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Conceptual Process Design
Chemical Process Technology

CONFIDENTIAL

Subject

Integration of a propane dehydrogenation and IPAproduction facility (175 kton/a IPA)

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ERRATA final report CPD3242

"Integration of a propane dehydrogenation and IPA production facility"

- Summary: first paragraph, production capacities are mentioned in *ton/a* (not kt/a)
- Summary: 4th paragraph last sentence, "...fed back to the dehydrogenation *section*."
- p.7: production capacities are mentioned in *ton/a* (not kt/a)
- p.10: 2.5, first paragraph, "... should be able to reach a 81% propylene *conversion*"
- p.17: "In table *18* a margin calculation..", (not table 7)
- p.18: "Appendix 9 (table *9*.24)"
- p.20: 4.4, first paragraph, "the process consist*s* of two reactors."
- p.21: Table 22, propylene fraction: *85* %
- p.23: Table, reference misses
- p.24: 5.1, end of first paragraph, "(see.. appendix 5.*2*)" (instead of 5.1)
- p.24: 5.3, second paragraph, "...from the propylene reactor product ~~is~~, since.."
- p.25: 5.4, second paragraph, *reference [17]*, Coulson & Richardson
- p.26: 5.5, *note 3* can be removed
- p.26: 5.6, first part refers to V102, last part to V203
- p.33: 7.5, 4th paragraph, "330 C, *additional* hot utilities are.." (instead of additive)
- p.34: 8.2, second paragraph, " results in a change in ~~of the~~ conv."
- p.35: Table 29, reference Rob Berger [20] should be *[18]*
- p.36: 6th paragraph, "overall- design *consists* of.." (not exists)
- p.36 - 49: [18] should be *reference [17]* (Coulson & Richardson)
- p.39: second paragraph, " can be found in *table 30* " (not 31)
- p.39: 4th paragraph, "..results... are listed in *table 31*" (not 32)
- p.45: "in a run down tank takes ~~a~~ 24 hours"
- p.47: last paragraph "the number in the table must *be* explained"

Bullet lists should have been used on pages: 13, 30, 41 and 47.

Appendix final report CPD3242
CLARIFICATIONS

Simulation of reactor R101

Introduction

In simulating the dehydrogenation reaction problems may arise due to the complex reaction mechanisms. In literature large amounts of research is presented, based on a pure propane feed stream.

After reviewing literature we felt that using the reported selectivities [1]would hurt the results of the simulation. The strategy that was used to obtain the most realistic simulation is presented in figure 1.

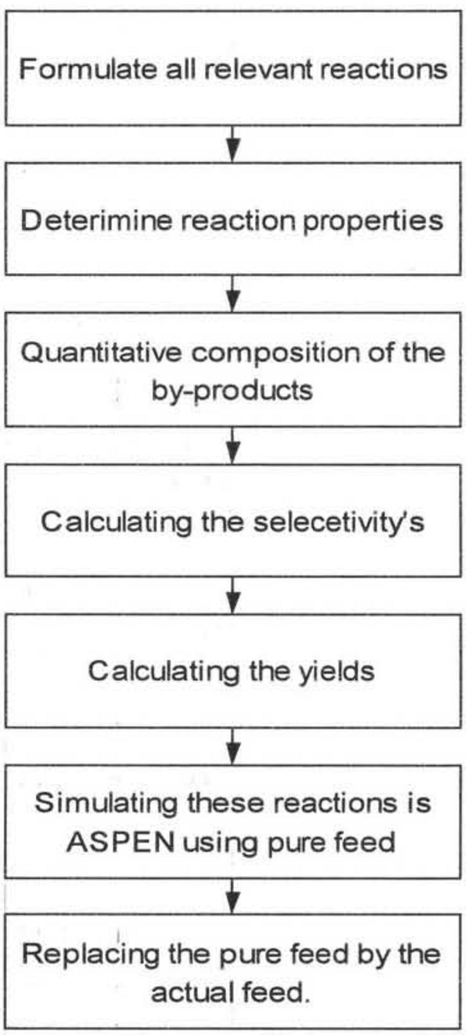

Figure 1, strategy applied in simulating the dehydrogenation reactor

Every step will be discussed in the following chapters.

[1] Because they were based on pure feed streams

Simulation steps

Formulate all relevant reactions

From literature:

$C_3H_8 \Leftrightarrow C_3H_6 + H_2$	$\Delta_r H^{298} = 127.7$ kJ/mol	(1)
$C_3H_8 \rightarrow 3C + 4H_2$	$\Delta_r H^{298} = 104.7$ kJ/mol	(2)
$C_3H_8 \rightarrow CH_4 + C_2H_4$	$\Delta_r H^{298} = -7.2$ kJ/mol	(3)
$C_3H_8 \rightarrow 2CH_4 + C$	$\Delta_r H^{298} = -44.1$ kJ/mol	(4)
$C_2H_4 + H_2 \Leftrightarrow C_2H_6$	$\Delta_r H^{298} = -46.3$ kJ/mol	(5)

Determine reaction properties

It can be seen that reaction (1) and (5) are equilibrium reactions. The reactants of reaction (5) are being produced by the other reactions.

Quantitative composition of the by-products

The composition (wt%) of the by-products are summarised in table 1 and are as described in literature [i] (hydrogen is not included).

Table 1, selectivities @ 650°C and 0.5 bara.

	Fraction [wt%]
Ethylene	1.1
Ethane	2.3
Methane	3.9
Carbon	7.7

Calculating the selectivities

Selectivities of reactions (2), (3), (4) and (5) were calculated[2] (see table 2) using:

1. By-product composition
2. Reported conversion and selectivity of reaction (1)

Table 2, reaction characteristics.

	Selectivity	Conversio	Yield
Reaction 1	0.879	0.900	0.791
Reaction 2	0.067	0.900	0.061
Reaction 3	0.003	0.900	0.003
Reaction 4	0.051	0.900	0.045
Reaction 5	0.753	0.900	0.677

[2] Note that the amount of cokes formed is the resultant of the by-product composition and the reported selectivity and conversion of reaction (1), and is therefore not a guess.

Calculating the yields

The conversion was arbitrarily set to 0.900. Using the calculated selectivities (table 2), yields were calculated, see table 2.

Simulating these reactions is ASPEN using pure feed

As mentioned before, two equilibrium reactions occur within the propylene reactor. Both equilibrium reactions were simulated in ASPEN using two equilibrium reactors. The equilibrium reaction (5), in which ethylene is converted to ethane, cannot be simulated without taking into account the irreversible reactions for the production of ethylene and hydrogen.

Schematically this looks like:

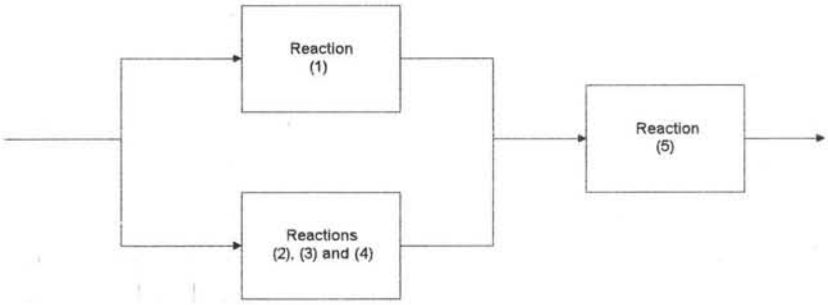

Figure 2, block diagram of the simulated propylene reactor.

The ASPEN-reactors were first simulated based on pure propane feed. By adjusting the available simulation parameters, the yields were obtained as reported in table 2.

The equilibrium constants used for the equilibrium reactions are calculated by ASPEN using the Gibbs energies at a specified temperature. A change in temperature at which the constant is calculated, results in a change in of the conversion, using this property the desired yields were obtained.

Replacing the pure feed by the actual feed.

After checking the conversions, selectivities and yields with the results from ASPEN (based on a pure feed), the pure feed was replaced by the actual feed, which contained o.a. propylene and ethane.

The effect of this change on e.g. reaction (1):
This equilibrium reaction was presented with propylene, which is also the product of the reaction. Therefore the amount of propane converted is likely to go down.

For this reason the conversion reported in literature, based on pure propane, is higher than the conversion found by the ASPEN simulation, see table 3.

Table 3, Propane conversion results.

Literature	90.0 %
Result ASPEN based on pure feed	90.0 %
Result ASPEN based on real feed	76.4 %

This result confirmed our initial assumption that selectivities as reported in literature cannot be used when changing the feed composition.

Column design history

Flowsheet subdivision

In order to let the Aspen model converge, the plant was split up in three sections:

Section 1: Unit 100
Section 2: Unit 200 except units C203, V203, C204 and V204
Section 3: The units C203, V203, C204 and V204

How the performance of these sections was calculated will be discussed in the following paragraphs.

Section 1

The thermodynamic models in Aspen Plus sufficed to describe the processes in this section. The whole section was designed using Aspens modelling techniques.

C101

First modelled as Shortcut column, results used as input for Radfrac model. This gave satisfying end results as soon as partial condensor was inserted. Aspen Reports were used as input for column design calculations.

C102

Modelled as Radfrac in Aspen. Initially satisfying mass flows were achieved at pressure of 1 bara. Later recommendations were made to operate the column at 29 bara (see Appendix 12.2). <u>No</u> changes were made to the simulation results (mass balance), however recommended operating conditions were used for column design calculations.

