn J factor from figure 5.2
f,g	[-]	0.056	0.056	0.056	Friction factor from figure 5.2
Pr,l	[-]	2.13	....	2.06	Prandtl number,all liquid (5.22)
Pr,g	[-]	5.49	0.998		Prandtl number,all vapour (5.26)
α,gas	W/m^{2}K	1096.17	833.56		Gas heat transfer coefficient (5.27)

α,Tl	W/m^{2}K	2473.38	....	2925.33	Liquid heat transfer coefficient (5.23)
α,cond	W/m^{2}K	6412.95	...		α,Tl corrected for the vapour volume (5.24)
φ	[-]	2.593	...		[(q*ρ,l /ρ,g) + 1-q] page 3.7
Z	[-]	0.192	...		[q*Cp,g/(dH/dT)] page 3.8
X	[-]	...	...	2.59	[(1-q)/q*[(f,l*ρ,g)/(f,g*ρ,l)]$^{\wedge 0.5}$] page 3.9
F	[-]	...	...	1.61	[2.35*(1/X+0.213)$^{0.736}$]
α,c	W/m^{2}K	...	...	4711.35	α,Tl corrected with boiling factor F (5.32)
St	[-]	0.00974	0.00621		[α/(m'*Cp,total)
α	W/m^{2}K	3022.51	833.563	4711.35	Heat transfer coefficient (5.29/***5.19***)
rd	m^{2}K/W	0.0001	0.0001	0.0001	Fouling factor stated in BOD
α,eff	W/m^2.K	2320.99	769.426	3202.53	Effective heat transfer coefficient (5.20)
t	M	5.08E-04	5.08E-04	5.08E-04	Fin thickness,selected in step 9
l,m	W/m.K	168	168	168	Aluminium thermal conductivity
hf(finh)	M	0.0089	0.0089	0.0089	Fin height,selected in step 10
x	[-]	1.038	0.598	1.219	[hf*(α,eff/(2*t*λ,Al))$^{0.5}$] page 3.4
tanhx	[-]	0.777	0.535	0.839	
ε,eff	[-]	0.749	0.896	0.689	Fin efficiency (5.10)
S,eff	m^2/m^2	132.594	963.951	38.15	Overall effective heat transfer surface (5.7)
α,m*S,m	W/m^{2}K	1085902	7894433	1868391	Local effective heat flow gradient (5.9)
(αS)eff	W/m^{2}K	239792	677992	114678	Effective heat flux gradient (5.1a)
dP/L	Pa/m	1224.77	619.77	1229.79	Friction pressure loss (7.1)^2
(αS)eff,cw	W/m^{2}K	239792	792669.1		Sum (αS)eff for cold/hot streams
(αS)flow	W/m^{2}K	184099.5			Overall incremental heat flow (6.19)
dQ/MTD	W/K	5.34E+05			incremental duty from heat curves/mean Temperature difference (5.6), corrected for heat loss to environment with factor 0.9
L,2	M	5.023			Zone length (6.20)
dP,f	Pa	6151.88	3113.03	6177.12	[dP/L*L]

zone3	**Units**	**23**	**20**	**12**	**7**	
T,in	K	240.15	223.73	223.73	223.73	Temperature,in stream
T,uit	K	238.9	226.8	226.8	226.8	Temperature,out stream
M'	kg/s	39.9892	118.913	19.9329	28.056	Total mass flow
m'	kg/s.m^2	212.291	141.258	572.551	433.555	Mass velocity (6.16)
Ax,t	m^2	0.18837	0.84182	0.03481	0.06471	Total flow cross sectional area (6.17)
d,h	M	0.00189	0.00189	0.00189	0.00189	Flow passage hydrolic diameter (5.17)
q	kg/kg	0.98353	1	0.169	0.39815	Vapour quality as mass fraction
dH/dT	J/kgK	24006		5016.83		Enthalpy/temperature gradient
ρ,l	kg/m^3	197.947	...	422.885	458.53	Liquid density
ρ,g	kg/m^3	46.7506	36.3514	39.4	36.9952	Vapour density
μ,l	Pa.s	3.6E-05	...	9.16E-05	0.00011	Dynamic viscosity of liquid
μ,g	Pa.s	1E-05	9.75E-06	9.8E-06	5.3E-05	Dynamic viscosity of vapour
Cp,l	J/kg.K	1679.46	...	2565.58	2329.2	Liquid specific heat
Cp,g	J/kg.K	2427.96	2475.72	2470.21	2490.34	Vapour specific heat

λ,l	W/m.K	0.0539	...	0.1097	0.11275	Liquid thermal conductivity
λ,g	W/m.K	0.0245	0.0259	0.0254	0.02595	Vapour themal conductivity
Re,Tl	[-]	11190.2		11783.2	7238.63	Reynolds,total flow (5.21)
j,Tl	[-]	0.0075		0.011	0.005	Colburn J factor from figure 5.2
Re,l	[-]	39256.4		9791.81	4356.57	Reynolds,liquid flow (5.30)
j	[-]	0.004				Colburn J factor from figure 5.2
f,l	[-]			0.056	0.056	Friction factor from figure 5.2
Re,g	[-]	38609.9	27309.2	18621.7	6099.35	Reynolds,gas flow (5.25/***5.13***^{1})
j,g	[-]	0.004	0.0072			Colburn J factor from figure 5.2
f,g	[-]	0.056	0.056	0.056	0.056	Friction factor from figure 5.2
Pr,l	[-]	1.1144	....	2.14238	2.33263	Prandtl number,all liquid (5.22)
Pr,g	[-]	1.01033	0.93212			Prandtl number,all vapour (5.26)
α,gas	W/m^{2}K	2047.65	2638.81			Gas heat transfer coefficient (5.27
α,Tl	W/m^{2}K	2487.64	....	8382.31	1911.87	Liquid heat transfer coefficient (5
α,cond	W/m^{2}K	5086.5	...			α,Tl corrected for vapour volume
φ	[-]	2.04471	...			[(q*ρ,l /ρ,g) + 1-q] page 3.7
Z	[-]	0.09947	...			[q*Cp,g/(dH/dT)] page 3.8
X	[-]		...	1.5009	0.42937	[(1-q)/q*[(f,l*ρ,g)/(f,g*ρ,l)]$^{\wedge 0.5}$]pa
F	[-]		...	2.13767	4.66957	[2.35*(1/X+0.213)$^{0.736}$]
α,c	W/m^{2}K		...	17918.6	8927.64	α,Tl corrected for boiling factor (
St	[-]	0.00791	0.00755			[α/(m'*Cp,total)]
α	W/m^{2}K	4078.66	2638.81	17918.6	8927.64	Heat transfer coefficient (5.29/***5.1***
rd	m^{2}K/W	0.0001	0.0001	0.0001	0.0001	Fouling factor stated in BOD
α,eff	W/m^{2}K	2897.05	2087.86	6418.16	4716.72	Effective heat transfer coefficient
t	M	5.08E-04	5.08E-04	5.08E-04	5.08E-04	Fin thickness,selected in step 9
l,m	W/m.K	168	168	168	168	Aluminium thermal conductivity
hf(finh)	M	0.0089	0.0089	0.0089	0.0089	Fin height,selected in step 10
x	[-]	1.15949	0.98433	1.72581	1.47948	[hf*(α,eff/(2*t*λ,Al))$^{0.5}$] page 3.
tanhx	[-]	0.82087	0.75493	0.93856	0.90137	
ε,eff	[-]	0.70796	0.76695	0.54384	0.60925	Fin efficiency (5.10)
S,eff	m^2/m^2	77.8038	347.701	14.3795	26.7282	Overall effective heat transfer sur
α,m*S,m	W/m^{2}K	637187	2847557	117764	218895	Local effective heat flow gradient
(αS)eff	W/m^{2}K	166502	578477	51741.2	79996.6	Effective heat flux gradient (5.1a)
dP/L	Pa/m	8770.06	5259.75	8771.68	6195.89	Friction pressure loss (7.1)^2
(αS)eff,c/w	W/m^{2}K	166502	710214.7461			Sum (αS)eff for cold/hot streams
(αS)flow	W/m^{2}K	134880.88				Overall incremental heat flow (6.
dQ/MTD	W/K	8.87E+04				incremental duty from heat curve Temperature difference (5.6), cor heat loss to environment with fact
L,2	M	1.1395				Zone length (6.20)
dP,f	Pa	9993.63	5993.57	9995.48	7060.32	[dP/L*L]

zone 4	**units**	**20**	**12**	**7**	**3**	
T,in	K	226.8	226.8	226.8	251.1	Temperature,in stream
T,uit	K	235	235	235	241.7	Temperature,out stream
M’	kg/s	118.913	19.9329	28.056	143.111	Total mass flow
m’	kg/s.m^2	89.7354	345.488	304.309	1281.67	Mass velocity (6.16)
Ax,t	m^2	1.32516	0.05769	0.0922	0.11166	Total flow cross sectional area (6.17)
d,h	m	0.00189	0.00189	0.00189	0.00892	Flow passage hydrolic diameter (5.17)
q	kg/kg	1	0.21035	0.43862	0.8256	Vapour quality as mass fraction
dH/dT	J/kgK		3059		4582	Enthalpy/temperature gradient
ρ,l	kg/m^3	...	433.505	469.805	351.94	Liquid density
ρ,g	kg/m^3	34.779	39.22	36.3497	109.59	Vapour density
μ,l	Pa.s	...	9.41E-05	0.00012	6.05E-05	Dynamic viscosity of liquid
μ,g	Pa.s	9.89E-06	9.9E-06	9.87E-06	1.38E-05	Dynamic viscosity of vapour
Cp,l	J/kg.K	...	2566.69	2333.07	2927.44	Liquid specific heat
Cp,g	J/kg.K	2428.86	2412.65	2432.91	4067.09	Vapour specific heat
λ,l	W/m.K	...	0.1097	0.1129	0.0964	Liquid thermal conductivity
λ,g	W/m.K	0.0264	0.0254	0.0262	0.0384	Vapour themal conductivity
Re,Tl	[-]		6924.23	4933.7	188906	Reynolds,total flow (5.21)
j,Tl	[-]		0.01	0.014	0.0033	Colburn J factor from figure 5.2
f,tg	[-]					Friction factor from figure 5.2
Re,l	[-]		5467.72	2769.71	32945.3	Reynolds,liquid flow (5.30)
j	[-]					Colburn J factor from figure 5.2
f,l	[-]		0.056	0.057	0.01	Friction factor from figure 5.2
Re,g	[-]	17113.2	13818.8	25489.4	684996	Reynolds,gas flow (5.25/***5.13***[1])
j,g	[-]	0.0072			0.0062	Colburn J factor from figure 5.2
f,g	[-]	0.056	0.056	0.056		Friction factor from figure 5.2
Pr,l	[-]	....	2.20087	2.40294	1.83861	Prandtl number,all liquid (5.22)
Pr,g	[-]	0.90949			1.46002	Prandtl number,all vapour (5.26)
α,gas	W/m^{2}K	1671.79			20729.8	Gas heat transfer coefficient (5.27)
α,Tl	W/m^{2}K	....	4337.53	3489.99	8248.31	Liquid heat transfer coefficient (5.23)
α,cond	W/m^{2}K	...			13865.4	α,Tl corrected for vapour volume (5.2
φ	[-]	...			1.681	[(q*ρ,l /ρ,g) + 1-q] page 3.7
Z	[-]	...			0.73282	[q*Cp,g/(dH/dT)] page 3.8
X	[-]	...	1.12914	0.35918		[(1-q)/q*[(f,l*ρ,g)/(f,g*ρ,l)]$^{\wedge0.5}$]page 3.
F	[-]	...	2.51845	5.27135		[2.35*(1/X+0.213)$^{0.736}$]
α,c	W/m^{2}K	...	10923.9	18396.9		α,Tl corrected for boiling factor (5.32)
St	[-]	0.00767				[α/(m’*Cp,total)]
α	W/m^{2}K	1671.79	10923.9	18396.9	9304.65	Heat transfer coefficient (5.29/***5.19***)
rd	m^{2}K/W	0.0001	0.0001	0.0001	0.0001	Fouling factor stated in BOD
α,eff	W/m^{2}K	1432.33	5220.77	6478.49	4819.9	Effective heat transfer coefficient (5.2
t	m	5.08E-04	5.08E-04	5.08E-04	6.10E-04	Fin thickness,selected in step 9
l,m	W/m.K	168	168	168	168	Aluminium thermal conductivity
hf(finh)	M	0.0089	0.0089	0.0089	0.0071	Fin height,selected in step 10
x	[-]	0.81529	1.55652	1.73391	1.08879	[hf*(α,eff/(2*t*λ,Al))$^{0.5}$] page 3.4

tanhx	[-]	0.6725	0.91486	0.93952	0.79643	
ε,eff	[-]	0.82486	0.58776	0.54185	0.73149	Fin efficiency (5.10)
S,eff	m^2/m^2	547.338	23.8301	38.0803	59.6353	Overall effective heat transfer su
α,m*S,m	W/m^{2}K	4482517	195161	311865	601681	Local effective heat flow gradien
(αS)eff	W/m^{2}K	667268	75977.4	137741	194513	Effective heat flux gradient (5.1a
dP/L	Pa/m	2218.55	3262.49	2576.32	3034.28	Friction pressure loss (7.1)^2
(αS)eff,c/w	W/m^{2}K	880987.219			194513	Sum (αS)eff for cold/hot streams
(αS)flow	W/m^{2}K	159333.8294				Overall incremental heat flow (6.
dQ/MTD	W/K	2.87E+05				incremental duty from heat curve Temperature difference (5.6), co heat loss to environment with fac
L,2	M	3.12				Zone length (6.20)
dP,f	Pa	6922.9	10180.5	8039.32	9468.35	[dP/L*L]

zone5	**units**	**20**	**7**	**3**	
T,in	K	235	235	251.1	Temperature,in stream
T,uit	K	242.5	242.5	256.07	Temperature,out stream
M'	kg/s	118.913	28.056	143.111	Total mass flow
m'	kg/s.m^2	117.126	328.076	925.474	Mass velocity (6.16)
Ax,t	m^2	1.01526	0.08552	0.15463	Total flow cross sectional area (6.17)
d,h	m	0.00189	0.00189	0.00892	Flow passage hydrolic diameter (5.17)
q	kg/kg	1	0.48982	0.86288	Vapour quality as mass fraction
dH/dT	J/kgK			5202	Enthalpy/temperature gradient
ρ,l	kg/m^3	...	482.745	370.33	Liquid density
ρ,g	kg/m^3	32.7362	35.6355	104.235	Vapour density
μ,l	Pa.s	...	0.00012	6.5E-05	Dynamic viscosity of liquid
μ,g	Pa.s	1.01E-05	1E-05	1.4E-05	Dynamic viscosity of vapour
Cp,l	J/kg.K	...	2340.19	2827.85	Liquid specific heat
Cp,g	J/kg.K	2374.62	2365.38	3758.14	Vapour specific heat
λ,l	W/m.K	...	0.11255	0.09735	Liquid thermal conductivity
λ,g	W/m.K	0.02715	0.0266	0.03785	Vapour themal conductivity
Re,Tl	[-]		5167.54	126405	Reynolds,total flow (5.21)
j,Tl	[-]		0.013	0.0037	Colburn J factor from figure 5.2
Re,l	[-]		2636.38	17333.3	Reynolds,liquid flow (5.30)
f,l	[-]		0.057	0.02	Friction factor from figure 5.2
Re,g	[-]	21884.1	28868	522642	Reynolds,gas flow (5.25/***5.13***^{1})
j,g	[-]	0.007		0.007	Colburn J factor from figure 5.2
f,g	[-]	0.056	0.056	0.012	Friction factor from figure 5.2
Pr,l	[-]	....	2.48864	1.89787	Prandtl number,all liquid (5.22)
Pr,g	[-]	0.8825		1.35382	Prandtl number,all vapour (5.26)
α,gas	W/m^2.K	2116.18		17164.5	Gas heat transfer coefficient (5.27)
α,Tl	W/m^2.K	....	3171.32	6315.66	Liquid heat transfer coefficient (5.23)
α,cond	W/m^2.K	...		11302.7	α,Tl corrected for vapour volume (5.24)
φ	[-]	...		1.78963	[(q*ρ,l /ρ,g) + 1-q] page 3.7