Section 2

Input for section based on output Section 1 (<137>). Columns were modelled separately as Radfrac and later combined to converge with recycles in the section as a whole. Results were used for the mass balance. Based on a (late) call of Shell columns C201 and C202 were altered separately. Resulting changes in mass balance were corrected manually in process stream summary to yield consistent balance over Section 2.

C201

First modelled as Shortcut column, results used as input for Radfrac model. Column design based on specification for bottom stream <202> (ref. Shell). Altered separately in Aspen with new specifications on reflux location (no condensor) and number of stages. Column design calculations based on output of this separate model.

V201

Thermodynamic models didn't suffice to model V201 as phase separator: no satisfying separation of IPA and NPA was achieved in Aspen. As a result V201 was modelled as ideal separator with split factors (appr. 1 for all components). Separator was included in section as a whole. Design calculation and assumptions as stated in report (Vessels, p.41).

C202

First modelled as Shortcut column, results used as input for Radfrac model. Column design based on specification for side stream <209> (CBM composition). Altered separately in Aspen with new specifications on side stream location, number of stages and new component flow input from separate C201 model (<209>). Column design calculations based on output of this separate model.

C205

First modelled as Shortcut column, results used as input for Radfrac model. Incorporated in section as a whole.

Section 3

Complete section modelled as ideal separator , since thermodynamic models proved inadequate for cyclohexane-water interactions (see Appendix 5.1). Calculation of split factors based on data (or assumptions based on practice) of incoming stream <220>, reflux ratio for C203, required output <221> and cyclohexane recovery. Results as in process stream summary.

C203

Modelled in Aspen as ideal separator with split factors. The number of trays and the relux ratio were provided by Shell. These figures were used to calculated the column dimensions in an ordinary manner. It should be noted that the comlumn dimensions were based on the top section only (as a result of unavailable data).

C204

Modelled in Aspen as ideal separator with split factors. Shell provided the dimension of a similar column.

Height: Was kept the same
Number of trays: Was kept the same
Diameter: The ratio feed to surface area was assumed to be constant
Pressure drop: Calculated in a similar manner as the other columns

For more details see the report, p.39.

LITERATURE

i G.F. Froment and K.C. Waugh et al., Reaction kinetics and the development of catalytic processes: proceedings of the international symposium, Brugge, Belgium, April 19-21, 1999 (paper: Kinetic Based Deactivation Modelling of an Isothermal Propane Dehydrogenation Reactor).

SUMMARY

Shell International Chemicals produces isopropyl alcohol (IPA) by hydration of propylene. IPA is used as a solvent in for example anti-freeze solutions or par~~fums. Shell~~ is market leader in selling IPA. Worldwide IPA production is 2 million kton/a. Shell produces 600.000 kton/a in six IPA production facilities. Exxon has only one IPA production facility, but produces 300.000 kton/a. Other competitors are Union Carbide and Condena. IPA production is based on propylene feedstock.

Propylene is produced as a by-product in catalytic cracking ~~or~~ and by steam cracking. Polymer grade propylene (99%) is made available by a PP-splitter (propane-propylene splitter). At 280 US$/ton it is expensive in comparison with lower purity grades, due to limited PP-splitting capacity. Demand for high purity propylene is expected to grow further in the future. Splitting capacity will grow, ~~which will put~~ ~~extra~~ pressure on the cost of both high purity and dilute propylene. Therefore it is important to look for cheaper alternatives ~~for~~ feedstock. with it, excerting additional

Propylene is already commercially produced by dehydrogenation of propane. Propane is relatively cheap (120 US$/ton), since it is produced in large quantities as a by-product in gas ~~extraction.~~ However this process still contains ~~the~~ an expensive PP-splitter. The combination of propane dehydrogenation and propylene hydratation may reduce the need of this separation step: propane-propylene mixtures can be ~~recycled to~~ produce IPA (550 US$/ton).

This report describes the integration of a propane dehydrogenation unit and an IPA unit on a green ~~field location~~, yielding an IPA production facility of 175 kton/a. Commercially four existing dehydrogenation processes and three IPA processes are available. Design of the dehydrogenation reactor in this concept is based on the Catofin process. Operating at high temperature and low pressure (650°C, 0.66 bara) it yields high conversion up to 90%. Break-even point for placing a PP-splitter is at a propylene concentration of 70 wt% in the dehydrogenation product stream. Current design contains 84.6 wt% in this stream. Design of the IPA reactor and the purification section is based on the Veba process (already used by Shell). Ninety percent of propane-propylene is recycled over the IPA reactor, the rest is fed back to the dehydrogenation. of $C_3^=$ in

The plant is designed for production of 182.5 kton IPA/a based on 8760 theoretical available hrs/a. Operating availability has been set to 8400 hrs/a, the difference including both actual downtime and slowdown of production. Result is a 175 kton/a IPA facility using 147 kton/a propane.

Economical plant life is set at 20 years, excluding 2 additional years for construction and plant start up. Maximum allowable investment is 582.6 M$. Total investment counts 78.28 M$. Rate ~~of~~ return is 5% (discount cash-flow-rate of return 1.14 %, POT 19.46 years). The utilities are the most important cost drivers of the process. A cut of 50% in the cost for electrical power would bring the POT to 12 years and the DCFFR above 6%.

Recommendations are made to optimise the performance of the purification section and to further analyse heat integration. Use of more accurate thermodynamic models could improve the design. Reductions should then be made in the use of utilities to make the process economically more viable.

TABLE OF CONTENTS

1 Introduction

Shell International Chemicals produces isopropyl alcohol (IPA)[1] by hydration of a propylene feedstock. At this moment Shell is market leader in selling IPA, which is mainly used as a solvent. IPA is a colourless, flammable liquid with a pleasant odour. It is the simplest secondary alcohol and was first prepared in 1885 by Berthelot while working with n-propanol. It was the first industrial chemical synthesized from a petroleum-derived olefin (1920).

Propylene is produced as a by-product in catalytic cracking or by steam cracking. For the production of IPA via direct hydration relatively pure propylene (>95%) is used whereas for indirect hydration a lower purity will suffice (>50%). High purity propylene is normally produced by a PP-splitter and is expensive relative to lower purity propylene due to limited PP-splitting capacity and high demand. The demand for high purity propylene is expected to grow further in the future. With it, splitting capacity will grow, which will put extra pressure on the propylene pricing of not only the pure but also the dilute propylene streams. It is therefore important to look for cheaper alternatives. The most promising option is propane.

Propane is produced in large quantities as a by-product in gas extraction; it is already dehydrogenated and purified to polymer grade propylene commercially. At this moment the production costs of this propylene are still relatively high. The combination of propane dehydrogenation and IPA production may reduce the need of some expensive separation steps, i.e. no PP-splitting capacity is required.

This report describes the integration of a propane dehydrogenation unit and an IPA unit yielding an integrated IPA production facility.

The new unit operates on a 100% propane feedstock for evaluation purposes. The propylene reactor converts the incoming propane to propylene. The light ends in the propylene reactor effluent are then removed by distillation and the process stream is ~~led into~~ the IPA reactor. The feed ~~incoming~~ propylene is then converted to IPA. Finally IPA is separated from the IPA reactor effluent by a separation section.

Table 1 clearly shows the incentive to make IPA from propane instead of propylene. Other properties of all components can be found in Appendix 5.3 'Pure Component Properties'.

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[1] Also known as; 2-propanol, secondary propyl alcohol, per-spirit, isopropanol, dimethylcarbinol, petrohol, avantine

Table 1, Price of main components

Component	Formula	Mole. Weight (g/mole)	Price (US$/ton)
Propane	C_3H_8	44	120
Propylene	C_3H_6	42	280 (Polymer grade, 99% propylene) 250 (Chemical grade, 95% propylene) 180 (Refinery grade, 70% propylene)
IPA	C_3H_8O	60	550

Commercial processes for converting propane to propylene and direct hydration of propylene to IPA are widely available, but are licensed to companies other than Shell. See Chapter 2 'Process Options and Selections' for further detail.

The worldwide IPA production is 2 million kton/a of which Shell produces 600.000 kton/a. Shell has six IPA production facilities. Exxon has only one IPA production facility, but produces 300.000(!) kton/a. Other competitors are Union Carbide and Condena.
The total IPA production is based on a propylene feedstock. There is no integrated facility which operates on a propane feedstock.

US $/t.

	280	250	180
	120	120	120
Up-grading	160	130	60

How much t C_3^0 / t IPA?

2 Process Options & Selections

Purpose of the plant to be designed is to produce isopropyl alcohol (IPA) from propane. At this moment, this is done in two different plants: in the first, propane is dehydrogenated to propylene; in the second, propylene is hydrated to IPA. For the dehydrogenation process, there are four commonly used processes, for the hydration process there are two. This chapter will explain which of the 8 possible combinations was chosen and why.

2.1 Process options

In the next section four dehydrogenation options are discussed. In section 2.1.2 a similar discussion is given for the two IPA production facilities. Integration possibilities can also be found in the realisation of recycles as will be discussed in paragraph 2.1.3.

At the end of this chapter a discussion is given of which combination of options yields maximum integration and which combination is chosen.

2.2 Dehydrogenation options

Commercially available processes for the production of the propylene are:

Oleflex	(UOP)[1]
Catofin	(ABB Lummus Crest)[2]
STAR	(Phillips Petroleum)[3]
FBD-4	(Snamprogetti-Yarsintez) [4]

In appendix 14.1 facts are compared for all options, this has resulted in the pro's and con's summarised in table 2.

Table 2, comparison of dehydrogenation plants.

	Oleflex	Catofin	STAR	FBD-4
Equipment	-	+	-	-
Reaction temperature	+	-	-	-
Reaction pressure	+	-	+	+
Conversion	--	++	--	?
Selectivity	+	+	+	?
PP-splitter needed	-	++	-	?

Based on table 2 it can be seen that the Catofin option has a high conversion. This is due to the fact that the operating pressure is sub atmospheric, which will be costly. On the other hand, the high conversion may make a PP-splitter unnecessary, which would be a big money saver. Besides the low pressure, Catofin operates at high temperature to create optimum conditions for the equilibrium dehydrogenation reaction.

2.3 IPA production options

The commercially available processes[5] for the direct hydration of propylene to isopropyl alcohol are:

Texaco	(Deutsche Texaco)
Veba	(Veba-Chemie)
Tokuyama-Soda	(Tokuyama Soda Co., Ltd.)

However the extreme process conditions of the Tokuyama-Soda process make this option liable to be left out from the discussion. Indirect hydration with sulphuric acid (H_2SO_4) has not been examined because of its obsolete technology.

In appendix 14.2 facts for the two relevant options are summarised.

5.6%

Some other considerations:
The Veba process needs voluminous recycles due to its low conversion of 5% per pass. It seems that the recycle will easily grow larger at higher propane/propylene ratios. An advantage of the Veba process is that Shell has knowledge about this process and will therefore be able to better evaluate the surplus value of the integration itself.

The Veba process employs a clay catalyst with phosphoric acid. The acid slowly leaches off, so the catalyst needs to be soaked every 2 years. The Texaco process uses an ion exchange catalyst (surface couple of 100 m^3) which should be replaced every nine months.

The Texaco process uses 3 times more reactor volume than Veba does, this has a negative effect on the fixed capital for the Texaco option. Due to this and other factors, such as the licence, the capital cost needed for the Texaco option will end up 40% more expensive. Coupled with capital cost, maintenance costs are higher (2% of capital/a). Because of the large recycle around the Veba rector, energy costs will be higher. However, high-pressure steam (12 bar) can be used twice: for the compressors (12 to 8 bar) and the distillation columns.

In table 3, all pros and cons are summarised.

Table 3, comparison of IPA plants.

	Veba	Texaco
Knowledge	++	-
Capital cost	++	-
Operational cost	=	=
Performance (yield, quality)	=	=
Feedstock flexibility	-	+
Comparison with existing facilities	+++	-
Catalyst	+	-
Internal recycle	--	++
Conversion per pass	--	++
Pressure	+	-
Reaction temperature	-	+
Minimal propylene/propane ratio	-	+
Water to propylene ratio	+	-

2.4 Possible recycles

Currently operated IPA processes are based on a high quality grade propylene, which is recycled around the reactor. To prevent a build-up of inerts, such as propane, part of the recycle is vented. Increasing the propane concentration would result in higher recycles around the IPA reactor and a larger vent. This vent contains valuable propane which can also be recycled to the dehydrogenation reactor.

2.5 Process concept chosen

As mentioned before, integration should result in a reduction in the use of energy and/or utilities and preferably equipment. From experience, Shell knows that the break-even point for placing a PP-splitter is at a propylene concentration of about 70 wt% in the outlet flow of the dehydrogenation section. The Catofin process is the only dehydrogenation process that is able to meet this, according to literature it should be able to reach a 81% propylene (see appendix 14.1). For this reason, the Catofin process was taken as the basis for the dehydrogenation process.

Although arguments may favour the Texaco process, the Veba process is chosen for strategical reasons: the added value of the better opportunities for comparing the outcome to existing plants outweigh the chance of a better overall solution.

From the PFS it can be seen that an increasing recycle back to the dehydrogenation reactor will effect relatively more units than sending it back to the IPA reactor. This means that minimising the recycle ~~back~~ will have a positive effect on the total investment required.

A split factor α can be instituted at the split point: a fraction α goes to one recycle, a fraction (1-α) to the other. In appendix 8 a summary is given of the mass balances at different recycle ratios.

$$\alpha = \frac{R_{Internal}}{R_{Internal} + R_{External}} \qquad [2.\ 1]$$

$R_{Internal}$ = recycle over the IPA reactor, kg/h

$R_{External}$ = recycle back to the Propylene reactor. kg/h

The results showed that α should be chosen as high as possible, only limited by the bleed that shouldn't become too large.

Because of these considerations the concept will be a Catofin-Veba process with optimal recycles.

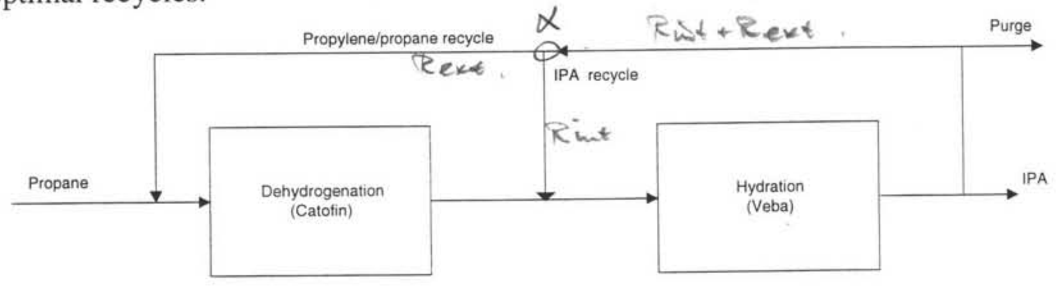


Figure 1, the chosen concept schematically presented.

3 BOD

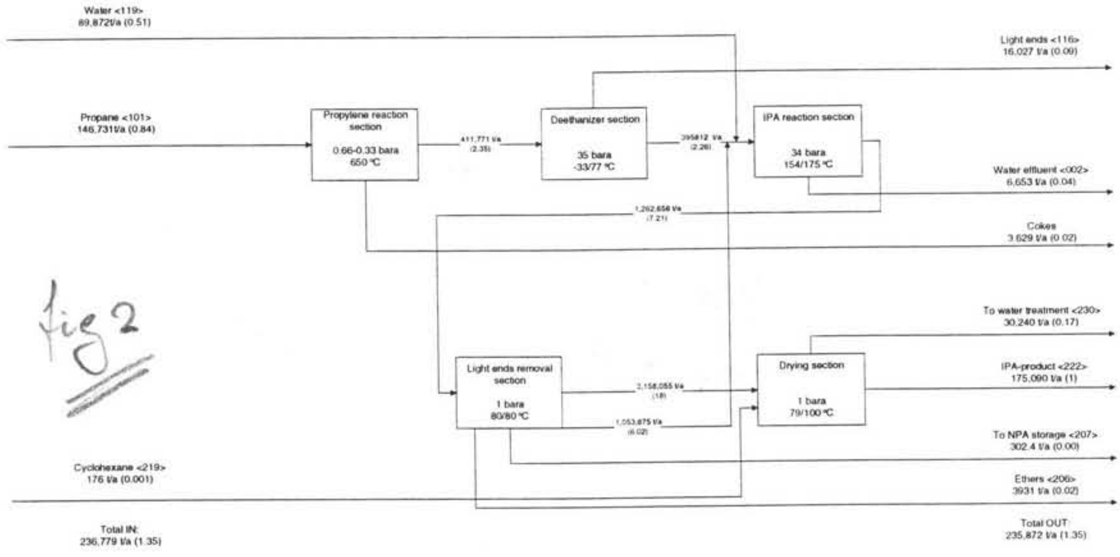
3.1 Description of Design

In chapter 2, several options for combining different processes have been discussed and the option of combining the Catofin process and the Veba process has finally been selected. Although the combination of the Catofin process with the Deutsche Texaco process also looks very promising, it was not chosen because of Shell's in-house expertise on the Veba process, which is already employed in three of the six Shell IPA facilities and offers a better opportunity to compare this design with these facilities.

The process uses a 100% propane feedstock (for evaluation purposes). This feedstock (stream <101>) is mixed with the overall recycle and then enters the propylene reactor (R101), where propane is converted to propylene under sub-atmospheric conditions. The reactor effluent (stream <107>) is then compressed to 34 bara and the light ends are removed in C101. Stream <122> then enters the IPA reactor, where only 5% of the ingoing propylene is converted to IPA. The IPA reactor therefore has a fairly large recycle. The IPA-water mixture and the propylene-propane mixture are then separated. 90% of the propylene-propane mixture is recycled back to the IPA reactor (stream <131>) and 10% is recycled back to the propylene reactor (stream <134>). The IPA-water mixture enters the separation section through stream <137>, IPA and water are separated and through stream <222> premium grade IPA enters the run down tanks for specification analysis.

3.2 Block schemes

Two levels of block schemes have been made. Figure 1 gives an overall view of the process. Figure 2 shows the process in more detail, all major equipment is also represented. In fig 2 the equipment coding and stream numbers correspond to the coding and numbers used in the Process Flow Scheme in appendix 1.
N.B.: Numbers between brackets () are t/t values.



0.09
0.04
0.02
0.17
1.00
0.00
0.02

Figure 2, Extended block scheme of the integrated process

3.3 Thermodynamic properties

An extensive collection of possible thermodynamic models is present in process simulators. The model that was used throughout the whole design is the Non Random Two Liquid method (Redlich Kwong, NRTL-RK). All needed parameters were estimated in Aspen Plus with the Unifac method. For further details see appendix 5.1.