Z	[-]	...		0.62338	[q*Cp,g/(dH/dT)] page 3.8
X	[-]	...	0.2855		[(1-q)/q*[(f,l*ρ,g)/(f,g*ρ,l)]$^{\wedge 0.5}$]page 3.9
F	[-]	...	6.17457		[2.35*(1/X+0.213)$^{0.736}$]
α,c	W/m^2.K	...	19581.5		α,Tl corrected for boiling factor (5.32)
St	[-]	0.00761			[α/(m'*Cp,total)]
α	W/m^2.K	2116.18	19581.5	8013.32	Heat transfer coefficient (5.29/***5.19***)
rd	m^2.K/W	0.0001	0.0001	0.0001	Fouling factor stated in BOD
α,eff	W/m^2.K	1746.58	6619.51	4448.55	Effective heat transfer coefficient (5.20)
t	m	5.08E-04	5.08E-04	6.10E-04	Fin thickness,selected in step 9
l,m	W/m.K	168	168	168	Aluminium thermal conductivity
hf(finh)	m	0.0089	0.0089	0.0071	Fin height,selected in step 10
x	[-]	0.90029	1.75268	1.046	[hf*(α,eff/(2*t*λ,Al))$^{0.5}$] page 3.4
tanhx	[-]	0.71644	0.94168	0.78025	
ε,eff	[-]	0.79579	0.53728	0.74593	Fin efficiency (5.10)
S,eff	m^2/m^2	419.34	35.3216	82.588	Overall effective heat transfer surface (5.7)
α,m*S,m	W/m^2.K	3434252	289272	833258	Local effective heat flow gradient (5.9)
(αS)eff	W/m^2.K	603667	129301	254975	Effective heat flux gradient (5.1a)
dP/L	Pa/m	4015.49	3980.33	4401.98	Friction pressure loss (7.1)^2
(αS)eff,c/w	W/m^2.K	732968.0842		254975	Sum (αS)eff for cold/hot streams
(αS)flow	W/m^2.K	189169.15			Overall incremental heat flow (6.19)
dQ/MTD	W/K	2.25E+05			
L,2	m	2.06			Zone length (6.20)
dP,f	Pa	8286.27	8213.73	9083.82	[dP/L*L]

zone6	**units**	**7**	**3**	
T,in	K	242.5	257.05	Temperature,in stream
T,uit	K	250	256.07	Temperature,out stream
M'	kg/s	28.056	143.111	Total mass flow
m'	kg/s.m^2	246.937	918.336	Mass velocity (6.16)
Ax,t	m^2	0.11362	0.15584	Total flow cross sectional area (6.17)
d,h	m	0.00189	0.00892	Flow passage hydrolic diameter (5.17)
q	kg/kg	0.53392	0.87998	Vapour quality as mass fraction
dH/dT	J/kgK		1426	Enthalpy/temperature gradient
ρ,l	kg/m^3	492.009	378.875	Liquid density
ρ,g	kg/m^3	35.1226	102.166	Vapour density
μ,l	Pa.s	0.00012	6.8E-05	Dynamic viscosity of liquid
μ,g	Pa.s	1.1E-05	1.4E-05	Dynamic viscosity of vapour
Cp,l	J/kg.K	2350.13	2787.35	Liquid specific heat
Cp,g	J/kg.K	2314.25	3635.71	Vapour specific heat
λ,l	W/m.K	0.1116	0.09785	Liquid thermal conductivity
λ,g	W/m.K	0.02695	0.0377	Vapour themal conductivity
Re,Tl	[-]	3832.68	121318	Reynolds,total flow (5.21)
j,Tl	[-]	0.0135	0.004	Colburn J factor from figure 5.2
f,tg	[-]			Friction factor from figure 5.2

Re,l	[-]	1786.33		Reynolds,liquid flow (5.30)
j	[-]			Colburn J factor from figure 5.2
f,l	[-]	0.057		Friction factor from figure 5.2
Re,g	[-]	23300.6	530250	Reynolds,gas flow (5.25/***5.13***)
j,g	[-]		0.006	Colburn J factor from figure 5.2
f,g	[-]	0.057	0.012	Friction factor from figure 5.2
Pr,l	[-]	2.55787	1.92422	Prandtl number,all liquid (5.22)
Pr,g	[-]		1.31156	Prandtl number,all vapour (5.26)
α,gas	W/m^2.K		14711.2	Gas heat transfer coefficient (5.27)
α,Tl	W/m^2.K	2273.66	6616.89	Liquid heat transfer coefficient (5.23)
α,cond	W/m^2.K		12171.1	α,Tl corrected for vapour volume (5.24)
φ	[-]		1.83939	[(q*ρ,l /ρ,g) + 1-q] page 3.7
Z	[-]		2.24357	[q*Cp,g/(dH/dT)] page 3.8
X	[-]	0.23323		[(1-q)/q*[(f,l*ρ,g)/(f,g*ρ,l)]$^{\wedge 0.5}$]page 3.9
F	[-]	7.11003		[2.35*(1/X+0.213)$^{0.736}$]
α,c	W/m^2.K	16165.8		α,Tl corrected for boiling factor (5.32)
St	[-]			[α/(m'*Cp,total)]
α	W/m^2.K	16165.8	4261.31	Heat transfer coefficient (5.29/***5.19***)
rd	m^2.K/W	0.0001	0.0001	Fouling factor stated in BOD
α,eff	W/m^2.K	6178.21	2988.02	Effective heat transfer coefficient (5.20)
t	m	5.08E-04	5.08E-04	Fin thickness,selected in step 9
l,m	W/m.K	168	168	Aluminium thermal conductivity
hf(finh)	m	0.0089	0.0089	Fin height,selected in step 10
x	[-]	1.69325	1.17755	[hf*(α,eff/(2*t*λ,Al))$^{0.5}$] page 3.4
tanhx	[-]	0.93456	0.82668	
ε,eff	[-]	0.55193	0.70203	Fin efficiency (5.10)
S,eff	m^2/m^2	46.9276	83.23	Overall effective heat transfer surface (5.7)
α,m*S,m	W/m^2.K	384321	839735	Local effective heat flow gradient (5.9)
(αS)eff	W/m^2.K	165259	191869	Effective heat flux gradient (5.1a)
dP/L	Pa/m	2397.13	2701.82	Friction pressure loss (7.1)[1]
(αS)eff,c/w	W/m^2.K	165259	191869	Sum (αS)eff for cold/hot streams
(αS)flow	W/m^2.K	88786.36		Overall incremental heat flow (6.19)
dQ/MTD	W/K	1.23E+05		
L,2	m	2.39		Zone length (6.20)
dP,f	Pa	5743.85	6473.94	[dP/L*L]

Appendix 8-5: Design product heat exchangers

Design Heat-exchanger Sales Gas: 1 shell, 2 tube passes

Number of units: 2

Data streams:

		inlet	**mean**	**outlet**	
C2-stream:	temperature	394.28	358.215	322.15	K
	thermal conductivity	0.0618	0.061	0.0602	W/m.K
	density	112.6719	135.1259	157.5799	kg/m3
	viscosity	1.88E-05	1.93E-05	1.98E-05	Pa.s
	flow	58.64095	58.64095	58.64095	kg/s
	avg molweight		18.3507		g/mol
	specific heat	2.91883	3.063196	3.207562	kJ/kg.K

Cooling water:	- Cp:	4.2	kJ/kg.K
	- Tin:	303.15	K
	- Tuit:	311.15	K
	- density	995	kg/m3
	- viscosity	8.00E-04	Ns/m2
	- thermal conductivity	5.90E-01	W/m.°C

Calculations:

type: one shell pass, two tube passes

Heat load: 12896.1 kW

Cooling water flow: 383.8125 kg/s

dTlm: -43.44949 °C

R = 0.110911
S = 0.791507

Ft = 0.940075

dTm = -40.84579 °C

U = 625 W/m2.°C

A = 505.1625 m2

Data tubes:

outside diameter:	38 mm
inside diameter:	33.8 mm
length:	4.88 m
material:	Carbon steel

Area one tube = 0.582282 m2

Number of tubes = 867.557 868

Square pitch: 1.25

Bundle diameter = 1639.13 mm

head type: Split-ring floating head

bundle diametrical clearance = 90 mm

Shell diameter = 1729.13 mm
1.72913 m

Tube cross-sectional area = 0.000286 m2

Tubes per pass = 434

Total flow area = 0.123955 m2

Volumetric flow = 0.433973 m3/s

tube-side velocity = 3.501058 m/s

Tube-side heat transfer coefficient:

Re = 8.28E+05

Pr = 0.969928

L/di = 144.3787

uit figuur 12.23
Jh = 1.70E-03

Nu = 1.39E+03

hi = 2514.467 W/m2.K

Shell-side coefficient:

Mean temperature = 307.15 K

Baffle spacing: 100 percentage of shell diameter

Baffle spacing = 1729.13 mm

Tube pitch = 47.5 mm

Cross-flow area = 597978.2 mm2
0.597978 m2

De = 36.93125 mm

Volumetric flow-rate = 0.385741 m3/s

Shell-side velocity = 0.645076 m/s

Re = 2.96E+04

Pr = 5.69E+00

baffles cut: 25%

jh 3.40E-03 fig. 12.29

hs = 2.86E+03 W/m2.K

Overall coefficient:

thermal conductivity metal = 17 W/m2.°C

fouling coefficients (hydrocarbons) 10 kW/m2.°C
fouling coefficients (water) 2 kW/m2.°C

1/U0 = 0.00159

U0 = 628.8879 W/m2.°C

Pressure drop:

Tube-side:

number of tubes = 868
passes = 2
I.d. = 33.8 mm
O.d = 38 mm
ut = 3.501058 m/s
Re = 8.28E+05

jf = 0.0018 fig. 12.24

dP = 7584.253 N/m2
0.075843 bar

Shell-side:

Gs = 98.06537 kg/s.m2

Linear velocity = 0.725733 m/s

Baffle cut: 25%

Re = 1.88E+05

jf = 3.00E-02 fig. 12.30

dP = 1128.495 N/m2
0.011285 bar

Costing:

Material:		Carbon steel		
Standard price:	hfl	175000 per		500 m2
Price:	hfl	176806.9		
number of units:		2		
pressure factor:		3		
Price (total):	hfl	1060841		

Design Heat-exchanger Sales Gas: Counter-current

Number of units: 2

Data streams:

		inlet	mean	outlet	
C2-stream:	temperature	394.28	358.215	322.15	K
	thermal conductivity	0.0618	0.061	0.0602	W/m.K
	density	112.6719	135.1259	157.5799	kg/m3
	viscosity	1.88E-05	1.93E-05	1.98E-05	Pa.s
	flow	58.64095	58.64095	58.64095	kg/s
	avg molweight		18.3507		g/mol
	specific heat	2.91883	3.063196	3.207562	kJ/kg.K

Cooling water:	- Cp:	4.2	kJ/kg.K
	- Tin:	303.15	K
	- Tuit:	311.15	K
	- density	995	kg/m3
	- viscosity	8.00E-04	Ns/m2
	- thermal conductivity	5.90E-01	W/m.°C

Calculation:

type:	counter-current	
Heat load:	12896.1	kW
Cooling water flow:	383.8125	kg/s
dTlm:	-43.44949	°C
dTm =	-43.44949	°C
U =	620	W/m2.°C
A =	478.7205	m2

Data tubes:

outside diameter:	25	mm
inside diameter:	19.8	mm
length:	4.88	m
material:	Carbon steel	

Area one tube =	0.38308	m2
Number of tubes =	1249.662	1250
Square pitch:	1.25	
Bundle diameter =	1269.51	mm
head type:	Split-ring floating head	
bundle diametrical clearance =	58	mm

Shell diameter =	1327.51 mm 1.32751 m
Tube cross-sectional area =	9.8E-05 m2
Tubes per pass =	1250
Total flow area =	0.122513 m2
Volumetric flow =	0.433973 m3/s
tube-side velocity =	3.542273 m/s

Tube-side heat transfer coefficient:

Re =	4.91E+05
Pr =	0.969928
L/di =	246.4646
using figure 12.23 Jh =	1.90E-03
Nu =	9.23E+02
hi =	2843.37 W/m2.K

Shell-side coefficient:

Mean temperature =	307.15 K
Baffle spacing:	100 percentage of shell diameter
Baffle spacing =	1327.51 mm
Tube pitch =	31.25 mm
Cross-flow area =	352456.7 mm2 0.352457 m2
De =	24.29688 mm
Volumetric flow-rate =	0.385741 m3/s
Shell-side velocity =	1.094436 m/s
Re =	3.31E+04
Pr =	5.69E+00
baffles cut:	25%

jh	3.20E-03	fig. 12.29
hs =	4.56E+03 W/m2.K	

Overall coefficient:

thermal conductivity metal =	17 W/m2.°C
fouling coefficients product:	10 kW/m2.°C
fouling coefficients water:	2 kW/m2.°C
1/U0 =	0.001566
U0 =	638.5695 W/m2.°C

Pressure drop:

Tube-side:

number of tubes =	1250	
passes =	1	
I.d. =	19.8 mm	
O.d =	25 mm	
ut =	3.542273 m/s	
Re =	4.91E+05	
jf =	0.002	fig. 12.24
dP =	10924.96 N/m2	
	0.10925 bar	

Shell-side:

Gs =	166.3777 kg/s.m2	
Linear velocity =	1.231279 m/s	
Re =	2.09E+05	
jf =	3.00E-02	fig. 12.30
dP =	4937.447 N/m2	
	0.049374 bar	

Costing:

Material:		Carbon steel	
Standard price	hfl	175000 per	500 m2
Price	hfl	167552.2	
Number of units:		2	
pressure factor:		3	
Price (total)	hfl	1005313	

Design Heat-exchanger C3-product

Number of units: 1

Data streams:

		inlet	mean	outlet	
C3-stream:	temperature	319.4615	314.8058	310.15	K
	thermal conductivity	0.0801	0.08215	0.0842	W/m.K
	density	450.6975	458.7726	466.8476	kg/m3
	viscosity	7.91E-05	8.4E-05	8.89E-05	Pa.s
	flow	11.3034	11.3034	11.3034	kg/s
	avg molweight		43.9732		g/mol
	specific heat	3.196639	3.086376	2.976112	kJ/kg.K

Cooling water:	- Cp:	4.2	kJ/kg.K
	- Tin:	303.15	K
	- Tuit:	311.15	K
	- density	995	kg/m3
	- viscosity	8.00E-04	Ns/m2
	- thermal conductivity	5.90E-01	W/m.°C

Calculation:

type: counter-current flow

Heat load: 324.69 kW

Cooling water flow: 9.663393 kg/s

dTlm: -7.636991 °C

dTm = -7.636991 °C

U = 550 W/m2.°C

A = 77.3008 m2

Data tubes:		
outside diameter:	20	mm
inside diameter:	16	mm
length:	3.66	m
material:	Carbon steel	

Area one tube = 0.229848 m2

Number of tubes = 336.3127 337

Square pitch: 1.25

Bundle diameter = 560.768 mm

head type: Split-ring floating head

bundle diametrical clearance = 58 mm

Shell diameter = 618.768 mm
0.618768 m

Standard diameter = ???

Tube cross-sectional area = 0.000064 m2

Tubes per pass = 337

Total flow area = 0.021568 m2

Volumetric flow = 0.024638 m3/s

tube-side velocity = 1.142357 m/s

Tube-side heat transfer coefficient:

Re = 9.98E+04

Pr = 3.155317

L/di = 228.75

figure 12.23:
Jh = 2.90E-03

Nu = 4.23E+02

hi = 2172.137 W/m2.K

Shell-side coefficient:

Mean temperature = 307.15 K

Baffle spacing: 45.45455 percentage of shell diameter

Baffle spacing = 281.2582 mm

Tube pitch = 25 mm

Cross-flow area = 34806.72 mm2
0.034807 m2

De = 19.4375 mm

Volumetric flow-rate = 0.009712 m3/s

Shell-side velocity = 0.279025 m/s

Re = 6.75E+03

Pr = 5.69E+00

baffles cut: 25%

jh 7.20E-03 fig. 12.29

hs = 2.62E+03 W/m2.K

Overall coefficient:

thermal conductivity metal = 17 W/m2.°C

fouling coefficient (product) 10 kW/m2.°C
fouling coefficient (water) 2 kW/m2.°C

1/U0 = 0.001814

U0 = 551.3324 W/m2.°C

Pressure drop:

Tube-side:

number of tubes = 337
passes = 1
I.d. = 16 mm
O.d = 20 mm
ut = 1.142357 m/s
Re = 9.98E+04

jf = 2.80E-03 fig. 12.24

dP = 2282.201 N/m2
0.022822 bar

Shell-side:

Gs = 324.7477 kg/s.m2

Linear velocity = 0.707862 m/s

Re = 7.52E+04

jf = 3.50E-02 fig. 12.30

dP = 13331.72 N/m2
0.133317 bar

Costing:

Material:	Carbon steel	
	44000 per	70 m2
Price:	48589.07	
pressure factor:	1.4	
number of units:	1	
Price:	68024.7	

Design Heat-exchanger C4-product

Data streams:

		inlet	mean	outlet	
C4-stream:	temperature	343.01	326.58	310.15	K
	specific heat	167187	156979	146771	J/kmol.K
	thermal conductivity	0.0837	0.0907	0.0977	W/m.K
	density	504.9235	527.5657	550.2079	kg/m3
	viscosity	0.000109	0.000129	0.000148	Pa.s
	flow	7.0567	7.0567	7.0567	kg/s
	avg molweight		58.2431		g/mol
	specific heat		2.695238		kJ/kg.K

Cooling water:	- Cp:	4.2	kJ/kg.K
	- Tin:	303.15	K
	- Tuit:	311.15	K
	- density	995	kg/m3
	- viscosity	8.00E-04	Ns/m2
	- thermal conductivity	5.90E-01	W/m.°C

Calculation:

type: counter-current flow

Heat load: 618.76 kW

Cooling water flow: 18.41548 kg/s

dTlm: -16.40446 °C

dTm = -16.40446 °C

U = 570 W/m2.°C

A = 66.17369 m2

Data tubes:

outside diameter: 16 mm
inside diameter: 12 mm
length: 4.88 m
material: Carbon steel

Area one tube = 0.245171 m2
Number of tubes = 269.9081 270

Square pitch: 1.25

Bundle diameter = 405.7464 mm

head type: Split-ring floating head

bundle diametrical clearance = 56 mm

Shell diameter = 461.7464 mm

0.461746 m

Tube cross-sectional area = 0.000036 m2

Tubes per pass = 270

Total flow area = 0.00972 m2

Volumetric flow = 0.013376 m3/s

tube-side velocity = 1.376128 m/s

Tube-side heat transfer coefficient:

Re = 6.77E+04

Pr = 3.823256

L/di = 406.6667

using figure 12.23:
Jh = 3.00E-03

Nu = 3.16E+02

hi = 2390.14 W/m2.K

Shell-side coefficient:

Mean temperature = 307.15 K

Baffle: 100 percentage of shell diameter

Baffle spacing = 461.7464 mm

Tube pitch = 20 mm

Cross-flow area = 42641.95 mm2
0.042642 m2

De = 15.55 mm

Volumetric flow-rate = 0.018508 m3/s

Shell-side velocity = 0.434033 m/s

Re = 8.39E+03

Pr = 5.69E+00

baffle cut 25%

jh 6.10E-03 fig 12.29

hs = 3.45E+03 W/m2.K

Overall coefficient:

thermal conductivity metal = 17 W/m2.°C

fouling coefficients (product): 10 kW/m2.°C
fouling coefficients (water): 2 kW/m2.°C

1/U0 = 0.00175

U0 = 571.4952 W/m2.°C

Pressure drop:

Tube-side:

number of tubes = 270
passes = 1
I.d. = 12 mm
O.d = 16 mm
ut = 1.376128 m/s
Re = 6.77E+04

jf = 0.003 fig. 12.24

dP = 6124.275 N/m2
0.061243 bar

Shell-side:

Gs = 165.4873 kg/s.m2

Linear velocity = 0.313681 m/s

Re = 2.00E+04

jf = 4.20E-02 fig. 12.30

dP = 2736.852 N/m2
0.027369 bar

Costing:

Material		Carbon steel		
Standard price:	hfl	34000	per	50 m2
Price	hfl	44998.11		
pressure factor:		1.1		
Price:	hfl	49497.92		

Design Heat-exchanger C5+-product

Data streams:

		inlet	mean	outlet	
C5-stream:	temperature	400.8481	355.4991	310.15	K
	thermal conductivity	0.081	0.09635	0.1117	W/m.K
	density	519.3174	572.775	626.2325	kg/m3
	viscosity	0.00013	0.000182	0.000234	Pa.s
	flow	5.7416	5.7416	5.7416	kg/s
	avg molweight		79.6805		g/mol
	specific heat	2.934019	2.605722	2.277424	kJ/kg.K

Cooling water:	- Cp:	4.2	kJ/kg.K
	- Tin:	303.15	K
	- Tuit:	311.15	K
	- density	995	kg/m3
	- viscosity	8.00E-04	Ns/m2
	- thermal conductivity	5.90E-01	W/m.°C

Calculation:

type: counter-current

Heat load: 1338.3 kW

Cooling water flow: 39.83036 kg/s

dTlm: -32.42377 °C

U = 550 W/m2.°C

A = 75.04596 m2

Data tubes:

outside diameter: 16 mm
inside diameter: 12 mm
length: 6.1 m
material: Carbon steel

Area one tube = 0.306464 m2
Number of tubes = 244.8769 245

Square pitch: 1.25

Bundle diameter = 388.2708 mm

head type: Split-ring floating head

bundle diametrical clearance = 55 mm

Shell diameter = 443.2708 mm
0.443271 m

Tube cross-sectional area = 0.000036 m2

Tubes per pass = 245

Total flow area = 0.00882 m2

Volumetric flow = 0.010024 m3/s

tube-side velocity = 1.136529 m/s

Tube-side heat transfer coefficient:

Re = 4.30E+04

Pr = 4.917201

L/di = 508.3333

using figure 12.23:
Jh = 3.20E-03

Nu = 2.33E+02

hi = 1867.202 W/m2.K

Shell-side coefficient:

Mean temperature = 307.15 K

Baffle: 100 percentage of shell diameter

Baffle spacing = 443.2708 mm

Tube pitch = 20 mm

Cross-flow area = 39297.81 mm2
0.039298 m2

De = 15.55 mm

Volumetric flow-rate = 0.040031 m3/s

Shell-side velocity = 1.018645 m/s

Re 1.97E+04

Pr 5.69E+00

baffle cut: 25%

jh 4.40E-03 fig. 12.29

hs = 5.84E+03 W/m2.K

Overall coefficient:

thermal conductivity metal = 17 W/m2.°C

fouling coefficient product: 10 kW/m2.°C
fouling coefficient water: 2 kW/m2.°C

1/U0 = 0.001787

U0 = 559.4792 W/m2.°C

Pressure drop:

Tube-side:

number of tubes = 245
passes = 1
I.d. = 12 mm
O.d = 16 mm
ut = 1.136529 m/s
Re = 4.30E+04

jf = 0.0033 fig. 12.24

dP = 5889.22 N/m2
0.058892 bar

Shell-side:

Gs = 146.1048 kg/s.m2

Linear velocity = 0.255082 m/s

Re = 1.25E+04

jf = 4.80E-02 fig. 12.30

dP = 2807.023 N/m2
0.02807 bar

Costing:

Material:		Carbon steel		
Standard price:	hfl	44000	per	70 m3
Price:	hfl	47171.74		
prssure factor:		1.1		
Price:	hfl	51888.92		

Appendix 8-6: Design of pumps

Pump:		P 01	P 02	P 03	P 04	P 05	P 06
Name:		Bottoms V-02	Bottoms V-03	Bottoms C-01	Reflux C-03	C3 Pump	Reflux C-04
feed nr.:		11	15	21	50	52	59
Temperature:	(K)	207.5000	220.0000	208.7	318.8	318.8	343
Density:	(kg/m3)	127.0600	363.6400	348.4	451.97	451.97	504.92
Viscosity:	(Pa.s)	7.52E-05	6.67E-05	6.73E-05	7.90E-05	7.91E-05	1.09E-04
Vapour pressure:	(bara)	14.2809	13.7657	21.72	16	16	9
at temperature:	(K)	207.5000	220.0000	208.7	318.8000	318.8000	343.0000

Power

mass flow:	(kg/s)	19.82	20.6	5.2	22.47	11.3	20.75	
Capacity	(m3/s)	1.56E-01	5.66E-02	1.49E-02	4.97E-02	2.50E-02	4.11E-02	
	(m3/h)	561.56	203.94	53.73	178.98	90.01	147.94	
Suction pressure:	(bara)	30	29.7	30.2	16	16	9	
elevation height:	(m)	14.3	21.6	16.9	33.5	0	24	
density:	(kg/m3)	127.06	363.64	348.4	451.97	451.97	504.92	
pressure height:	(bar)	0.17824358	0.770538614	0.577608876	1.48533161	0	1.188783648	
pressure drop:	(bar)	0.3	0.3	0	0	0	0	
pressure rise:	(bar)	0.15	0	0.15	0	5	0	
pressure "drop":	(bar)	0	0	0.2	0	0	0	
Discharge pressure:	(bara)	30.62824358	30.77053861	30.72760888	17.48533161	21	10.18878365	
Theoretical power:	(kW)	9.799927398	6.064540605	0.787475934	7.38442845	12.5008297	4.88538	
Pump efficiency:	(-)	83	80	75	80	76	78	fig. 10.62
Power at shaft:	(kW)	11.80714144	7.580675756	1.049967912	9.230535563	16.44846013	6.263307692	

		P 01 Bottoms V-02	P 02 Bottoms V-03	P 03 Bottoms C-01	P 04 Reflux C-03	P 05 C3 Pump	P 06 Reflux C-04
vapour pressure:	(bara)	14.2809	13.7657	21.7200	16.0000	16.0000	9.0000
pressure before pump:	(bara)	30	29.7	30.2	16	16	9
liquid height before pump:	(m)	2.35	7.9	1.5	3	3	3
NPSH available	(m)	1263.45085	454.5757065	249.6125307	3	3	3

Costing:

elevation hight:	(m)	50.4022821	30.00970184	15.43707234	33.5	112.7694307	24
Capacity:	(m3/h)	561.56	203.94	53.73	178.98	90.01	147.94
Power:	(kW)	11.81	7.58	1.05	9.23	16.45	6.26
RPM:		1450	1450	1450	1450	2900	1450
Price:	**(hfl)**	**62000**	**20500**	**14400**	**20500**	**15000**	**20500**
Power:	(kW)	12.5	10	1.3	11	18.5	7.5
Price electromotor:	(hfl)	2785	2165	505	3700	5570	3090
Price (total)	**(hfl)**	**64785**	**22665**	**14905**	**24200**	**20570**	**23590**

* price 315m3/h taken twice

Appendix 8-7: Design of liquid-vapour separators

For the design of the liquid-vapour separators [45], p 144 is used to determine the maximum allowable throughput $U_{g,max}$.

$$U_{g,\max} = \lambda_{\max} \cdot \left(\frac{\rho_l - \rho_g}{\rho_g} \right)^{0.5}$$

with:

$U_{g,max}$ = maximum superficial gas velocity at the entrance of the separator package (m/s)

λ_{max} = maximum gas-load factor (m/s)

ρ_l = liquid density (kg/m^3)

ρ_g = gas density (kg/m^3)

The gas load factor, λ, incorporates the influence of gas density on the maximum allowable gas velocity and is, for vane-type demisters, a function of surface tension, liquid density and geometry. The following expression is used for the gas load factor ([45], p.146)

$$\lambda_{\max} = C \cdot \left(\frac{\sigma \cdot g}{\rho_l} \right)^{0.25}$$

with:

C = parameter for different types of vane demisters (-)

g = gravitational acceleration (m/s^2)

σ = surface tension (kg/s^2)

In [45], p 145-147 three vane-type demisters are described. A vane-type II demister is used, because at natural gas treating conditions it can separate smaller droplets than the other types ([45], p.148). For vane-type II demister C is 1.20.

Using these equations the $U_{g,max}$ can be calculated.

Assuming that the separator package has the same diameter as the column, $U_{g,max}$ is the maximum allowable gas velocity in the column.

The minimum vessel cross-sectional area (m^2) is calculated with equation xx:

$$A_{\min} = \frac{Q_v}{U_{g,\max}}$$

with:

Q_v = vapour flow rate (m^3/s)

The minimum diameter is calculated using:

$$D_{\min} = \sqrt{\frac{4 \cdot A_{\min}}{\pi}}$$

The column diameter is chosen equal to the minimum diameter.