3.4 Pure component properties

A list of pure component properties is given in appendix 5.3 'Pure Component Properties'.

3.5 Basic assumptions

Plant capacity

Plant capacity is set to 175 kton per year. This was done after consulting Shell Chemical Nederland BV. The plant will operate 8400 h/a, therefore all streams are based on a production of 20.81 ton/h IPA. The plant life is taken 22 years.

Plant availability

Theoretical technical availability of the plant: 365 days/yr * 24 hrs/day = 8760 hrs/yr. However, operating availability has been set at 8400 hrs/yr (= 23 hrs/day) for two reasons.
Technical unavailability: this is actual downtime. Equipment is under construction or being repaired. There is no production.
Production slowdown. Production is slowed down in order to meet certain (product) specifications. This can also be expressed in downtime.

In order to make up for this down time, the plant has been designed for a production of 182.5 kton IPA/year. Operating production will be 182.5 *(8400/8760) = 175 kton IPA/year.

Table 4, plant availability

	Theoretical	In practice	Efficiency
Availability [h/a]	8760	8400	0.96
Annual production [kton/a]	182.5	175	0.96

u ?

Location

During this design study a green field location in Malaysia will be assumed to be the location.

Utilities

site

Because of the green ~~field~~ location , utilities will be assumed to be the same as on other Shell locations. For a complete overview of the available utilities and their conditions see appendix 11 'BOD Utility conditions'.

Product specifications

The product stream contains 99,8% IPA and is the only desired product of this process. Other products, which are formed, are considered by-products.

Table 5. IPA-product specifications.

IPA product purity	>99.8%
Water%	<0.1%
Acidity	<0.002 mg/l

Definition of all in-and outgoing streams

There are three ingoing streams:

Table 6 , ingoing streams, stream <101>

Stream number	<101>	
Stream name	PROPANE FEED	
Composition [wt%]	Propane	100%
Mass stream [kg/h]	17468	
Mass stream [kton/a]	147	
Pressure [bara]	15	
Temperature [°C]	35	
Phase S/L/V	L	
Costs [US$/ton]	120	
Remarks	100% propane for evaluation purposes 7x f. Shell [illegible]	

Table 7, ingoing streams, stream <119>

Stream number	<119> 11.10	
Stream name	MAKE UP WATER [illegible]	
Composition [wt%]	Water	100%
Mass stream [kg/h]	10699	
Mass stream [kton/a]	90	
Pressure [bara]	1	
Temperature [°C]	25 x10	
Phase S/L/V	L	
Costs [US$/ton]	1.25	
Remarks	Process water	

Table 8, ingoing streams, stream <219>

Stream number	<219>	
Stream name	MAKE UP CYCLO HEXANE	
Composition [wt%]	Cyclohexane	100%
Mass stream [kg/h]	21	
Mass stream [kton/a]	0.18	
Pressure [bara]	1	
Temperature [°C]	25	
Phase S/L/V	L	
Costs [US$/ton]	n.a. ("miscellaneous materials")	
Remarks	Azeotrope agent	

There are seven outgoing streams:

Table 9, outgoing streams, stream <116>

Stream number	<116>	
Stream name	FROM ACCUMULATOR C101	
Composition [wt%]	HYDROGEN	42.7
	METHANE	15.4
	ETHYLENE	11.2
	ETHANE	13.6
	ETHANOL	0.0
	PROPYLEN	15.8
	PROPANE	1.3
Mass stream [kg/h]	1900	
Mass stream [kton/a]	15.96	
Pressure [bara]	34	
Temperature [°C]	-33	
Phase S/L/V	V	
Price [US$/ton]	120	
Remarks	Used as fuel gas	

Table 10, outgoing streams, stream <206>

Stream number	<206>	
Stream name	VENT	
Composition [wt%]	PROPYLEN	52.8
	PROPANE	12.7
	ACETONE	16.9
	DIPE	17.6
Mass stream [kg/h]	452	
Mass stream [kton/a]	3.8	
Pressure [bara]	1	
Temperature [°C]	34	
Phase S/L/V	V	
Price [US$/ton]	120	
Remarks	Used as fuel gas	

Table 11, outgoing streams, stream <207>

Stream number	<207>	
Stream name	TO NPA STORAGE	
Composition [wt%]	NPA	100
Mass stream [kg/h]	19	
Mass stream [kton/a]	0.16	
Pressure [bara]	1	
Temperature [°C]	34	
Phase S/L/V	L	
Price [US$/ton]	120	
Remarks	Used as fuel gas if it can not be sold at a higher price	

? = liquid!

Table 12, outgoing streams, stream <222>

Stream number	<222>	
Stream name	IPA PRODUCT	
Composition [wt%]	IPA	99.8
	NPA	0.1
	Water	0.1
Mass stream [kg/h]	20845	
Mass stream [kton/a]	175	
Pressure [bara]	1	
Temperature [°C]	30	
Phase S/L/V	L	
Price [US$/ton]	550	
Remarks	Main product	

Design

Better ? 99.9

Table 13, outgoing streams, stream <230>

Stream number	<230>
Stream name	TO WATER TREATMENT
Composition [wt%]	Cyclohexane 0.6
	Water 0.94
Mass stream [kg/h]	3584
Mass stream [kton/a]	30
Pressure [bara]	1
Temperature [°C]	30
Phase S/L/V	L
Costs [US$/ton]	1.67
Remarks	Waste water

?

Ref. Steele.

Table 14, outgoing streams, stream <001>

Stream number	<001>
Stream name	COKES
Composition [wt%]	C 100
Mass stream [kg/h]	432
Mass stream [kton/a]	3.6
Pressure [bara]	0.5
Temperature [°C]	650
Phase S/L/V	S
Price [US$/ton]	n.a.
Remarks	Used to heat the bed during regeneration

Table 15, outgoing streams, stream <002>

Stream number	<002>
Stream name	WATER EFFLUENT
Composition [wt%]	100
Mass stream [kg/h]	800
Mass stream [kton/a]	6.7
Pressure [bara]	1
Temperature [°C]	35
Phase S/L/V	L
Costs [US$/ton]	10.5
Remarks	Waste, evaporated during regeneration of zeolite

The amounts and conditions of the utilities used can be found in appendix 11 'BOD utility conditions'.

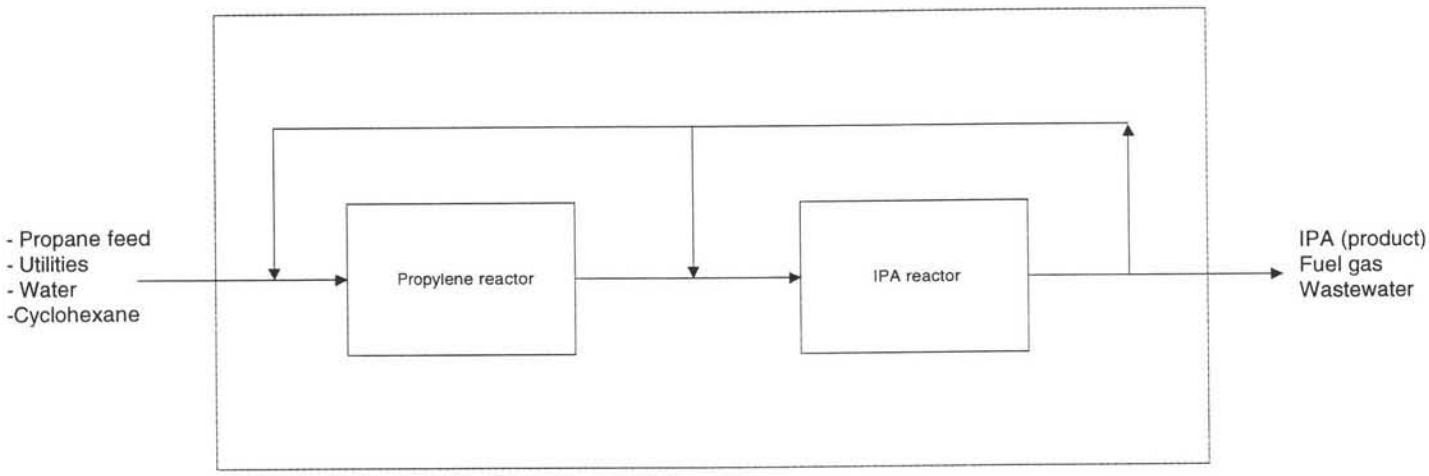
3.6 Wastes

Only stream <230> TO WATER TREATMENT and stream <002> WATER EFFLUENT are considered waste streams. Stream <230> needs water treatment to remove the small amount of cyclohexane, stream <002> is evaporated into the open air.

3.7 Battery limit

, originating
The next figure shows the battery limit that was set after consulting Shell Nederland BV. Everything within these boundaries is designed. Streams coming from outside the battery limit are considered to be constantly available; Streams leaving the battery limit are left out of the design as soon as they cross the battery limit.

Figure 3, battery limit



Economic margin

As discussed in detail previously, the aim of this project is to design a plant producing IPA based on a propane feedstock. In table 7 a margin calculation is given based on an annual production of 175,000 tons.