For a vertical drum the vapor disengaging height is chosen 1.5. ([44], p. 156)

For the liquid-vapour separators a liquid hold-up time of 5 minutes is chosen [46]. The liquid height is calculated using equation xxx.

$$H_{liquid} = \frac{Q_l \cdot t_h}{A_{\min}}$$

with:

H_{liquid} = liquid height (m)
Q_l = liquid flow rate (m^3/s)
t_h = liquid hold-up time (s)

The total height of the column is equal to the sum of liquid height and vapour disengaging space.

The calculations for the process can be found on the following page.

Calculation vane-type separators

		V01	V02	V03	
surface tension	(kg/s2)	0.0027	0.0035	0.0033	
C		1.2	1.2	1.2	
Lapda max	(m/s)	0.112794	0.118078	0.116562	
Ug,max	(m/s)	0.157708	0.331096	0.311543	
Ug	(m/s)	0.157708	0.331096	0.2	Vlager l.v.m. diameter
Qv	(m3/s)	0.996636	2.289668	0.429376	
A	(m2)	6.319513	6.91542	2.146879	
D	(m)	2.837311	2.968072	1.653746	
Disengaging height	(m)	4.255967	4.452108	2.48062	
liquid vol. Flow	(m3/s)	0.08222	0.054124	0.056637	
Vol for 5min. Hold-up	(m3)	24.66612	16.23729	16.99125	
Liq. Height	(m)	3.903167	2.347983	7.914395	
Total height	(m)	8.159134	6.800091	10.39501	
Qsi	(-)	5.5	5.5	5.5	
Pressure drop	(Pa)	7.85415	12.45833	4.911929	
	(Pa)	11.87177	14.04283	13.58708	
Costing:					
Material:		Carbon Ste	Carbon Ste	Carbon Steel	
L	(m)	8.159134	6.800091	10.39501	
D	(m)	2.837311	2.968072	1.653746	formule 5.5, 1e deel
					formule 5.5, 2e deel
V	(m3)	51.56176	47.02549	22.31683	
wall thickness	(mm)	13	13	13	
V	(m3)	60	50	25	
price:	(hfl)	83000	76500	71400	
Isolation:	type:	pur/pir	foamglas	pur/pir	
	Area:	91	80	48	
					DACE-prijzenboekje
	price (hfl/m2)	252	490	292	
	Price isolation: (hfl)	22932	39200	14016	
Total costs		105932	115700	85416	m2

Appendix 8-8: Calculation of column diameters and efficiencies

Table 8.8.1: massflows absorber (kg/s)

Topsection		Bottomsection	
liquid	gas	Liquid	gas
9.15	118.91	5.22	115.17

Table 8.8.2: calculation of column diameter of absorber

	Tray (theor.)	Mean density		σ [N/m]	F_{lg} [-]	T_s [m]	C_{tr}* [m/s]	C_{tr} [m/s]	$U_{g,\ max}$ [m/s]	D, column [m]
		Liquid [kg/m3]	Gas [kg/m3]							
Top	1	317.75	44.3527	0.002	0.029	0.45	0.085	0.054	0.133	5.7
Bottom	10	348.43	42.3264	0.003	0.016	0.45	0.086	0.059	0.159	5.2

Table 8.8.3: massflows deethaniser (kg/s)

Topsection		Feed top		Tray 11		Feed middle		Feed bottom		Bottomsection	
liquid	gas	liquid	gas	liquid	gas	liquid	gas	liquid	gas	Liquid	gas
11.66	40.67	16.15	39.30	15.07	39.52	32.39	38.87	55.87	37.23	99.15	76.81

Table 8.8.4: calculation of column diameter of deethaniser

	Tray (theor.)	Mean density		σ [N/m]	F_{lg} [-]	T_s [m]	C_{tr}* [m/s]	C_{tr} [m/s]	$U_{g,\ max}$ [m/s]	D, column [m]
		Liquid [kg/m3]	Gas [kg/m3]							
Top	2	398	47.2	0.004	0.287	0.45	0.060	0.044	0.119	3.4
Feed top	8	417	45.1	0.005	0.411	0.45	0.051	0.039	0.111	3.5
Tray	11	428	44.6	0.005	0.381	0.450	0.053	0.040	0.118	3.5
Feed m	12	438	43.6	0.005	0.833	0.45	0.033	0.025	0.074	4.4
Feed bottom	16	446	44.4	0.005	1.500	0.45	0.045	0.034	0.103	3.6
Bottom	27	352	77.9	0.001	1.291	0.45	0.036	0.020	0.037	6.5

Table 8.8.5: massflows depropaniser (kg/s)

Topsection		Bottomsection		Feedstage	
liquid	gas	liquid	gas	liquid	Gas
18.88	30.14	34.77	22.38	30.07	29.36

Table 8.8.6: calculation of column diameter of depropaniser

	Tray (theor.)	Mean density		σ [N/m]	F_{lg} [-]	T_s [m]	C_{tr}* [m/s]	C_{tr} [m/s]	$U_{g,\ max}$ [m/s]	D, column [m]
		Liquid [kg/m3]	Gas [kg/m3]							
Top	2	486	22.7	0.007	0.135	0.45	0.074	0.060	0.270	2.8
Bottom	44	501	26.4	0.006	0.357	0.45	0.055	0.043	0.182	2.7
Feed	18	506	22.9	0.007	0.218	0.45	0.066	0.054	0.246	2.9

Table 8.8.7: massflows debutaniser (kg/s)

Topsection		Bottomsection		Feedstage	
liquid	gas	liquid	Gas	liquid	gas
12.60	19.71	23.99	19.11	26.14	21.05

Table 8.8.8: calculation of column diameter of debutaniser

	Tray (theor.)	Mean density		σ [N/m]	F_{lg} [-]	T_s [m]	C_{tr}* [m/s]	C_{tr} [m/s]	$U_{g,\ max}$ [m/s]	D, column [m]
		Liquid [kg/m3]	Gas [kg/m3]							
Top	2	505	22.5	0.006	0.135	0.45	0.074	0.058	0.269	2.3
Bottom	30	520	25.1	0.006	0.276	0.45	0.061	0.048	0.213	2.4
Feed	13	515	22.5	0.006	0.260	0.45	0.062	0.049	0.229	2.5

Table 8.8.9: calculation of column efficiency [1] [36, p. 504]

Column	Tray	α_a	μ_a	E_0
Deethaniser	2	3.50	0.1	
	14	2.96	0.1	
	28	1.66	0.1	
	mean	2.70	0.1	69
Depropaniser	2	2.31	0.1	
	23	2.08	0.1	
	45	1.82	0.1	
	mean	2.07	0.1	73
Debutaniser	2	1.93	0.1	
	16	1.75	0.1	
	31	1.54	0.1	
	mean	1.74	0.1	76

1) Values taken from Chemcad

Table 8.8.10: calculation of absorber efficiency [36, p. 504+505] [1]

μ_s (mNs/m^2)	ρ_s (kg/m^3)	K (-)	M_s kg/kmol	x	E_0 (%)
0.07	348.4	5.13	27.28	2.30	43

1) The values are taken from Chemcad and refer to the bottomstream of the absorber, which has the highest viscosity (viscosity has a depreciating effect on the efficiency; therefore this is a conservative estimation)

Appendix 8-9: Design of condensers

Design condenser de-propaniser

Utilities:			
Water:	- Cp	4.18	kJ/kg
	- Tdesign	30	°C
	- Tmax	38	°C
	- density	993	kg/m3

Data pipe:		
- outside diameter	16	mm
- inside diameter	12	mm
- length	4.88	m

Tuit =	318.75	K	45.60
dTmin =	4.00	°C	
Twater, max =	38.00	°C	41.60
Average mol. Weight	43.7929	g/mol	
Heat transferred from vapour:	9942.2	kW	3314.067
Condensation rate:	33.77004	kg/s	294.4089
Cooling water flow =	297.3146	kg/s	
Assumed overall coefficient:	600	W/m2.°C	
dTmn =	11.12467	°C	
R =	0		
S =	0.512821		
shell:	1		
tubes:	2		
Ft =	1		
dTm =	11.12467		
Trial area =	1489.513	m2	
Max. Area per unit:	600	m2	
Number of units:	3		
Trial area per unit:	496.5043	m2	
Surface area of one tube =	0.245171	m2	
Number of tubes =	2025.133		
	2030		

Pitch: square pitch
1.25

Pt = 20 mm

Tubes bundle diameter = 1047 mm

Number of tubes in centre row = 52.32993
53

Shell-side coefficient:

Condensing coefficient: 1500 W/m2.°C

Mean temperature:
- Shell-side 45.60 °C
- Tube-side 34 °C

Tw = 38.64 °C

Mean condensate temperature = 36.32 °C

Physical properties a	36.32 °C	
viscosity:	8.88E-05	Pa.s
density liquid:	467.97	kg/m3
Cp:	3.00	kJ/kg.°C
thermal conductivity:	0.0845	W/m2.oC

rho,v = 52.16 kg/m3

1.00E-02 kg/s.m

Nr = 35.33333

Re = 1.49E+04
Pr = 3.15
figure 12.43: 0.31

Hc = 1632.71 W/m2.°C

Tube-side coefficient:

Tube cross-sectional area = 0.115 m2

Tube velocity = 0.87 m/s

Hi = 5169.68 W/m2.°C

Overall coefficient:

Fouling factor: 10000 W/m2.°C

Fouling factor (water): 2000 W/m2.°C

Kw: 17 W/m.°C

1/U = 0.001639

U = 610.088 W/m2.°C

Shell-side pressure drop:

type head: pull-through floating head

baffle spacing: shell diameter
baffle cut: 25%

Clearance 96 mm fig. 12.10

Shell i.d. 1143 mm

cross-flow area = 0.261106 m2

Mass flow-rate = 43.11148 kg/s.m2

equivalent diameter = 15.7988 mm

vapour viscosity = 2.00E-05 Pa.s

Re = 3.41E+04

jf 4.00E-02 fig. 12.30

Us = 0.826593 m/s

assumption: pressure drop is 50 per cent of that calculated using the inlet flow

dP = 880.5831 N/m2
8.81E-03 bar

Tube-side pressure drop:

viscosity water = 6.00E-04 Pa.s

Re = 1.73E+04

jf = 4.20E-03 fig. 12.24

dP = 12144.88 N/m2
0.121449 bar

Costing:

Material:	Carbon steel		
Standard price:	175000	per	500 m2
Price:	173776.5		
Number of units:	3		
Price:	521329.5		
pressure factor:	1.2		
Price (total)	625595.4		

Design condenser debutaniser

Utilities:			
Water:	- Cp	4.18	kJ/kg
	- Tdesign	30	°C
	- Tmax	38	°C
	- density	993	kg/m3

Data pipe:		
- outside diameter	25	mm
- inside diameter	21.8	mm
- length	4.88	m

Tout = 343.01 K 69.86

Average mol. Weight 79.67 g/mol

Heat transferred from vapour: 6176.3 kW 305.1847

Condensation rate: 20.23790993 kg/s

Cooling water flow = 184.6979665 kg/s

Assumed overall coefficient: 575 W/m2.°C

dTmn = 35.71077676 °C

R = 0
S = 0.200702459

shell: 1
tubes: 2

Ft = 1

dTm = 35.71077676 °C

Trial area = 300.7885092 m2

Surface area of one tube = 0.38308 m2

Number of tubes = 785.1845807
786

Pitch: square pitch
1.25

Pt = 31.25 mm

Tubes bundle diameter = 1075 mm

Number of tubes in centre row = 34.40792335
35

Shell-side coefficient:

Condensing coefficient: 1500 W/m2.°C

Mean temperature:
- Shell-side 69.86 °C
- Tube-side 34 °C

Tw = 47.75 °C

Mean condensate temperature = 40.87 °C

Physical properties at: 40.87 °C

viscosity: 1.43E-04 Pa.s
density liquid: 545.26 kg/m3
Cp: 2.55 kJ/kg.°C
thermal conductivity: 0.096 W/m2.oC

rho,v = 94.88 kg/m3

4.82E-02 kg/s.m

Nr = 23.33333333

Re = 9.17E+03
Pr = 3.81
fig. 12.43 0.26

Hc = 1223.08 W/m2.°C

Tube-side coefficient:

Tube cross-sectional area = 0.147 m2

Tube velocity = 1.27 m/s

Hi = 7535.609872 W/m2.°C

Overall coefficient:

Fouling factor: 10000 W/m2.°C
Fouling factor water: 2000 W/m2.°C

Kw: 17 W/m.°C

1/U = 0.001685179

U = 593.4086342 W/m2.°C

Shell-side pressure drop:

type head: pull-through floating head

baffle spacing: shell diameter
25 per cent cut

Clearance 96 mm fig. 12.10

Shell i.d. 1171 mm

cross-flow area = 0.27436419 m2

Mass flow-rate = 73.76294227 kg/s.m2

equivalent diameter = 24.685625 mm

vapour viscosity = 2.00E-05 Pa.s

Re = 9.10E+04

jf 3.30E-02 fig. 12.30

Us = 0.777402453 m/s

assumption: pressure drop is 50 per cent of that calculated using the inlet flow

dP = 748.1759924 N/m2
7.48E-03 bar

Tube-side pressure drop:

viscosity water = 6.00E-04 Pa.s

Re = 4.58E+04

jf = 3.20E-03 fig. 12.24

dP = 13154.01441 N/m2
0.131540144 bar

Costing:

Material: Carbon steel
Standard price: hfl 120000 per 300 m2.
Price: hfl 120315.4

pressure factor: 1.1

price: hfl 132346.9

Appendix 8-10: Design of reboilers

Design reboiler de-ethaniser

Type of reboiler: Thermosyphon Vertical

Max. heat-flux = 37900 W/m2
Vaporisation rate = 74.9494 kg/s
Tbottom 377.8683 K

Physical properties:
Boiling point at 30 bar 377.8683 K
Tc 411.7318 K

Steam saturation temp. 150 oC

total heatload = 16706.7 kW (chemcad)

Mean temp difference = 45.2817

Reduced temperature = 0.92

design heat flux = 60000 W/m2 figure 12.59

Chosen heat flux 37900 W/m2

U: 836.9827 W/m2.°C

Required area = 440.81 m2

Data tubes:

length 3.66 m
inside diameter 25 mm
outside diameter 30 mm

Area of one tube = 0.28731 m2
Number of tubes = 1534.266

pitch: 1.25 square pitch
tubes: fixed tubes

diameter of bundle = 1671.627 mm

diametrical clearance = 25 mm figure 12.10

Shell inside diameter = 1696.627 mm

Costing:

Material: Carbon steel
Standard price: hfl 120000 per 300 m2
price: hfl 176324

pressure factor: 1.6

price: hfl 282118.4

Design reboiler de-propaniser

Type of reboiler: Thermosyphon

Max. heat-flux = 37900 W/m2

Vaporisation rate = 25.5748 kg/s
Tbottom 389.7 K

Physical properties:
Boiling point at 30 bar 389.7 K
Tc 450.8727 K

Steam saturation temp. 150 °C

total heatload = 8087 kW (chemcad)

Mean temp difference = 33.45 °C

Reduced temperature = 0.864324

design heat flux = 60000 W/m2 figure 12.59

chosen heat flux: 37900 W/m2

U = 1133.034

Required area = 213.3773 m2

Data tubes:		
length	3.66	m
inside diameter	25	mm
outside diameter	30	mm

Area of one tube = 0.28731 m2
Number of tubes = 742.6728

pitch: 1.25 square pitch
tubes: fixed tubes

diameter of bundle = 1203.275 mm

diametrical clearance = 17 mm figure 12.10

Shell inside diameter = 1220.275 mm

Costing:

Material:		Carbon steel		
Standard price:	hfl	90000	per	200 m2
price:	hfl	96019.79		
pressure factor:		1.2		
price:	hfl	115223.7		

Appendix 8-10

Design reboiler de-butaniser

Type of reboiler: Thermosyphon Vertical

Max. heat-flux = 37900 W/m2

Vaporisation rate = 19.4728 kg/s
Tbottom 400.81 K

Physical properties:	
Boiling point at 30 bar	400.81 K
Tc	450.8727 K

Steam saturation temp. 150 oC

total heatload = 5372.8 kW (chemcad)

Mean temp difference = 22.34

Reduced temperature = 0.888965

design heat flux = 42000 W/m2 figure 12.59

chosen heat flux 37900 W/m2

U = 1696.509

Required area = 141.7625 m2

Data tubes:	
length	3.66 m
inside diameter	25 mm
outside diameter	30 mm

Area of one tube = 0.28731 m2
Number of tubes = 493.4132

pitch: 1.25 square pitch
tubes: fixed tubes

diameter of bundle = 999.7693 mm

diametrical clearance = 17 mm figure 12.10

Shell inside diameter = 1016.769 mm

Material:		Carbon steel		
Standard price:	hfl	48000	per	100 m2
Price:	hfl	68046.02		
Pressure factor:		1.1		
Price:	hfl	74850.62		

Appendix 8.11: Liquid accumulator design

Units		V 04	V 05
"Feed"	nr	<48>	<58>
	(kg/s)	33.77	20.75
avg. density	(kg/m3)	451.9672	504.9159
Flow	(m3/s)	0.074718	0.041096
Hold-up time	(s)	300	300
Hold-up volume	(m3)	22.41534	12.32879
Liquid volume fraction		0.65	0.65
Volume	(m3)	34.48514	18.96736
Chosen vessel:			
V	(m3)	35.00	20.00
L	(m)	4.6	4.5
D	(m)	3	2.3
gem. Plaatdikte	(mm)	13	10
Materiaal		Steel	Steel
Price:	(hfl)	70000	56000

Appendix 8-12: Calculation of properties of product streams

	Sales gas, 32				C3, <54>			
Component	Flow	molfraction	Pvp	yi/Pvp	Flow	molfraction	Pvp	yi/Pvp
	[kmol/s]	[-]	[bar]	[1/bar]	[kmol/s]	[-]	[bar]	[1/bar]
N2	0.027594	0.00428	0.214017	0.02	0	0	-	
CH4	5.61793	0.871447	43.57237	0.02	0	0	-	
C2H6	0.563954	0.08748	0.854328	0.102396	0.003698	0.014388	-	
CO2	0.232104	0.036004	0.806651	0.044634	0	0	-	
H2S	1.76E-05	2.73E-06			0	0		
C3H8	0.005066	0.000786	0.056503	0.013907	0.251884	0.979983	12.79219	12.53612
I-C4H10	0	0	0.008405	0	0.001304	0.005074	4.876376	0.024743
N-C4H10	0	0	0.003718	0	0.000143	0.000556	3.487068	0.001937
I-C5H12	0	0	0.000356	0	0	0	1.124143	0
N-C5H12	0	0	0.00044	0	0	0	1.284774	0
N-C6H14	0	0	1.89E-05	0	0	0	0.33122	0
N-C7H16	0	0	1.27E-06	0	0	0	0.107512	0
H2O	0	0	1.55E-07	0	0	0	0.062836	0
N-C8H18	0	0	7.28E-08	0	0	0	0.035566	0
Total	6.446666	1			0.257029	1		

Results:

sum-inerts: 0.160937 -
Pcorr= 6.213612 bar
P-1/sum(yi, 1.63E-08 -
Dew,T 181.3587 K
Dew,T -91.79132 °C

P-c2h6 12.5628 bar
Pbubl 12.74619 bar

	C4, <61>				C5+, <64>			
Component	Flow	molfraction	Pvp	yi/Pvp	Flow	molfraction	Pvp	yi/Pvp
	[kmol/s]	[-]	[bar]	[1/bar]	[kmol/s]	[-]	[bar]	[1/bar]
N2	0	0	-		0	0	-	
CH4	0	0	-		0	0	-	
C2H6	0	0	-		0	0	-	
CO2	0	0	-		0	0	-	
H2S	0	0			0	0		
C3H8	0.000252	0.002089	12.79219	0.026719	0	0	12.79219	0
I-C4H10	0.042127	0.349577	4.876376	1.704667	1.55E-05	0.000215	4.876376	0.001048
N-C4H10	0.076851	0.63773	3.487068	2.223809	0.000933	0.012941	3.487068	0.045127
I-C5H12	0.000897	0.007441	1.124143	0.008365	0.021512	0.298531	1.124143	0.335592
N-C5H12	0.000381	0.003163	1.284774	0.004064	0.022371	0.310457	1.284774	0.398866
N-C6H14	0	0	0.33122	0	0.016889	0.234375	0.33122	0.07763
N-C7H16	0	0	0.107512	0	0.008271	0.114786	0.107512	0.012341
H2O	0	0	0.062836	0	0	0	0.062836	0
N-C8H18	0	0	0.035566	0	0.002068	0.028695	0.035566	0.001021
Total	0.120508	1			0.07206	1		

Results:

Pbubl 3.967624 bar

Pbubl 0.871625

Design Basis

Gas Treatment Plant

Raytheon E&C

Raytheon Project No. 00231

ISSUE	DATE	DESCRIPTION	BY	CH'K'D	APP'D	CLIENT
1	13-Jan-99	For comments	PR	-	-	Raytheon E&C

ID. NO.		SHEET	DOC.NO.
C:\CPD01.DOC\I	Raytheon Engineers & Constructors	1 of 4	231-10-basis

Table of Contents

1. Introduction

The purpose of the 'Conceptual Process Design' is to find an optimum design for the recovery of NGL from feedgas originating from a Northsee gasfield.

The minimum task for the new facility is to treat the feed gas to meet dewpoint and other sales gas quality specifications and then to compress the treated gas into an export pipeline. Another task is the extraction of NGL which involves optimizing the recovery of propane and heavier hydrocarbons.

2. Feed and Product Specifications

2.1 Feed Gas

Composition of Feed Gas:

	mol%
N2	0.400
CO2	3.365
C1	81.430
C2	8.230
C3	3.730
iC4	0.630
nC4	1.130
iC5	0.325
nC5	0.330
nC6	0.245
nC7	0.120
nC8	0.030
H2S	2.5E-4
H2O	0.035

Feed Rate: 5.00 Gsm^3/y (based on 8400 operating hours per year)

Turndown: 50%

Feed Conditions:
Temperature: 5 °C
Pressure: 11000 kPa

2.2 Sales Gas Specification

Hydrocarbon dew point: -15°C at 5000 kPa
Water dewpoint: <0.030 g/Sm^3
Export pipeline inlet pressure: 19000 kPa
Export pipeline max. temperature: 50°C

2.3 NGL Export Specification

C_2/C_3 ratio:	<1.5 wt%
C_3 recovery:	>80%, maximize NGL recovery (no economic parameters available, validate additional investment for higher recovery by 'good engineering judgement').
NGL product temperature:	35°C
NGL Battery Limit pressure:	> 500 kPa above bubble point

3. Utility Conditions

LP Steam:	
Thermal design conditions:	150°C, saturated
MP Steam:	
Thermal design conditions:	175°C, saturated
Cooling Water:	
Supply temperature:	30°C
Maximum return temperature:	38°C
Air (for aircoolers):	
For thermal design:	30°C

4. Technical Options

- Consider applying a wash section in the deethanizer top section. Use heavy natural gasoline components, present in the feed gas, as heavy wash oil (an additional distillation column will be required).
- Consider using plate-fin heat exchangers to reduce equipment size and cost.
- Consider heat integration by using an interstage reboiler.
- Consider using vane packages in vapor-liquid separators to reduce size.
- Consider the effect of an expander trip
- Study limitations in expander operation with respect to retrograde condensation, allowable liquid fractions in expander outlet (Raytheon is interested in expander design criteria, however, by lack of a proper reference use a maximum liquid fraction of 25 wt%).

5. Some Patent Literature

EP 0 182 643 (19-11-84): Process and apparatus for separating C3 and heavier components from hydrocarbon gases, The Ortloff Corporation.

EP 0 240 188 (01-04-86): Process for separating hydrocarbon constituents, McDermott International Inc.

GB 2 102 931A (30-6-82): Recovery of condensable hydrocarbons from gaseous streams, Snamprogetti S.p.A.

GB 1 532 334 (9-8-76): Hydrocarbon gas processing, The Ortloff Corporation

GB 1 532 335 (9-8-76): Hydrocarbon gas processing, The Ortloff Corporation

US 4 171 964 (4-10-76): Hydrocarbon gas processing, The Ortloff Corporation

Appendix 9-2: Relevant data given in correspondence with Raytheon

Taken from email of 22 January 1999

[1] De volgende specificaties gelden voor C3-LPG, C4-LPG, en C5+:

C3-LPG: C2 < 1.5 wt%
C4 < 1.0 wt%
H2S < 40 ppmwt
geen verdere specificaties
C3-LPG product condities naar opslag: Temp < 38 °C
Druk: 20 barg

C4-LPG: C3 < 0.2 wt%
C5+ < 1.5 wt%
H2S < 15 ppmwt
geen verdere specificaties
C4-LPG product condities naar opslag: Temp < 38 °C
Druk: 8.0 barg

C5+" C4 < 1.0 wt%
C5+ product condities naar opslag: Temp < 38 °C
Druk: 7.0 barg

Voor het sales gas zijn geen nadere specificaties (N2, CO2, H2S).

Taken from email of 2 February 1999

Wat betreft jullie vragen, ik heb niet overal een antwoord op:

[1] Locatie: Kies maar een plaats aan de Noorse kust.
[2] Economische levensduur: 25 jaar
[3] Prijzen: Geen idee, indien jullie niets vinden neem als basis:
Sales Gas: Nederlandse export prijs van aardgas
Feed Gas : Sales Gas x 75%
[4] C4 en iC4 vormen samen het C4-LPG produkt, scheiding niet nodig. C5+ is een tussenprodukt en hoeft niet verder gescheiden te worden in jullie ontwerp.
[5] Water en CO2 moeten gescheiden worden om bevriezing te voorkomen. Het voorbeeld dat ik heb gepresenteerd omvatte alleen waterscheiding, terwijl het CO2 gehalte toch 3.3 mol% CO2 was.

Taken from email of 26 February 1999

Hier enkele aandachtspunten:

* Turbo-expanders: Als maximum voor een grote turbo-expander zou ik een motor grootte van 7500 kW aanhouden en een aanzuigvolume van ongeveer 10000 m3/h. Om binnen deze limiet te blijven kan het helpen om de voeding naar de turbo-expander wat af te koelen. De adiabatische efficientie is +/- 85%.

* Wat drijft de Turboexpander aan? Zorg voor een 'match' van duties.

* Denk aan vloeistoffractie limitaties in turboexpanders.

* Warmte-integratie: Bij dit soort energiebehoeften is warmte-integratie erg belangrijk. Dit valt uit de flowsheets niet op te maken. B.V. hoe wordt in het Reflux HE process de koude van warmtewisselaar no. 16 benut? Ook zie ik grote verschillen wat betreft de mogelijkheden tot warmteintegratie tussen de verschillende processen.

* Denk aan een juist drukprofiel

* Het Sales Gas moet ook nog gecomprimeerd worden. Deze compressor is het duurste stuk apparatuur in het proces. Het is daarom belangrijk wat de aanzuigdruk is. (Het Reflux HE proces lijkt een belangrijk voordeel te hebben).

* Reflux Heat Exchanger Process:
- Wat is bepalend voor de keus van het 50 bar druk niveau?
- Wat indien er direkt naar 30 bar wordt geexpandeerd?
- De C3 recovery is inderdaad laag vergeleken bij de andere twee processen, en lijkt een goede reden om het proces af te laten vallen (zie echter de vorige opmerking).
- Waarom wordt geexpandeerd in '6' en warmte toegevoegd in '19'? Hetzelfde geldt voor '1' en '17'?
- Extra koeling bij deze lage temperaturen is erg kostbaar. Indien jullie voor dit proces kiezen moet ook dit gesimuleerd worden.

* Cryomax:
- Klopt de warmtebalans rond 'unit 2'?
- Extra koeling bij deze lage temperaturen is erg kostbaar. Indien jullie voor dit proces kiezen moet ook dit gesimuleerd worden.
- Wat is de gevoeligheid van recovery en drukniveau, doen jullie nog een soort van optimalisatie?

*Turboexpander proces:
- Lijkt erg veel op het besproken voorbeeld, maar is op sommige punten afwijkend. Wat is het effect van koelen van de voeding? Kan hierdoor de eerste expander '1' komen te vervallen?
- Zijn de voedingslocaties in de Deethanizer geoptimaliseerd? Is warmteintegratie van de voedingen naar de kolom met andere stromen voordelig?

Appendix 9-3: Correspondence with Mafi-Trench

We have performed a preliminary analysis of your process conditions and determined that two Frame 5 turboexpander-compressors operating in series will handle this duty. We would recommend placing the compressor flow in series and not in parallel as noted on your data sheets. This allows a better match between the expander and compressor impellers. Performance of these units is detailed in the attached file. A budget price for the turboexpanders, auxiliary support skid (API 614 style lube oil system to supply oil to the bearings and a seal gas supply system to supply seal gas to the labyrinth seals), and control system is approximately USD 2,000,000 to 2,250,000 for a European location. This price would normally be affected by the level of specifications imposed by our client and the materials of construction required for the support system (i.e. carbon steel vs stainless steel). Normal delivery time for the package would be 52 weeks. Typical materials of construction for our turboexpander unit are detailed in the attached file. Additional comments are provided in your text below.

I hope this helps you on your project. Please let me know if you have troubles opening the files (if you do, let me know what type of graphics package you have or a fax number I can send to) or if I can answer any additional questions.

Dean Rupprecht
Applications Engineering Manager

PS. Be sure to tell your colleagues to specify Mafi-Trench Turboexpanders when they go to work for Fluor, Stork Comprimo, ABB Lummus, Gasunie, NAM, etc!

Dear Sirs,
Hereby I would like to ask for a rough cost estimation of a turbo-expander for a none-commercial project, which is carried out by four students at Delft University of Technology, The Netherlands. The project is a design case, which is compulsory for students Chemical Technology in their fourth year. It concerns the complete design and modelling of a chemical plant.
I have taken the liberty of attaching a file, which contains the involved stream compositions. It concerns two turbo-expanders used in a process retrieving C3+ hydrocarbons from a stream of natural gas. The stream names denote which turbo-expander they enter or leave. Expander 1 is linked to compressor 1, and expander 2 to compressor 2.
Furthermore, I would like to ask MTC the following questions:

- what is the maximum suction volume (sm3/h)?