Table 16, raw materials

	Flow	Prices	Total
	ton/a	[$/ton]	[k$/a]
Propane	146748	120.00	17610
Process Water	89880	1.25	112
Total cost			17722

Table 17, products

	Flow	Prices	Total
	[kton/a]	[$/ton]	[k$/a]
IPA	175	550	96142
Self made fuel gas	15.96	120	1915
NPA	0.17	120	20
Total			98057

Table 18, margin

	Total
	[million $/a]
Products	98
Raw materials	18
Margin [million $/a]	80

The economical margin is positive. Based on this margin the project seems to be economically feasible.

The economical plant life is set for 20 years, excluding 2 additional years for construction and plant start up. Appendix 9 (table 24) shows the calculation of the maximum allowable investment based on this economical plant life, an earning power of 10% and an annual cash flow equal to the economic margin. It can be found that the maximum allowable investment is equal to 582.6 M$.

4 Thermodynamic Properties and Reaction Kinetics

For reliable simulations regarding the reaction and separation sections in the process, adequate kinetics and algorithms are needed. This chapter indicates the applied models and the assumptions that were made. It describes the reaction kinetics used for the reactor design.

4.1 Operating windows

Regarding the operating conditions the process flow scheme can be divided in three main parts, containing respectively the propylene reactor, IPA reactor and separation units C201 to C205. Operating windows regarding temperature and pressure for these sections are tabulated below.

Table 19. Operating windows for different process sections (listed by main containing equipment)

Equipment nr.	Name	Temperature [°C]	Pressure [bara]
R101	Propylene reactor	650	0.66
R102	IPA reactor	175	34
C201-205	Purification section	30-80	1

4.2 Component properties and azeotropic data

In appendix 5.3 pure component properties are shown. Table 20 contains enthalpies and Gibbs energies of formation for selected components. Pressure-temperature dependencies were calculated based on data from Reid and Prausnitz [6] for the design of shortcut separation columns.

Table 20, thermodynamic properties

Component	ΔH_f^0 [kJ/kg]	ΔG_f [kJ/kg]	Cp [kJ/kgK]
HYDROGEN	0,00	0,00	
METHANE	-4,68	-3179,38	2,24
ETHYLENE	1,87	-1176,79	1,57
ETHANE			
ETHANOL	-5,11	-3660,87	1,44
PROPYLENE	0,49	1494,29	1,54
PROPANE	-2,36	-533,86	1,71
ACETONE			
IPA	-4,54	-2891,67	1,55
NPA	-4,28	-2698,33	1,46
CYCLOHEXANE			
N-HEXANE			
DIPE			
WATER	-13,44	-12711,11	1,87
CARBON			

Remarks: (1) enthalpies of formation at T=298,2 K
(2) properties of the ideal gas state
(3) heat capacities of the ideal gas at 30 °C

The IPA process deals with an IPA-water azeotrope, i.e. IPA and water at a certain composition cannot be separated in a normal distillation column. However making use of cyclohexane as an entrainer makes it possible to obtain almost pure IPA with azeotropic distillation. Liquid-liquid inmiscibility of cylohexane and water enables the recovery of the entrainer (concept according to Perry, p.13-74 6 [7]). Table 21 shows the azeotropic compositions of the involved components.

Table 21. Azeotropic compositions for water-IPA-cyclohexane [7]

Components			Azeotropic data			
A	B	C	B.P. [°C]	Wt.% A	Wt.% B	Wt.% C
Water	IPA		80.3	12.6	87.4	
Water	Cyclohexane		68.95	9	91	
IPA	Cyclohexane		68.6	33	67	
Water	IPA	Cyclohexane	64.3	7.5	18.5	74

4.3 Thermodynamic models and data validation

An extensive collection of possible thermodynamic models is present in process simulators. The model that was used throughout the whole design is the Non Random Two Liquid method (Redlich Kwong, NRTL-RK). All needed parameters were estimated in Aspen Plus with the Unifac method.

To evaluate if this model gives good descriptions of the behaviour of the mixtures in the operating windows, temperature-composition pressure-composition diagrams were checked on validity (appendix 5.1).

It can be seen that the interactions of cyclohexane and water are not properly estimated by the NRTL-RK method. Using the Peng-Robinson method or importing Dechema parameters in the process simulator gave no improvements [8]. Parameter estimation with the ASOG method (Analytical Solution Of Groups) was not possible since the procedure does not apply for the combination of cycloalkanes and water [9]. All separation units treating mixtures of cyclohexane and water were thus modeled as ideal separators with split factors based on practice.

Data accuracy

All property parameters were estimated with the Unifac method. Appendix 51 (p A.5.1-2&3) shows calculation of the vapour liquid equilibrium of selected mixtures with Unifac [10]. Results are plotted against experimental data. Accuracy of the estimations is within limits of 5%.

4.4 Reaction kinetics

As can be seen in de PFS, the process consist of two reactors. The first dehydrogenates propane to produce propylene. The second hydrates propylene to obtain IPA. The reactions that take place, in both reactors, will be discussed in the following section.

Propylene production

Reactions that take place in the dehydrogenation reaction [11]:

$C_3H_8 \Leftrightarrow C_3H_6 + H_2$	$\Delta_r H^{298} = 127.7$ kJ/mol	(1)
$C_3H_8 \rightarrow 3C + 4H_2$	$\Delta_r H^{298} = 104.7$ kJ/mol	(2)
$C_3H_8 \rightarrow CH_4 + C_2H_4$	$\Delta_r H^{298} = -7.2$ kJ/mol	(3)
$C_3H_8 \rightarrow 2CH_4 + C$	$\Delta_r H^{298} = -44.1$ kJ/mol	(4)
$C_2H_4 + H_2 \Leftrightarrow C_2H_6$	$\Delta_r H^{298} = -46.3$ kJ/mol	(5)

It can be seen that reaction (1) and (5) are equilibrium reactions. The reactants of reaction (5) are being produced by the other reactions. In section 8.1 a more detailed report is given on the configuration used in Aspen to simulate these reactions.

Calculations concerning the selectivity and yields of the reactions (2)-(5) are discussed in appendix 5, the conversion was reported to be 0.90[12]. All calculations are based on the 'by-product' composition as reported by Stitt[13], see table 22. A summary of the selectivity's and conversions of the reactions are presented in table 23 at 650 °C and 0.5 bara.

Table 22, ~~selectivity's~~ @ 650°C and 0.5 bara.

	Fraction [wt%]
Ethylene	1.1
Ethane	2.3
Methane	3.9
Carbon	7.7
Propylene	
Total	100.0

Table 23, reaction characteristics.

	Selectivity	Conversion	Yield
Reaction 1	0.879	0.900	0.791
Reaction 2	0.067	0.900	0.061
Reaction 3	0.003	0.900	0.003
Reaction 4	0.051	0.900	0.045
Reaction 5	0.753	0.900	0.677

The selectivity of the main reaction (1) lies within the range as reported[12]. These calculated yields determine the product composition which is given in table 22.

It should be noted that these results are based on a pure propane feed, which is not used in this design as feedstock for the propylene reactor (see appendix 2).

Literature [2] suggests equation 4.1, which means that lowering the pressure will increase the conversion of propane.

$$K_p = \frac{\xi^2 \cdot P}{\left(1-\xi^2\right)} \qquad [4.\ 1]$$

K_P = Equilibrium constant, Pa

ξ = Conversion of propane, - -

P = Reaction pressure, Pa

The equilibrium constant is temperature dependent, and can be found in literature [14]

$$K_p \propto 10^{\sum \frac{v_i K_i}{R \cdot T}} \qquad [4.\ 2]$$

v_i = Stoichiometric coeffcient, - -

K_i = Equilibrium constant of the formation of compound i

R = Gas constant

T = Temperature

In this case the latter equation implies that an increase in temperature will have a positive effect on the position of the equilibrium. The other reactions will also favour from a temperature rise but less than the main equilibrium, which results in a positive effect on the selectivity.

IPA production

Inside the IPA reactor a vapour and solid phase are present. The solid phase is the phosphoric catalyst and the vapour phase contains the steam and the propylene/propane feed stream. Propylene and steam are converted to IPA in a thin layer of liquid H_3PO_4, which forms inside the capillaries of the catalyst-sphere, when steam enters the catalyst. The rate determining step is the diffusion of water and propylene into the thin layer of liquid H_3PO_4.

The limiting factor will probably be the diffusion of the propylene in the condensed water phase. It can be seen that the kinetics for the production of IPA are very complex, and are therefore not used. Instead reported conversions have been used.

Table 24 Operating Conditions of the IPA reactor

Reactor temperature	175 ^{0}C
Reactor pressure	34 bar
Conversion/pass on C_3H_6	0.0574
Main product	IPA
Byproducts	Acetone / Coke / DIPE / n-propanol / Propane / Polymer
Water/propylene ratio (moles)	0.3-0.35
Deactivation of catalyst	± 2 years
Phases in the reactor	V/S

Assumptions for by-products and their selectivity.

Main reaction

Isopropyl alcohol	$C_3H_6 + H_2O$	$\rightarrow C_3H_8O$	(1)

By-product reactions

Acetone	$C_3H_6 + H_2O$	$\rightarrow C_3H_6O + H_2$	(2)
Di-Isopropyl Alcohol	$2\ C_3H_8O$	$\rightarrow C_6H_{14}O + H_2O$	(3)
1-Propanol (NPA)	$C_3H_6 + H_2O$	$\rightarrow C_3H_8O$	(4)

Some by-products, which are formed, are not included in this model. The form~~ing~~ of propane, cokes and polymer are not included as selectivity data could not be found.

Product/by-product	Selectivity $C_3^{=}$
IPA	0.97500
Acetone	0.00375 0.99405
DIPE	0.00530 $\neq$ 1.0000
n-propanol	0.01000
Polymer	No data available
Propane	No data available
Coke	No data available

5 Process Structure & Description

In chapter 2 the process concept was presented. However this did not reveal the details about the selected unit operations, which will be discussed in this chapter.

5.1 Propylene reactor

In the propylene reactor propane is converted to propylene by catalytic dehydrogenation. The design of the reactor is based on the configuration used in the Houdry Catofin process [15]. The reactor consists of a fixed bed containing a chromium oxide catalyst. Reactions take place at a low pressure (average 0.5 bara), selected to achieve an optimum between product selectivity and energy consumption. The reaction is largely endothermic. Three reactors are placed parallel for frequent regeneration of the catalyst bed. Burning of cokes in the regeneration cycle provides heat for the reaction; in this design fuel gas is led through the bed to heat it to the final temperature (different ways of providing the heat of reaction are proposed in several patents, see appendix 13). Every 7 minutes the bed needs to be regenerated. Large power-operated valves are placed to enable this swing operation under vacuum. Operation temperature of the dehydrogenation reaction is 650°C; under these conditions a high conversion of 90% can be achieved. Selectivity to propylene is 87.9% (see chapter 4 and appendix 5.1). The vessel designed to operate under vacuum can withstand an equivalent pressure of 3.6 bara.

5.2 IPA Reactor feed

Operating at 175°C and 34 bara the IPA reactor converts propylene to isopropyl alcohol. This exothermic reaction has a low conversion per pass of 5.7%, asking for large recycles around the reactor. The recycle is split in two: one going ~~around the~~ IPA reactor, the other flowing back to the propylene reactor. Split fraction was set to 0.9 in direction of the IPA reactor cycle (see appendix 8 for recycle mass flows against split factor 'alpha').

Kinetics for the production of IPA are rather complex. The catalyst in the fixed bed reactor consists of porous particles with H_3PO_4 acid deposited on them. The steam entering the reactor will condense in the particle, as a result some acid will dissolve in the condensed water. When the propylene also dissolves in the condensed water the reaction can take place.

5.3 Columns

The separation columns in the process are sieve plate columns. R101 effluent leaves the reactor at 650°C and 0.33 bara. To limit the loss of product through polymerisation the product gas pressure is raised to 36 bara. This pressure level is also required to achieve temperatures above –50°C in the partial condenser of C101.

Deethanizer column C101 removes hydrogen and light ends. Hydrogen must be removed from this stream since it is an explosive gas; leakage of hydrogen would invoke the reverse Joule-Thompson effect creating heat for explosion. Keeping hydrogen gas in the R102 feed stream would ask for the use of stronger vessel materials. It is required to separate the light ends from the propylene reactor product ~~is~~, since too much light alkenes in the IPA reactor feed will cause high levels of ethanol and di-ethylether in the product stream. Ethanol spec in IPA product should be less than 50 ppm. C4 and C6 alcohols in IPA product stream give trouble with the strict odour specifications.

Refrigeration duty at –33°C in the condenser of C101 is delivered by a heat pump. The closed heat pump cycle contains a propane/propylene mixture in the same ratio as the C101 bottom stream. A valve causes the stream of 34 bara of 40°C to drop adiabatically to 1 bara and –47°C, arriving at the right temperature level to deliver cooling duty. Compressor K103 brings this stream back at 34 bara, after which it is cooled from 140°C to 40°C with cooling water (E110).

Reactor effluent from the IPA reactor still contains vast amounts of propane and propylene. These components are separated from the IPA product stream in the washing column C102. This column functions as an absorber, a fixed bed in which water strips the stream from IPA product. Propane and propylene are recycled. Pressure difference over compressor K104 is 5 bar. The absorber should work at 29 bara and 30°C[2]

The separation section works under atmospheric conditions. Light ends column C201 operates at 1 bar and removes DIPE and NPA over the top. The bottom product <202> should contain 15% IPA in water. Therefore, a sufficient part of water stream <213> is recycled to the light ends column, which enters in the top (no condenser needed). A wet distillation is needed in order to meet odour specifications (higher alkanes would affect the specs). Aromatic compounds would leave the columns with DIPE and light ends over the top because of the high activity coefficient.

However water forms an azeotrope with IPA; water is removed in distillation column C202 such that the side stream contains 85% of IPA and 15% water (azeotropic condition). The azeotrope is separated with the aid of cyclohexane as entrainer in C203. The IPA product stream <222> contains maximum 0.05% NPA and 0.1% water. The reflux of the cyclohexane rich stream from vessel V203 counts 4 over feed.

5.4 Heat exchangers

Two types of heat exchangers were used: fixed tube sheets and floating head. According to guidelines [18] fixed tube sheets were placed for the lower temperature differences and lower pressures ($\Delta T<30°C$, $P<8$ bar), for other exchangers floating heads were used which allow higher pressures and temperature differences.

Air was selected as cooling medium if hot stream end temperatures exceeded 60°C. For other streams cooling water was used. As a design practice cooling water was located in the tubes (being the stream with biggest chance on fouling) because these can be easily removed and cleaned. Exceptions on this rule are the exchangers with the hydrocarbon streams at high pressures. Allocating the high pressures to the tube side is cheaper than high-pressure shells.

[2] However it should be noted that in the process simulation this absorber was modelled at a pressure of 1 bar. This causes the duty of compressor K104 to increase unrealistically. Therefore in the design of the equipment operating conditions were assumed that are indicated in appendix 12A (Recommended flow scheme).

5.5 Dryer

The recycle from washing column C102 contains water. This water should be removed since it would freeze in the condenser of C101 and could damage the catalyst in R101. Furthermore the risk of formation of gas hydrates should be minimised. Gas hydrates are ice structures containing light alkanes. Alcohols function as inhibitor for gas hydrate[3] formation. Gas hydrates are formed at temperatures below 10°C [16]. Dryer D101 is placed in the recycle stream to the propylene reactor in order to minimise the volume flow through the dryer bed. The dryer contains molecular sieves (zeolite) that absorb water. Two beds are placed parallel to allow frequent regeneration; heating it with burning fuel gas dries the bed.

5.6 Decanters, knockout drums

In the first section a high-pressure knockout drum is used[4] to separate the liquid phase from the vapour phase. This was done to minimise the load on the absorber.
Top stream from the light ends column C201 and the bottom from the NPA-recovery column C205 are fed to the three-phase-decanter (V201). In this vessel the ethers (gas), the NPA as dispersed phase and an inorganic as the continuous phase are separated. The last decanter is used to separate the azeotropic agent (cyclohexane) from the water phase. The azeotropic agent is sent back to the IPA drying column.

5.7 Pumps

All pumps are centrifugal pumps. These pumps are widely used in the process industry and can be used for the operating conditions of this process. In two cases a multi stage pump is necessary (P102 and P103) this is due to the large head, which has to be met; exact specifications to be determined by experts. The pump P203 used to transfer the NPA from decanter V201 has to be able to pump very small volumetric flow rates; this may be done through by-passing the pump.

5.8 Compressors

Besides the 'centrifugal compressor' two other types of compressors are used within the chemical industry. If a high capacity is desired an *axial flow compressor* is used, therefore compressor K104 in the recycle around the IPA reactor has been chosen to be such a compressor (ΔP = 5 bar). There is a big pressure change between the two reactors. Another constraint is the vacuum, which is desired within the propylene reactor section. A normal centrifugal compressor cannot obtain this. For this reason the pressure difference is achieved by installing a *reciprocal compressor* to create the vacuum. A second *multi stage centrifugal compressor* produces the large increase in pressure from 0.57 to 36 bar within three stages (pressure ratio of 4).

5.9 Process Flow Scheme (PFS)

Appendix 1 contains the complete process flow scheme.

Description of UNIT 100, REACTION SECTION

The propane feedstock enters the process at 15 bara in stream <101>. After letting the pressure off, it is mixed with the so-called overall-recycle, stream <134>. It is then

[3]

[4] Note that in the PFS the vessel is a flash vessel, but due to the improved configuration a decision was made to insert a knock out drum at high pressure. See appendix 12A 'Recommended process flow scheme'

warmed up by the reactor effluent, stream <108>, and heated in the furnace (F101). It enters the dehydrogenation reactor (R101), where it is converted to propylene. (R101) consists of 3 reactors, of which one is used, one is being regenerated and one is stand-by. After being cooled by the aforementioned feed-product heat exchange, the stream <109> is brought up to 0.67 bara by compressor (K101), and again to 36 bara by compressor (K102). After both compressors, cooling is used. Stream <114> enters the deethanizer (C101), where the light ends are removed. The bottom stream <118> is mixed with some make-up water and the IPA-recycle stream <132>, which is to be discussed later. The resulting stream <121> is, after heating, conducted to the hydration reactor (R102), where propylene is converted to isopropyl alcohol (IPA). In the high pressure separator (V102), most of the C3s are removed; the remaining C3s are stripped off in the washing column (C102) with the water stream <129>, which has it origin in Unit 200 and will be discussed there. The top of C102, stream <130> which contains mainly unconverted propylene and some unconverted propane, is split in two: one part to the overall-recycle (stream <133>, which is dried over a zeolite bed before it is mixed with the feed), and the greater part to the so-called IPA-recycle, stream <131>. This stream is compressed and mixed with the C101 effluent. The bottom of column C102 and separator V102, stream <134>, is led to Unit 200, the separation and purification section.