Our guide vanes are designed to allow 125% of design flow.

- what is the maximum allowable liquid fraction at the of the expander?

We have experience with up to 40% by weight at expander discharge with no adverse affects. The liquid concentrations for this project are well within acceptable values.

- what material would MTC use for the manufacture these particular turbo-expanders?

See attached file.

- what are the operating areas: when does retrograde condensation occur? Is it likely to occur in these turbo-expanders?

Not sure I understand your question -

- what is the efficiency of the expander and of the compressor?

See attached data sheets

Finally, I would like to stress that I am interested in just a rough cost estimation for calculation of the total investment needed in my conceptual natural gas plant (which is, of course, only a concept and not to be implemented in any way).

I would like to thank MTC in advance for its effort and hope to hear from MTC shortly.

Yours sincerely,
Bob Tulleken

REFERENCE ONL

◂◂◂◂◂
FILE=

MAFI-TRENCH CORPORATION MACHINE CHARACTERISTICS
INQUIRY No FOR: Delft University of Technology
PREP. BY: DEV DATE: 3/29/99 TYPE/SIZE: (2) EC 5.0

(2)EC5 IN SERIES

CONDITION	DESIGN STAGE 1		DESIGN STAGE 2	
STAGE	EXP.	CMP.	EXP.	CMP.
CH4	81.46	87.14		87.14
C2H6	8.23	8.75		8.75
C3H8	3.73	.08		.08
I-C4H10	.63			
N-C4H10	1.13			
I-C5H12	.33			
N-C5H12	.33			
N-C6H14	.24			
N-C7H16	.12			
N-C8H18	.03			
N2	.40	.43		.43
CO2	3.37	3.60		3.60
H2S			VAPOR	
LB/HR	1129723	938,799	922,983	938,799
IN DENSTY	9.44	2.42	4.45	2.00
NA/REC. ?	10.00	-	16.95	-
MW	20.64	18.34	18.92	18.34
CP	.77	.41	.63	.40
Z	.72	.87	.80	.86
a	921	1204	974	1149
P1, PSIA	1,595.4	584.4	805.9	435.1
T1, °F	41.1	14.2	-18.8	-27.7
P2, PSIA	805.9	809.0	435.1	584.4
T2, °F	-18.8	63.7	-68.2	14.2
Wt% liq	18.3%	-	8.0%	-
H	19.80	15.29	19.80	12.48
ACFM	3730	6462	6257	7836
RPM	10,000	10,000	10,100	10,100
Ns	57	91	75	118
EFF,%	86.0%	75.0%	86.0%	75.0%
HP	7559	7520	6175	6136
psi	-	1.07	-	1.00
WHL OD	15.80	19.4	15.80	17.9
U2, fps	689	846	696	789
U2/Co	.69		.70	
Ma. No.	.40	.36	.43	.40
BRG SIZE	4	E,ACFM IN	4	E,ACFM IN
BRG LOSS	39	1994	39	3459
%Q/N				
%U/C opt	100%		100%	
INLET	12(47)900	24(43)600	12(82)600	24(52)600
OUTLET	18(44)900	24(34)600	18(74)600	24(42)600

MAFI-TRENCH TURBOEXPANDER-COMPRESSOR SYSTEM

STANDARD MATERIALS OF CONSTRUCTION

ITEM NO.	DESCRIPTION	MATERIAL
41	Expander housing	Cast ASTM A351 Gr.CF8 stainless steel
46	Expander follower	Stainless steel AISI 410
55	Inlet vane	Stainless steel AISI 416
56	Adjusting links	Stainless steel AISI 410
61	Pressure ring	Stainless steel AISI 410
65	Expander housing cover	Cast ASTM A352 Gr.LC3 nickel steel
71,135	Nose cone	Aluminum alloy 6061-T6 or 7075-T6
73	Center stretch rod	Stainless steel 17-4 PH
75	Expander wheel	Aluminum alloy 6061-T6 or 7075-T6
79,134	Shaft key	Stainless steel AISI 410
80	Expander wheel seal	Brass
81	Heat Barrier	Micarta Gr.221
83,124	Seal insert	Micarta Gr.221
87,126	Shaft seal ring	Stainless steel 17-4 PH
91	Shaft	Stainless steel 17-4 PH
95,114	Bearing	Brass
100	Bearing housing	Cast ASTM A216 Gr.WCB carbon steel
122	Compressor shaft seal	Aluminum alloy 6061-T6
128	Compressor wheel seal	Carbon steel
130	Compressor wheel	Aluminum alloy 6061-T6 or 7075-T6
136	Compressor follower	Aluminum alloy 6061-T6
140	Compressor housing	Cast ASTM A216 Gr.WCB carbon steel
146	Diffuser bracket	Cast or fabricated carbon steel
153	Compressor inlet spool	Cast or fabricated carbon steel
606	Expander casing studs	Stainless steel per ASTM A320
603	Compressor casing studs	Carbon steel per ASTM A193

NOTE:

These materials are typical for cryogenic gas plant and petrochemical plant turboexpanders. Alternative materials are available to meet the requirements of any application.

Appendix 10-1: Explanation of choices made in determination of F&EI

- General process hazards

a) Exothermic reactions: exothermic reactions are not present in the process. No penalty.
b) Endothermic reactions: endothermic reactions are not present in the process. No penalty.
c) Material handling & transfer:
d) Enclosed process units: all units are located in open air. Maybe some sort of roof or covering is provided, leaving all walls open. No penalty.
e) Access: due to the fact that no walls are surrounding the unit, no penalty is applied.
f) Drainage: it is assumed that the plant has an adequate drainage design to deal with spills and leakage of mainly the heavier hydrocarbons. Penalty: 0.25.

- Special process hazards for turbo-expander (K01)

a) Process temperature: the processed components are above their boiling point, but below their autoignition temperature. Penalty: 0.6
b) Low pressure (sub-atmospheric): operating pressure is high. No penalty.
c) Operation in or near flammable range: no oxygen is present. No penalty.
d) Dust explosion: no dust particles are present. No penalty.
e) Pressure: the highest operating pressure is the pressure of the feed of the expander part of the turbo-expander, which is 110 bara (1580 psig). Penalty used: 0.92.
f) Low temperature: inlet temperature of the compressor part is 240 K, which is lower than 244 K, so a penalty of 0.5 is used, because of the use of carbon steel in the compressor housing.
g) Quantity of flammable material: based on a throughput of 1 minute, which is rather large considering the fact that the volume of the turbo-expander is rather small. In calculating this quantity the throughput of gas and liquid in the expander part but also in the compressor part is used. In this case, using the heat of combustion of methane (mass bases) [53, appendix A], which is the highest of all components, and a total input of 188 kg/s this yields a penalty of 1.0.
h) Corrosion and erosion: chosen is option 1. Penalty: 0.1
j) Leakage - joints and packing: the pressure ring may leak. Also the seal gas system may fail. Also the shaft connecting the expander and compressor part can cause problems. Penalty: 0.3
k) Use of fired heaters: the fired heater and flare are more than 70 m dislocated from the turbo-expander. Penalty: 0.1.
l) Hot oil heat exchange system: not present. No penalty.
m) Rotating equipment, pumps, compressors: present. Penalty: 0.5

Explanation of choices made in determination of Fire and Explosion Index of coldbox (E01), deethaniser (C02) and compressor (K02), which deviate from those made for turbo-expander (K01)

- Special process hazards for coldbox (E01)

e) Pressure: the highest operating pressure is the pressure of output of the expander part of the turbo-expander, which is 72.5 bara (1037 psig). Penalty used: 0.86
f) Low temperature: no carbon steel is used, only aluminum. No penalty.
g) Quantity of flammable material: based on a throughput of 1 minute. In this case, using the heat of combustion of methane [53, appendix A] and a total input of 348 kg/s this yields a penalty of 1.5.
m) Rotating equipment, pumps, compressors: not present. No penalty.

- Special process hazards for deethaniser (C02)

e) Pressure: the highest operating pressure is the pressure of output of the expander part of the turbo-expander, which is 72.5 bara (1037 psig). Penalty used: 0.86

g) Quantity of flammable material: based on total mass flow through the column (64 kg/s) during 1 minute. This amount is larger than the total mass hold-up in the column (2351 kg, taken from Chemcad). Penalty: 0.4.

m) Rotating equipment, pumps, compressors: not present. No penalty.

- Special process hazards for compressor (K02)

e) Pressure: the highest operating pressure is the discharge pressure, which is 110 bara (2741 psig). Penalty used: 0.99
f) Low temperature: lowest temperature in the compressor is the feed temperature (285K). No penalty.
g) Quantity of flammable material: based on a throughput of 1 minute using a throughput of 118 kg/s. Penalty: 0.72.

Appendix 10-2: Fire and Explosion Index, calculation turbo-expander (K01)

Table 10.2.1: Fire and Explosion Index calculation turbo-expander (K01)

Exhibit A		
Material factor	21	
1. General process hazards	**Penalty**	**Penalty used**
Base factor		1.00
A. Exothermic reactions		
B. Endothermic reactions		
C. Material handling & transfer		
D. Enclosed process units		
E. Access		
F. Drainage		0.25
General process hazard factor (F1)		1.25
2. Special process hazards		
Base factor		1.00
A. Process temperature		
1. Above flash point		
2. Above boiling point		0.60
3. Above autoignition		
B. Low pressure (sub-atmospheric)		
C. Operation in or near flammable range		
1. Tank farm storage flammable liquids		
2. Process upset or purge failure		
3. Always in flammable range		
D. Dust explosion		
E. Pressure		0.92
F. Low temperature		0.50
G. Quantity of flammable material		
1. Liquids or gases in process		1.00
2. Liquids or gases in storage		
3. Combustible solids in storage		
H. Corrosion and erosion		0.10
J. Leakage – joints and packing		0.30
K. Use of fired heaters		0.10
L. Hot oil heat exchange system		
M. Rotating equipment, pumps, compressors		0.50
Special process hazard factor (F2)		5.02
Unit hazard factor (F3=F1*F2)		6.28
Damage factor		0.78
Fire and explosion index (F3 x MF)		132

Table 10.2.2: Loss Control Credit Factors Calculation

Exhibit B Loss Control Credit Factors			
1. Process control (C1)	**Credit**		**Credit**
a) Emergency power	0.97	e) Computer control	0.79
b) Cooling	0.98	f) Inert gas	1.00
c) Explosion control	0.74	g) Operating procedures	0.90
d) Emergency shutdown	0.90	h) Reactive chemical review	0.85
C1 Total	0.38		
2. Material isolation (C2)	**Credit**		**Credit**
a) Remote control valves	0.94	c) Drainage	0.95
b) Dump/blowdown	0.94	d) Interlock	0.96
C2 Total	0.81		
3. Fire protection (C3)	**Credit**		**Credit**
a) Leak detection	0.90	f) Sprinkler systems	1.00
b) Structural steel	0.92	g) Water curtains	1.00
c) Buried tanks	1.00	h) Foam	1.00
d) Water supply	0.90	j) Hand extinguishers	1.00
e) Special systems	1.00	k) Cable protection	0.90
C3 Total	0.67		
Credit factor (C1 x C2 x C3)	0.21		

Table 10.2.3: Turbo-expander (K01) analysis summary

Exhibit C	
Unit analysis summary	
Fire and explosion index	132
Radius of exposure (m) [1]	34
Value of area of exposure (Mln Dfl @ 1999) [2]	4.37
Damage factor	0.78
Base MMPD (Mln Dfl @ 1999)	3.41
Credit factor	0.21
Actual credit factor	0.40
Actual MMPD (Mln Dfl @ 1999)	1.36
Days outage (days) [3]	20
Value of product manufactured (Mln Dfl @ 1999) [4]	9.10
Total loss (Mln Dfl @ 1999) [5]	10.46

1) [53], figure 8, p 29

2) This includes a sales gas loss during 15 minutes, but that is insignifant relative to the costs of the turbo-expanders. The other turbo-expander is assumed to within the radius of Exposure. The price calculation of the turbo-expanders can be found in appendix 11-1.
   Calculation of product loss:
   The output of the two turbo-expanders is the total sales gas stream (118 kg/s).
   So: 118 kg/s * 60 s * 0.13 Dfl/kg = 14,000 Dfl @ 1999

3) From [53], figure 10; ROE = 1.94 Dfl/$. It should be noted that this a high underestimation: delivery time of turbo-expanders is about 52 weeks. Therefore spares should be bought.

4) During 20*24 h / 8400 h * 159.5 Mln Dfl

5) Value of product manufactured + Actual MMPD

Appendix 10-3: Fire and Explosion Index, calculation coldbox (E01)

Table 10.3.1: Fire and Explosion Index calculation coldbox (E01)

Exhibit A		
Material factor	21	
1. General process hazards	**Penalty**	**Penalty used**
Base factor		1.00
A. Exothermic reactions		
B. Endothermic reactions		
C. Material handling & transfer		
D. Enclosed process units		
E. Access		
F. Drainage		0.25
General process hazard factor (F1)		1.25
2. Special process hazards		
Base factor		1.00
A. Process temperature		
1. Above flash point		
2. Above boiling point		0.60
3. Above autoignition		
B. Low pressure (sub-atmospheric)		
C. Operation in or near flammable range		
1. Tank farm storage flammable liquids		
2. Process upset or purge failure		
3. Always in flammable range		
D. Dust explosion		
E. Pressure		0.86
F. Low temperature		
G. Quantity of flammable material		
1. Liquids or gases in process		1.50
2. Liquids or gases in storage		
3. Combustible solids in storage		
H. Corrosion and erosion		0.10
J. Leakage – joints and packing		0.30
K. Use of fired heaters		0.10
L. Hot oil heat exchange system		
M. Rotating equipment, pumps, compressors		
Special process hazard factor (F2)		4.46
Unit hazard factor (F3=F1*F2)		5.58
Damage factor		0.77

Appendix 10-4: Fire and Explosion Index, calculation deethaniser (C02)

Table 10.4.1: Fire and Explosion Index calculation deethaniser (C02)

Exhibit A		
Material factor	21	
1. General process hazards	**Penalty**	**Penalty used**
Base factor		1.00
A. Exothermic reactions		
B. Endothermic reactions		
C. Material handling & transfer		
D. Enclosed process units		
E. Access		
F. Drainage		0.25
General process hazard factor (F1)		1.25
2. Special process hazards		
Base factor		1.00
A. Process temperature		
1. Above flash point		
2. Above boiling point		0.60
3. Above autoignition		
B. Low pressure (sub-atmospheric)		
C. Operation in or near flammable range		
1. Tank farm storage flammable liquids		
2. Process upset or purge failure		
3. Always in flammable range		
D. Dust explosion		
E. Pressure		0.62
F. Low temperature		0.50
G. Quantity of flammable material		
1. Liquids or gases in process		0.40
2. Liquids or gases in storage		
3. Combustible solids in storage		
H. Corrosion and erosion		0.10
J. Leakage – joints and packing		0.30
K. Use of fired heaters		0.10
L. Hot oil heat exchange system		
M. Rotating equipment, pumps, compressors		
Special process hazard factor (F2)		3.62
Unit hazard factor (F3=F1*F2)		4.53
Damage factor		0.71
Fire and explosion index (F3 x MF)		95