Description of UNIT 200, SEPARATION AND PURIFICATION SECTION

Stream <137>, consisting of impure IPA and water, is fed to the light ends column (C201). The stream <203>, which leaves the top, is mixed with stream <234>, which leaves the bottom of the NPA recovery column (C205) and contains mainly n-propylalcohol (NPA). The resulting stream <204> is completely condensed and fed to the phase separator vessel (V201). Light ends and ethers are removed as a gas, stream <206>; NPA is tapped off as the organic phase in stream <207; and the aqueous phase with IPA in it is mixed up with the feed of column (C201).
The bottom stream <202> of column (C201) enters the IPA CBM column (C202). This column has as a top product some light ends, ethers and NPA in stream <212>, which is fed to the NPA recovery column (C205), which is to be discussed later. The bottom stream <213> consists of pure water and is split in three: a part is led back to column (C102), another part to the IPA reactor (R102) and the remaining part is used as reflux in column (C201). The side stream <209> of column (C202) consists of an almost azeotropic mixture of IPA and water and gives the column its name (CBM = Constant Boiling Mixture). It is mixed with a make-up stream of cyclohexane <219> and enters the last separation step.
This combined stream <220> enters the IPA drying column, where the IPA/water azeotrope is broken with the aid of an azeotropic agent, cyclohexane. The product IPA is recovered at 99.9% purity at the bottom of this column, in stream <221> which becomes, after cooling, stream <222> which leads to the storage tanks. The top stream <223> is condensed and fed to the phase separation vessel (V203), where cyclohexane with some IPA in it is recovered as the organic phase (stream <224>) and water with some IPA in it as the aqueous phase (stream <225>). The former stream is used as the reflux for column (C203); the latter is fed to the cyclohexane recovery column (C204), where almost clean water leaves the bottom in stream <229> and is led to the waste treatment to have the remaining organic impurities removed. Stream <228> leaves the top of (C204) and is mixed with the feed of the (C203) column.

Finally, the aforementioned NPA recovery column (C205) essentially separates NPA (bottom) and IPA (top), which are led back respectively in stream <234> to separator (V201) and in stream <233> to column (C201).

5.10 Process Stream Summary

The process stream summary is indispensable for a good understanding of the process flows. It indicates pressure, temperature, enthalpy, phase and composition for each stream. The stream summary can be found in appendix 2.

5.11 Heat integration

Heat integration is usually performed with the main process streams. The diagram in Appendix 10.1 shows a graphical representation of the process streams that have to be cooled or heated.

Only the two feed streams to the reactors need heating. According to practice it seems plausible to place product-feed exchangers ([ref [6], [18]). However to heat the R101 feed stream <103> from 5°C to 650°C in one exchanger would ask for expensive equipment because of the high end temperature and large temperature difference. Therefore medium pressure steam of 15 bar and 300°C is produced by exchanging the R101 product stream <107> with boiling feed water. This stream is then further exchanged with stream <103>. A furnace is required to bring the R101 feed to the entrance temperature of 650°C (see process flow scheme). This also improves temperature control around the reactor.

The use of a product-feed exchanger around reactor R102 proved to be unfavourable during the design, since the small temperature difference asks for a very large heat exchange area and thus expensive equipment. It was decided to heat the reactor feed <121> with hot utilities.

To make use of further available heat in the process streams, exchangers E107 and E113 were placed to form the reboilers of respectively C203 and C201. Evaluation of the heat integration regarding control options and pinch analysis follows in respectively chapter 6 and 7.

5.12 Utilities

Major consumers of utilities in the process are the compressors (K101 to 104), reboilers of the columns (e.g. the reboiler for C202 asking for 58MW) and the coolers. Large quantities of air and cooling water are used for cooling process streams and providing the necessary duties for the condensers.

Compressors K101 to 103 use electricity, under the assumption that power is sufficiently available (investment is included in cost). Compressor K104 however is driven by MP steam of 15 bar, that is produced in the plant (E105). After being used for the compressor this steam exits as low-pressure steam on 5 bar and is employed for heating of the reboilers of columns C101 and C204; 0.38 t/h is left over and appears as net steam production.

Fuel gas is used for the regeneration of the catalyst in the propylene reactor and for drying the fixed bed of dryer D101.

Optimisation of operating conditions and connections between unit operations could reduce temperature differences and thus cooling and heating duties.

5.13 Process Yields

Tables below present the yields of product, feed, by-products, process chemicals and utilities. Appendix 4.2 shows the results in a block diagram.

Table 25, process yields of the streams

Process Yields - Streams							
Name	Ref.	kg/s		t/h		t/t IPA Product	
	Stream	IN	OUT	IN	OUT	IN	OUT
Propane feed	<101>	4.85		17.47		0.84	
Make up water	<119>	2.97		10.70		0.51	
Make up cyclohexane	<219>	0.01		0.02		0.001	
Light ends from C101	<116>		0.53		1.90		0.09
Vent	<206>		0.13		0.45		0.02
To NPA storage	<207>		0.01		0.02		0.001
IPA product stream	<222>		5.78		20.81		1.00
To water treatment	<230>		1.00		3.58		0.17
Cokes	<001>		0.12		0.43		0.02
Water effluent	<002>		0.22		0.80		0.04
Total		7.830	7.778	28.188	28.000	1.3543	1.3453

-0,052

Table 26, process yields of the utilities

Process Yields – Utilities							
Name	Ref. Stream	kg/s	kW	t/h	kWh/h	t/t IPA Product	kWh/t IPA Prod.
LP Steam (1)	-	-0.11		-0.38		-0.02	
MP Steam	-	6.80		24.48		1.18	
HP Steam	-	50.00		180.00		8.65	
Fuel Gas	-	0.70		2.51		0.12	
Cooling Water	-	2045.86		7365.11		353.92	
Air	-	4321.90		15558.84		747.66	
Electricity	-		17954.72		17954.72		862.81
Remarks: (1) LP steam is net produced in the process							

The yield with respect to propane feed of 0.84 t/t is higher than the reported value. 0.82 t/t ref [5]

6 Process Control

In this chapter, the basic control of the process will be described.

6.1 Basic controls

Some basic control that appears at several units is described below; exceptions will be mentioned in the text.

The temperature of the outgoing stream of a heater/cooler is controlled by a TC controlling the valve of the heating/cooling medium.
The pressure control after compressors and pumps is usually done by varying the bypass over the unit.

The level control of vessels is done by controlling the valve after the discharge pump. With a two-liquid-phases separation vessel, the level of the lightest fluid, which flows over the baffle, is controlled by a level control as described above. The interface level, and thus the level of the heaviest fluid, is controlled by an interface controller connected to a valve after the other pump.

The control around most distillation columns has the following basic configuration:
top stream condenser temperature control over cooler
reflux flow control over the reflux (left) pump
accumulator level control over the remaining (right) pump
reboiler temperature control over the reboiler heater
liquid level control after the discharging pump.

6.2 Following the process streams

The feed stream is preheated by exchanging heat with the product stream, in heaters E101 through E104. The temperature is controlled by letting a TC control the bypass of E104.
The stream is then heated in the furnace F101. The outgoing temperature is kept constant by a TC controlling the furnace fuel and air feed.
There are two types of control in the reactors R101A/B/C. In the reactor that is on stream, the pressure has to be controlled. This is done by controlling the compressor K101. The reactor that is being regenerated needs a temperature control by controlling the inflow of air and fuel gas.

The pressure after compressor K102 is controlled by K102s bypass.
At the top section of column C102, there is an exception to the standard configuration: while there is only one pump, reflux as well as accumulator liquid level will have to be controlled by it. To flatten out fluctuations to ensure a smooth operation, a cascade control with a TC- and a FC- control is used here.

At reactor 102, the risk of a runaway reaction exists. There is a temperature control at heater E112, but would a runaway occur, it can be quenched with water by letting a TC open the valve after P102.

Arriving at Unit 200, we see a reboiler temperature control not by varying a heating medium flow but by varying the bypass over the heat exchanger E113.
The liquid level of column C202 is controlled by a valve in stream <218>, because the flow controls in streams <214> and <216> have the priority: first, a sufficient stream must be led to C102 respectively C201; then, with the remaining stream, the level can be controlled.

As with C201, the reboiler temperature of C203 is controlled by the bypass over a heat exchanger.
From now, control is straightforward, as described under "Basic Controls".

6.3 Explanation of process streams-reboiler connections

Exchangers E107 and E113 were placed to form the reboilers of respectively C203 and C201. Although this is usually unfavourable in terms of process control, measures can be taken to avoid problems. Changing cooling duties of E106 and E108, in order to secure enough energy for the reboilers of C203, could control temperature differences in stream <112>. However temperature allowances for compressor K102 limit this control loop. Placing a by-pass over E113 could provide control over temperature differences in R102 effluent, so that the reboiler of C201 runs on constant duty. Normally heat exchangers for steam would be added in the reboiler cycle to complement required duty.

7 Mass and Heat Balances

In this chapter the mass and heat balances are presented. For mass and heat flows that cross the battery limit, in- and outgoing amounts should be equal. Assessing the mass and heat balance forms a good evaluation tool for the design. This chapter also demonstrates an analysis of the heat integration possibilities in the process.