Appendix 10-5: Fire and Explosion Index, calculation compressor (K02)

Table 10.5.1: Fire and Explosion Index calculation compressor (K02)

Exhibit A		
Material factor	21	
1. General process hazards	**Penalty**	**Penalty used**
Base factor		1.00
A. Exothermic reactions		
B. Endothermic reactions		
C. Material handling & transfer		
D. Enclosed process units		
E. Access		
F. Drainage		0.25
General process hazard factor (F1)		1.25
2. Special process hazards		
Base factor		1.00
A. Process temperature		
1. Above flash point		
2. Above boiling point		0.60
3. Above autoignition		
B. Low pressure (sub-atmospheric)		
C. Operation in or near flammable range		
1. Tank farm storage flammable liquids		
2. Process upset or purge failure		
3. Always in flammable range		
D. Dust explosion		
E. Pressure		0.99
F. Low temperature		
G. Quantity of flammable material		
1. Liquids or gases in process		0.72
2. Liquids or gases in storage		
3. Combustible solids in storage		
H. Corrosion and erosion		0.10
J. Leakage – joints and packing		0.30
K. Use of fired heaters		0.10
L. Hot oil heat exchange system		
M. Rotating equipment, pumps, compressors		0.50
Special process hazard factor (F2)		4.31
Unit hazard factor (F3=F1*F2)		5.39
Damage factor		0.76
Fire and explosion index (F3 x MF)		113

Appendix 10-6: Hazop study on turbo-expander (K01)

Table 10.6.1: Autoignition Temperature, LEL and UEL

Component	AIT[1] (K)[4]	LEL[2] (v%)	UEL[3] (v%)
Methane	616	5.3	15
Ethane	788	3	12.5
Propane	723	2.3	9.5
N-butane	678	1.9	8.5
N-Pentane	582	1.4	7.8
N-Hexane	498	1.2	7.4
N-Heptane	496	1.1	6.7
N-Octane	493	1	4.7

1) Autoignition Temperature
2) Lower Explosion Limit
3) Upper Explosion Limit
4) [60]

Table 10.6.2: Hazop study on turbo-expander (K01)

Guide word	Deviation	Possible Causes	Consequences	Action required
None	No flow	Line blockage Line fracture Control valve shut in error Molsieve beds give no output	If expander feed is blocked, compressor part will stop working. Pressure drop over compressor; reduced output; output pressure of sales gas not sufficient.	Installed pressure control on expander initiates shutdown.
		Absorber gives no output Coldbox blockage	If compressor feed is blocked, turbo-exp. will crash, due to destructive vibrations (due to very high expander wheel rotating speed)	A bypass should be installed on the expander to the compressor part
More	More flow	Temporarily more flow from gas field	Higher expander and compressor wheel velocities. In extreme cases turbo-exp. could overheat and crash.	Turbo-exp. is designed to allow 125% of design flow. Installed control valves regulates flow.
	Higher pressure	Higher feed pressure	In worst case rupture of e.g. pressure rings or expander housing. Higher pressure in rest of process (less good separations, product off spec, possible equipment damage)	Pressure control by installed control valves
Less	Less flow	Internal blockage of molsieve beds Line blockage Line fracture Valve leakage Absorber gives less output	Lower exp. and comp. wheel velocity. No serious damage is done.	Good communications with feed operator. Good maintenance of molsieve beds, timed replacing of zeolite. Regular controlling of transport lines and equipment.
	Lower			

Less	pressure	Lower feed pressure Valve leakage	Less cold is produced in exp.: product possibly off spec. Higher compression duty required in final comp.	Also covered by good communications. In extreme cases refrigeration required (if this occurs regularly).
Part	High water concentration High conc. of C_3^+	Different conditions at supply field; slug catcher (outside batt. limit) is not working well. Supply from another gas field	Regeneration cycle of molsieve shortens Higher streams of C_3^+; lower sales gas stream	Good communications with gas supplier Do nothing! (besides adjusting reboiler duties and main compressor duty)
More	Water entrainment.	Failure of slug catcher (which is outside batt. limit)	Regeneration cycle of molsieve shortens	Good comm.: make sure slug catcher is well maintained.
Other	Leakage of turbo-exp. Lube oil or seal gas system fails Control system breakdown	Pressure ring cracked Leakage in these systems. Electrical failure or wear.	Gas release into to the air; LEL can be reached: risk of expl. or fire. Turbo-expander could overheat Turbo-expander out off control	Gas detection system, which sounds alarm at for example 25% LEL and initiate shutdown at 70% LEL. Good maintenance of systems. Backup system, good maintenance and inspection.

Appendix 11-1: Purchase costs per unit of equipment

Table 11.1.1: Costs of absorber and columns

Absorber & Columns @ Mln Dfl 1997								
Name	Equip.	Type Source Quant.	Diam. (m) Page Price [1] range	H (m) Costs (Dfl)	Mat. Factor	Press. (bar) Factor [5]	Cost equip. (Mln Dfl.)	Total Costs (Mln Dfl)
C-01	Vessel	Vert.	5.7	16.4	Steel	30.0		
		[51]	p 17,		H II	1.4		
			20/40 [2]	294,000	1		0.412	
	Trays	Valve	5.7		AISI			
		[51]	p 18	625 [3]	405			
		24	AISI 405	15,900 [4]	1		0.383	0.794
C-02	Vessel	Vert.	3.5	6.2	Steel	30.0		
		[51]	p 17,		H II	1.4		
			20/40 [2]	71,000	1		0.099	
	Vessel	Vert.	6.5 [3]	16.9	Steel			
		[51]	p 17,		H II			
			20/40 [2]	343,000	1		0.481	
	Trays	Valve	3.5		AISI			
		[51]	p 18	775 [3]	405			
		11	AISI 405	7,500 [4]	1		0.082	
	Trays	Valve	6.5 [3]		AISI			
		[51]	p 18	625 [3]	405			
		28	AISI 405	15,900 [4]	1		0.447	1.108
C-03	Vessel	Vert.	2.9	33.5	Steel	16.0		
		[51]	p 17,		H II	1.2		
			12/15 [2]	224,000	1		0.269	
	Trays	Valve	2.9		AISI			
		[51]	p 18	820 [3]	405			
		61	AISI 405	5,400 [4]	1		0.330	0.599
C-04	Vessel	Vert.	2.5	24.0	Steel	9.0	0.175	
		[51]	p 17,		H II	1.1		
			12/15 [2]	159,000	1			
	Trays	Valve	2.5		AISI			
		[51]	p 18	885 [3]	405			
		40	AISI 405	4,300 [4]	1		0.174	349,000
Total	Absorber		(incl. C-01)					0.794
Total	Columns		(excl. C-01)					2.056

1) The prices for columns were, where necessary, extrapolated from the relevant prices ranges (see remark (2)) and then adjusted for the total weight of the column.

2) Ratio of mean wall thickness and length of tangent lines and indicates from which prize range the prices are taken.

3) The price for the large diameter valve trays was extrapolated from the price range on p. 18. By taking the natural logarithm for both the diameter as the price per m^2, a linear dependency was found. The price here given is the cost per m^2 in Dfl @ 1997.

4) Total cost per tray

5) [36, p. 223]

Table 11.1.2: Costs of heat exchangers

Heat Exchangers @ Mln Dfl 1997										
Name	M of C Sh/Tubes	Source	Area (m^2)	Based On (m^2) [3]	Costs/ Unit (Dfl)	Type[4]	Pres (bar)	Number	Pres. Fac.	Costs (Mln Dfl)
E-01	Alumina[1]	[51]	1000		265,000	Plate	30/72	4	2.3	2.434
E-02	CS/CS [2]	[51]	441	500	176,000	F.T.	30/21	1	1.6	0.282
E-03	CS/CS	[51]	497	300	170,000	F.H.	16	3	1.2	0.612
E-04	CS/CS	[51]	214	200	96,000	F.T.	16/21	1	1.2	0.115
E-05	CS/CS	[51]	479	500	168,000	F.H.	190	2	3.0	1.008
E-06	CS/CS	[51]	301	300	120,000	F.H.	9	1	1.1	0.132
E-07	CS/CS	[51]	142	100	68,000	F.T.	21/9	1	1.1	0.075
E-08	CS/CS	[51]	75	70	47,000	F.H.	9	1	1.1	0.052
E-09	CS/CS	[51]	77	70	49,000	F.H.	21	1	1.4	0.069
E-10	CS/CS	[51]	66	50	45,000	F.H.	9	1	1.1	0.050
Total										4.828

1) Plate heat exchanger
2) CS = Carbon Steel
3) This is the standard area from which the unit price is extrapolated. Extrapolated from the range of 240 to 300 m^2 for AISI 316 SS, which is more expensive than aluminum [30]
4) F.T. = Fixed Tubes, F.H. = Floating Head

Table 11.1.3: Costs of furnace

Furnace @ Mln Dfl 1997 [1]						
Name	Size S (kW)	Const. C	Equipm. Index n [2]	Costs C*S^n UK Pnds	Costs @ Mln Dfl 1992[3]	Costs @ Dfl 1997 [4]
F-01	1000	290	0.77	59,200	0.1776	0.193
Total						0.193

1) [36, p. 225]
2) for Carbon Steel
3) ROE, UK Pnds to Dfl 1992: 3.00
4) Index correction Ned, 1992: 140, 1997: 152.05, factor 1.09 [61, p. 9-64]

Table 11.1.4: Costs of compressor

Compressor @ Mln Dfl 1997 [1]							
Name	Size S [2] (kW)	Const. C	Equipment Index n	Correction factor	Costs C*S^n UK Pnds	Costs [3] @ Mln Dfl 1992	Costs @ Mln Dfl 1997 [4]
K-02	24,000	500	0.8	48	3,462,000	10.387	11.281
Total							11.281

1) [36, p. 225]
2) The size of this compressor greatly exceeds the range from which prices for compressors can be derived. Therefore, it was decided to use 48 (24.000/500) smaller units for cost estimation. The maximum pressure for the compressor type from which this assumption has been made is 50 bara, while the output pressure should be 190 bara. This is a reason for estimating the total compressor cost even higher. On the other hand, 48 units are much more expensive than one single unit. Therefore it is decided to multiply the cost of a 500 kW compressor by 48 to arrive on the estimation for the 24 MW compressor. This is called the correction factor.
3) ROE, UK Pnds to Dfl 1992: 3.00
4) Index correction Ned, 1992: 140, 1997: 152.05, factor 1.09 [61, p. 9-64].

Table 11.1.5: Costs of vessels

Vessels @ Mln Dfl 1997								
Name	Equip.	Type Source Quant.	Diam. (m) Page Price range	H or L (m) Costs (Dfl)	Mat. Factor	Press. (bar) Factor	Cost equip. (Dfl.)	Total Costs (Mln Dfl)
S-01/03	Vessel	Vert. [51] 3	2.53 p 16, 40/16[1]	6.65 79,000	Steel H II 1	110.0 2.5	197,500	0.593
V-01	Vessel	Vert. [51]	2.84 p 16, 60/13[1]	8.16 83,000	Steel H II 1	30.0 1.4	116,000	0.116
V-02	Vessel	Vert. [51]	2.97 p 16, 40/13[1]	6.80 77,000	Steel H II 1	30.0 1.4	108,000	0.108
V-03	Vessel	Vert. [51]	1.65 p 16, 20/13[1]	10.40 71,000	Steel H II 1	30.0 1.4	99,000	0.099
V-04	Vessel	Hor. [51]	3.00 p 16, 40/13[1]	4.60 70,000	Steel H II 1	16.0 1.2	84,000	0.084
V-05	Vessel	Hor. [51]	2.30 p 16, 20/10[1]	4.50 56,000	Steel H II 1	9.0 1.1	62,000	0.062
Total	Molsieves		(incl. S-01/03)					0.593
Total	Vessels		(excl. V-01/05)					0.469

1) Ratio of volume and mean plate thickness from which the prices is extrapolated.

Table 11.1.6: Costs of pumps

Pumps @ Mln Dfl 1997[1]					
Name	Power Calc (kW)	Power Actual (kW)	Cap. (m^3/h)	Unit Cost (Dfl)	Total cost (Mln Dfl)
P-01	11.81	12.5	562	65,000	0.065
P-02	7.58	10	204	22,500	0.022
P-03	1.05	1.3	54	15,000	0.015
P-04	9.23	11	179	24,000	0.024
P-05	16.45	18.5	90	20,500	-0.021
P-06	6.26	7.5	148	24,000	0.024
Total					0.171

1) [51]

Table 11.1.7: Costs of turbo-expanders

Turbo-expanders @ Mln Dfl 1997[1]					
Name	Type	Unit Cost (Mln Dfl)	Number[3]	Total Cost (Dfl @ 1999)	Total Cost[4] (Dfl @ 1997)
K-01&03	Mafi-Trench Frame 5	2.183[2]	2	4.366	4.205
Total			2	4.366	4.205

1) Source: Cost estimation provided by Mafi-Trench Company, see appendix 9-3.

2) This price includes an auxiliary support skid, API 614 style lube oil system, a seal gas supply system and a control system for an European location

3) Installed spares for both the Turbo-expanders

4) Index correction Ned, 1997: 152.05, 1999: 157.88, factor 0.96 [61, p. 9-64].

Appendix 11-2: Calculation of direct, indirect and fixed cap. costs

Table 11.2.1: Calculation of direct, indirect and fixed cap. costs

Equipment	Mln Dfl @ 1997
Absorber	0.79
Columns	2.06
Heat Exchangers	4.83
Furnace	0.25
Molsieves	0.59
Vessels	0.47
Pumps	0.17
Turbo-expander	4.21
Compressor	11.28
Purchase costs equipment	24.59
Direct capital costs[1]	83.61
Fixed capital costs[2]	121.23
Indirect capital costs[3]	37.62

Table 11.2.2: Conversion Dfl @ 1997 to Dfl @ 1999

Type of costs	Mln Dfl @ 1997	Mln Dfl @ 1999[4]
Direct	83.61	86.81
Indirect	37.62	39.07
Fixed	121.23	125.88

1) Direct capital costs are calculated from the purchased costs equipment by multiplying these by the Lang factor for a “fluids type” process: 3,40
2) Fixed capital costs are calculated from the direct capital costs by multiplying these by the Lang factor for “fluids type” process: 1,45
3) Indirect capital costs = fixed capital costs - direct capital costs
4) Dfl @ 1999 = Dfl @ 1997 * (157.88/152.05) ; International Plant Cost Index correction. In 1997: 152.05; in 1999: 158.77 (for the Netherlands). These values were extrapolated from [61, p. 9-64].

Appendix 11-3: Calculation of production costs

Table 11.3.1: Utility requirements

Utility Requirements						
Name	Function	Utility Type	Load (kW)	Requirements		
				CW (kt/a)	LP Steam (kt/a)	Electr. (MWh/a)
E-02	Reboiler	LP Steam			229.6	
E-04	Reboiler	LP Steam			111.2	
E-07	Reboiler	LP Steam			73.9	
E-03	Condensor	CW		8,990.8		
E-05	Cooler	CW		23,213.0		
E-06	Condensor	CW		5,585.3		
E-08	Cooler	CW		1,204.5		
E-09	Cooler	CW		292.2		
E-10	Cooler	CW		556.9		
P-01	Pump	Electr.	11.81			99.1
P-02	Pump	Electr.	7.58			63.8
P-03	Pump	Electr.	1.05			8.4
P-04	Pump	Electr.	9.23			77.3
P-05	Pump	Electr.	16.45			137.8
P-06	Pump	Electr.	6.26			52.9
K-02	Compresssor		24,088			202,309.2
F-01	Furnace	None				
Total				39,800	414.7	202,800
CW = cooling water						
Stream Hours/a :			8,400			
Steam, Heat of Condensation :			2,200 kJ/kg			
Cooling Water :			$T_{in} = 30°C$, $T_{out} = 38°C$; $C_P.\Delta T = 4.20*8 = 33.6$ kW/kg			

Table 11.3.2: Raw materials from mass balance (Mln Dfl @ 1999)

In/Out	Name	Stream number	Kton/a	Ton/ton Propane	Price (Dfl/kton)	Costs (Mln Dfl/a)
In	Feed	1	4329		0.168	500.00
	Total in		4329		0.168	500.00
Out	Sales Gas	34	3564		0.189	463.47
	Propane	54	342			
	Butane	61	212			
	Pentane$^+$	64	174			
	Fuel/Stack Gas	40	37		0	0
	Total Out		4329			463.47
In – Out =Raw materials cost						36.53

Table 11.3.3: Utilities (Mln Dfl @ 1999)

In/Out	Name	Unit	Units/a	Dfl/unit	Mln Dfl/a
In	Cooling Water	ktons	39,800	50.0	1.99
	LP Steam	ktons	414.7	30,000	12.44
	Electricity	MWh	202,800	130.0	26.36
Total = Utilities					40.79

Table 11.3.4: Zeolite (Mln Dfl @ 1999)

Zeolite	Per Bed fill (ton)	Dfl/ton	Per fill (Dfl)	Life Time (years)	Number of beds	Mln Dfl/a
	16.15	6180	99,830	4	3	0.075

Table 11.3.5: Labour costs (Mln Dfl @ 1999)

Labour	Workers/shift	Mln Dfl /(worker*a)	Dfl/shift	Number of shifts	Mln Dfl/a
Costs	10	0.10	1.00	5	5.00

Table 11.3.6: Production costs (Mln Dfl/a)

Variable Costs	Dfl/a (Mln)	Remarks	% of annual prod. Costs
Raw materials	36.53		28,50
Miscellaneous	0.63	10% of maintenance	0,49
Utilities	40.79		31,83
Catalist	0.08		0,06
Shipping & packaging	0.000	Negligible	0,00
Subtotal A	78.03		60,88
Fixed costs			
Maintenance	6.29	5% of Fixed Capital	4,91
Operating labour	5.00		3,90
Laboratory costs	1.00	20% of Operating Labour	0,78
Supervision	1.00	Idem	0,78
Plant overheads	2.50	50% of Op. Labour	1,95
Capital charges	12.59	10% of Fixed Capital [1]	9,82
Insurance	1.26	1% of Fixed Cap.	0,98
Local taxes	2.52	2% of Fixed Cap.	1,96
Royalties	1.26	1% of Fixed Cap.	0,98
Subtotal B	33.42		26,07
Direct costs (A+B)	111.45		86,96
Indirect costs			
Sales expense	16.72	15% of Direct Costs [2]	
General overheads			
R&D			
Subtotal C	16.72		13,04
Annual production costs (A+B+C)	128.16		100,00
Production costs	375 Dfl/ton propane [3]		
Raw material costs	107 Dfl/ton propane [3]		

1) The lifetime of the plant is 25 years. Therefore Capital charges totalling 10% of Fixed Capital are assumed to be quite adequate.

2) Sales expense, General overheads and R&D all together amount to 15% of Direct Costs. For this amount was decided considering the low R&D and Sales expense costs.

3) Propane, as main product, is assumed to account for all Annual Production Costs.

Appendix 11-4: Net Future Value and Net Present Value at 6.5% interest

Table 11.4.1: Net Future Value and Net Present Value at 6.5% interest

Total Investment: 163.64 Mln Dfl Net Annual Cash Flow: 31.37 Mln Dfl					Fixed Capital: 125.88 Mln Dfl				
End year no.	Net Future Values No discount – Annual & Accum.					Net Present Values Discounted – Accum.			
	Capital ann.	Costs Accum.	Nett ann. Cash flow	Acc. Cash flow	NFV	Disc. fact.; interest= 6.5 %[2]	Capital Costs	Acc.Cash Flow	NPV
	Dfl Mln	Dfl Mln	Dfl Mln	Dfl Mln	Dfl Mln	-	Dfl Mln	Dfl Mln	Dfl Mln
1	81.82	81.82			-81.82	1.00	81.82		-81.82
2	81.82	163.64			-163.64	0.94	158.65		-158.65
3			31.37	31.37	-132.27	0.88		27.66	-130.99
4			31.37	62.74	-100.90	0.83		53.63	-105.02
5			31.37	94.11	-69.53	0.78		78.01	-80.63
6			31.37	125.48	-38.16	0.73		100.91	-57.74
7			31.37	156.85	-6.79	0.69		122.41	-36.24
8			31.37	188.22	24.58	0.64		142.60	-16.05
9			31.37	219.59	55.95	0.60		161.55	2.90
10			31.37	250.96	87.32	0.57		179.35	20.70
11			31.37	282.33	118.69	0.53		196.16	37.41
12			31.37	313.70	150.06	0.50		211.75	53.12
13			31.37	345.07	181.43	0.47		226.49	67.84
14			31.37	376.45	212.81	0.44		240.32	81.68
15			31.37	407.82	244.18	0.41		253.31	94.67
16			31.37	439.19	275.55	0.39		265.51	106.86
17			31.37	470.56	306.92	0.37		276.96	118.32
18			31.37	501.93	338.29	0.34		287.72	129.07
19			31.37	533.30	369.66	0.32		297.82	139.17
20			31.37	564.67	401.03	0.30		307.30	148.65
21			31.37	596.04	432.40	0.28		316.20	157.55
22			31.37	627.41	463.77	0.27		324.56	165.91
23			31.37	658.78	495.14	0.25		332.41	173.76
24			31.37	690.15	526.52	0.23		339.78	181.13
25			31.37	721.52	557.88	0.22		346.70	188.05
26			31.37	752.89	589.25	0.21		353.196	194.55
27			31.37	784.26	620.62	0.19		359.30	200.65
Total			784.26	784.26	620.62	13.39	158.65	359.30	200.65
Ratio [Cash Flow/Capital]								2.27	
Net Present Value [Cash Flow - Capital]								200.65	

1) Construction time of 2 years is assumed.

2) Interest Rate is 6.5%. Discount Factor (D.F.) = $(1/(1+0,065))^n$, in which n = the year number from the first column minus 1

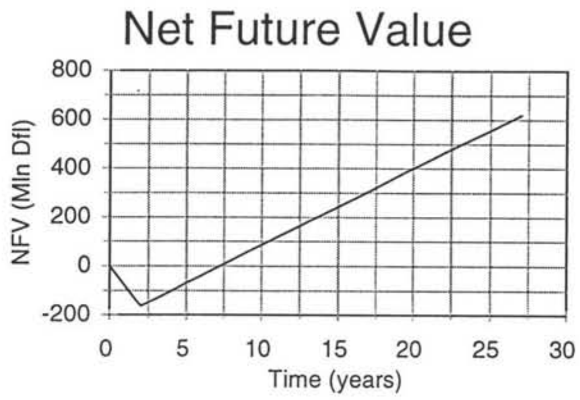

Figure 11.4.1: Net Future Value at 6.5% interest

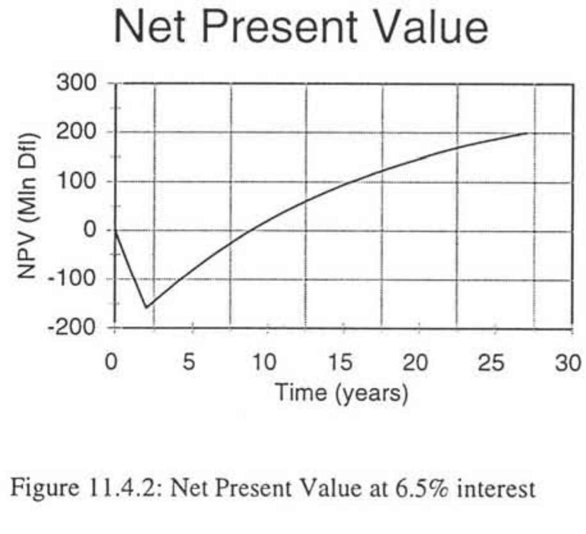

Figure 11.4.2: Net Present Value at 6.5% interest

Appendix 11-5: DCFROR

Table 11.5.1: Net Future Value and Net Present Value at 17.3% interest

Total Investment: 163.64 Mln Dfl Net Annual Cash Flow: 31.37 Mln Dfl					Fixed Capital: 125.88 Mln Dfl				
End year no.	Net Future Values No discount – Annual & Accum.					DCFROR			
	Capital Ann.	Costs Accum.	Nett ann. Cash flow	Acc. Cash flow	NFV	Disc. fact.; inte-Rest= 17.3 % [2]	Capital Costs	Acc.Cash Flow	NPV
	Dfl Mln	Dfl Mln	Dfl Mln	Dfl Mln	Dfl Mln	-	Dfl Mln	Dfl Mln	Dfl Mln
1	81.82	81.82			-81.82	1.00	81.82		-81.82
2	81.82	163.64			-163.64	0.85	151.56		-151.56
3			31.3	31.37	-132.27	0.73		22.79	-128.77
4			31.37	62.74	-100.90	0.62		42.22	-109.34
5			31.37	94.11	-69.53	0.53		58.78	-92.78
6			31.37	125.48	-38.16	0.45		72.90	-78.67
7			31.37	156.85	-6.79	0.38		84.93	-66.63
8			31.37	188.22	24.58	0.33		95.19	-56.38
9			31.37	219.59	55.95	0.28		103.93	-47.63
10			31.37	250.96	87.32	0.24		111.38	-40.18
11			31.37	282.33	118.69	0.20		117.73	-33.83
12			31.37	313.70	150.06	0.17		123.15	-28.42
13			31.37	345.07	181.43	0.15		127.76	-23.80
14			31.37	376.45	212.81	0.13		131.70	-19.87
15			31.37	407.82	244.18	0.11		135.05	-16.51
16			31.37	439.19	275.55	0.09		137.91	-13.66
17			31.37	470.56	306.92	0.08		140.34	-11.22
18			31.37	501.93	338.29	0.07		142.42	-9.14
19			31.37	533.30	369.66	0.06		144.19	-7.37
20			31.37	564.67	401.03	0.05		145.70	-5.86
21			31.37	596.04	432.40	0.04		146.99	-4.58
22			31.37	627.41	463.77	0.03		148.08	-3.48
23			31.37	658.78	495.14	0.03		149.02	-2.55
24			31.37	690.15	526.51	0.03		149.81	-1.75
25			31.37	721.52	557.88	0.02		150.49	-1.07
26			31.37	752.89	589.25	0.02		151.07	-0.49
27			31.37	784.26	620.62	0.02		151.56	0.00
To-tal			784.26	784.26	620.62	6.68	151.56	151.56	0.00
Ratio [Cash Flow/Capital]								1,00	
Net Present Value [Cash Flow - Capital]								0.00	

1) Construction time of 2 years is assumed.

2) Interest Rate is 17.3%. Discount Factor (D.F.) = $(1/(1+0,173))^n$, in which n = the year number from the first column minus 1

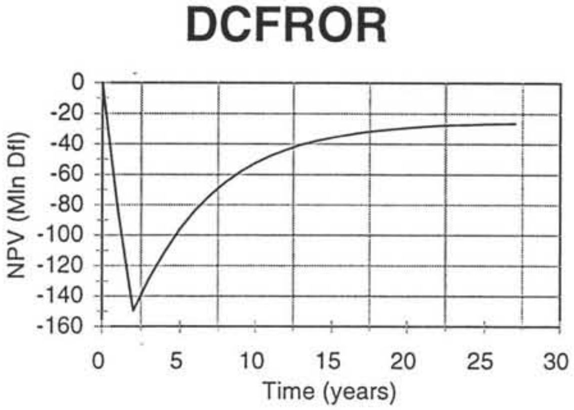

Figure 11.5.1: Discounted Flow Rate Of Return at 17.3% interest

Appendix 11-6: Maximum allowed investment calculation based on margin

Table 11.6.1: Maximum allowed investment calculation based on margin at an interest rate of 6.5%

Margin: 123.0 Mln Dfl Maximum allowable investment: 1453 Mln Dfl

End year no.	Net Future Values No discount – Annual & Accum.					Net Present Values Discounted – Accum.			
	Capital Ann.	Costs Accum.	Nett ann. Cash flow	Acc. Cash flow	NFV	Disc. fact.; inte-Rest= 6.5%[2]	Capital Costs	Acc.Cash Flow	NPV
	Dfl Mln	Dfl Mln	Dfl Mln	Dfl Mln	Dfl Mln	-	Dfl Mln	Dfl Mln	Dfl Mln
1	727	727			-727	1.00	727		-727
2	727	1453			-1453	0.94	1409		-1409
3			123	123	-1330	0.88		108	-1300
4			123	246	-1207	0.83		210	-1199
5			123	369	-1084	0.78		306	-1103
6			123	492	-961	0.73		396	-1013
7			123	615	-838	0.69		480	-929
8			123	738	-715	0.64		559	-850
9			123	861	-592	0.60		633	-775
10			123	984	-469	0.57		703	-706
11			123	1107	-346	0.53		769	-640
12			123	1230	-223	0.50		830	-579
13			123	1353	-100	0.47		888	-521
14			123	1476	23	0.44		942	-467
15			123	1599	146	0.41		993	-416
16			123	1722	269	0.39		1041	-368
17			123	1845	392	0.37		1086	-323
18			123	1968	515	0.34		1128	-281
19			123	2091	638	0.32		1168	-241
20			123	2214	761	0.30		1205	-204
21			123	2337	884	0.28		1240	-169
22			123	2460	1007	0.27		1273	-136
23			123	2583	1130	0.25		1303	-105
24			123	2706	1253	0.23		1332	-77
25			123	2829	1376	0.22		1359	-49
26			123	2952	1499	0.21		1385	-24
27			123	3075	1622	0.19		1409	0
Total			3075	3075	1622	13.4		1409	0
Ratio [Cash Flow/Capital]								1,00	
Net Present Value [Cash Flow - Capital]								0.00	

1) Construction time of 2 years is assumed.

2) Interest Rate is 6.5%. Discount Factor (D.F.) = $(1/(1+0,065))^n$, in which n = the year number from the first column minus 1

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