7.1 Overall mass and heat balance

Table 28 shows the overall mass and heat balance. Data for the streams is taken from the Process Stream Summary, appendix 2.

Table 27, overall mass and heat balance

Overall mass and heat balance					
		Mass		Heat	
Name	Ref.	kg/s		kW	
	Stream	IN	OUT	IN	OUT
Propane feed	<101>	4.85		13160.49	
Make up water	<119>	2.97		35537.36	
Make up cyclohexane	<219>	0.01		10.21	
Light ends from C101	<116>		0.53		763.25
Vent	<206>		0.13		143.91
To NPA storage	<207>		0.01		17.02
IPA product stream	<222>		5.79		34,980.91
To water treatment	<230>		1.00		15,795.21
Cokes	<001>		0.12		6,460.86
Water Effluent	<002>		0.22		2,979.72
Hot utility				152,189.59	
Cold Utility					223,682.80
Power (1)				16,021.51	
TOTAL		7.83	7.79	216,919.16	284,823.68
Error [%]			0.55	23.84	
Remarks: (1) Includes theoretical duties of pumps and compressors					

7.2 Mass balance

The overall mass balance indicates a small deviation of 0.55%. This is due to reduced convergence accuracy that was required to run the simulation in Aspen. Maximum relative error that was allowed counted 1%.

7.3 Heat balance

In the heat balance a rather large error of 23.84% is encountered. Three possible reasons for this deviation can be found. First is that the separation columns C203 and C204 were simulated as ideal separators (no adequate thermodynamic models were available), resulting in deviations in enthalpy of the streams. Second cause of errors in the heat balance could be found in the fact that column C201 was simulated separately (correction of feed stages and way of reflux). Third is that the flow scheme was split into three sections in order to complete the simulation. Rounding errors in the input of the connecting streams may have contributed to the deviation. Exact reasons should be closer investigated during future design.

7.4 Component mass and heat balance

An overall component mass balance and stream heat balance is included in the appendix (appendix 4).

7.5 Heat integration

A pinch analysis follows to evaluate design decisions of the heat integration (see chapter 5). Pinch technology [Linhoff et al 17, p.100] is used for the efficient design of heat exchanger networks and can lead to reduction in energy requirements. Reboilers and condensers were left out of the calculation of the pinch. Table 29 shows the hot and cold streams ~~that~~ were taken into account.

which

Table 28, hot and cold streams used for heat integration analysis

Hot streams				Cold streams			
Stream nr.	Duty [kW]	T (in) [°C]	T (out) [°C]	Stream nr.	Duty [kW]	T (in) [°C]	T (out) [°C]
<107>	21677	650	100	<103>	24097	5	650
(2)<112>	7655	180	30	<121>	14061	140	160
(2)<123>	46025	180	90				
<126>	39427	90	30				
(2)<216>	6809	90	30				
(2)<204>	2615	80	30				
<222>	4765	80	30				
Remarks: (1) Stream <229> left out of discussion because of small duty (2) Approximation of T-intervals for easy calculation							

From this data composite curves can be calculated and plotted on a temperature-enthalpy diagram (appendix 10.2). The cold composite curve is shifted on the enthalpy axis below the hot composite with a minimum temperature difference of 10°C.

Hot utility requirements are relatively small, 1185 kW, but cold utilities should at least provide 92000 kW of cooling duty. From this diagram the Grand Composite curve can be drawn (appendix 10.3). The pinch temperature is found to be 145°C.

The grand composite shows that all cold streams between 650 °C and 145°C can be heated with process streams. However if heating capacity is used to form MP steam of 300°C, additive hot utilities are required (±12500 kW).

Evaluation with pinch analysis demonstrates that heat flows across the pinch in exchanger E101; this will cause the consumption of utilities to be greater than the minimum values that could be achieved. An improvement could be to increase the duties of furnace F101 and E105, respectively to heat stream <104> and cool stream <107>.

8 Process equipment Design

In this chapter the use of the simulation tools will be discussed and explained. Simulation problems and calculations concerning the equipment from the PFS will be presented. The 'Design Specification Sheets' (and their summary) can be found in appendix 6. In this appendix the results of the simulation and sizing calculations per piece of equipment are summarised.

8.1 Integration by Process Simulation

The process is simulated with the process flow sheet simulator ASPEN+ version 10. ASPEN combines a graphical flow sheet interface with thermodynamic model calculations to simulate the different process units and streams. For other simple calculations the spreadsheet program Excel was used (MS Office2000).

8.2 Special modelling issues

As mentioned in chapter 4, two equilibrium reactions occur within the propylene reactor. Both equilibrium reactions are simulated in Aspen using two equilibrium reactors. The equilibrium reaction, in which ethylene is converted to ethane, cannot be simulated without taking into account the irreversible reactions for the production of ethylene and hydrogen. (5) & (3)

Schematically this looks like:

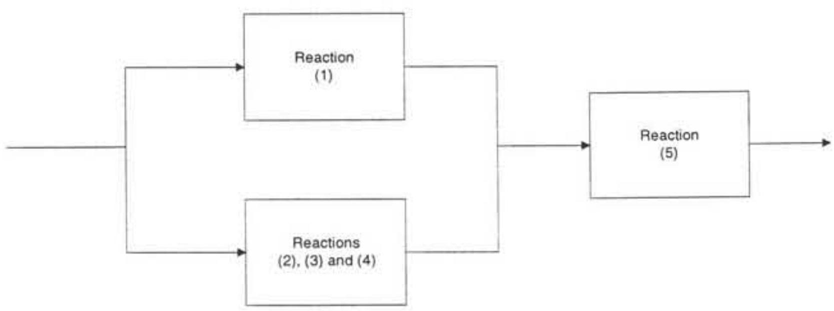


Figure 4, block diagram of the simulated propylene reactor.

The reactors were first simulated based on a pure propane feed. By adjusting the available simulation parameters, the yields were obtained as calculated in chapter 4. The equilibrium constants used for the equilibrium reactions are calculated by Aspen using the Gibbs energies at a specified temperature. A change in temperature at which the constant is calculated, results in a change in of the conversion.

Within the purification section (Unit 200), a ~~lot of~~ many ideal separation blocks have been used. This was due to problems with the thermodynamic models, as mentioned in chapter 4.

Because of the large numbers of circular calculations (due to the recycles) ASPEN had difficulties converging. The decision was made to split the total process up into three sections.

Section 1; up to the absorber
Section 2; columns C201, C202 and C205
Section 3; columns C203 and C204

8.3 Equipment Selection and Design

Reactors

Propylene reactor (R101)

For the design of the propylene reactor assumptions had to be made and these are stated in table ~~30~~.

Table 29, Assumptions made for the design of the propylene reactor

Assumptions	Ground for assumption
d_p = 5.3 mm	Same as IPA-catalyst; Rob Berger[18] says this is a common diameter.
Cat. Cost: 20 DM/kg	Educated guess after consulting Rob Berger ~~[20]~~ [18]
Void fraction = 0.4	Common value; same as in IPA reactor
GSHV = 4000 m^3 gas/m^3 cat/hr	Interpreted from fig.2, article of EH Stitts[13] m^3 cat incl void.
Viscosity of gas in reactor = 2e-5 Ns/m^2	This is the average value of the viscosity's of propylene and propane at 1 bara and 500 °C.

The Gas Hourly Space Velocity enables one to calculate the amount of catalyst needed for the volumetric rate of the gas entering the reactor.

$$m^3_{cat.required} = \frac{Volumerate}{GHSV} \qquad [8.1]$$

The amount of catalyst determines the required reactor volume. Reactor dimensions can then be determined. Usually L/D equals 4, however the propylene reactor operates at sub atmospheric conditions and therefore a maximum pressure drop of 0.33 bara is allowed. So the pressure drop then determines the height of the bed. Ergun's equation[19] gives the maximum height of the catalyst bed:

$$\frac{\Delta p}{\rho L} = \frac{\upsilon^2_0}{d_p}\frac{1-\varepsilon}{\varepsilon}(170\frac{\nu}{\upsilon_0 d_p}(1-\varepsilon)+1.75) \qquad [8.2]$$

Once L (= height of the catalyst bed) has been determined, the diameter of the catalyst bed can be calculated, which is also the diameter of the reactor. L/D for the propylene reactor equals 1.3 instead of 4. This is not unusual for propane dehydrogenation reactors, which are reported to have large cross sectional areas.[15]. The results are summarised in appendix 6, in the 'Design Specification Sheets'

IPA reactor
The design of the IPA reactor has been done according to the IPA reactor, operated by Shell on their Pernis site. The GHSV of the Pernis IPA reactor has been calculated and this GHSV has been used to determine the dimensions of this IPA reactor.

In Aspen a stoichiometric reactor models the IPA reactor, where the reactions occur in series. All reactions and conversion of propylene are known. Selectivity for the main (product forming) reaction is also known. Catalyst deactivation and coke deposition have not been modelled because little literature data was available and because of time limitation.

The dimensions of the propylene reactor were calculated with the same method used to calculate the dimensions of the propylene. For the IPA reactor L/D has been set to 4, which is a common value. For comparison, the L/D of the Shell IPA reactor equals 3.67. The results are summarised in appendix 6 'Design Specification Sheets'.

Distillation columns

In this section the procedures will be discussed as they have been used for design of the ~~distillation~~ columns. An example calculation and the results can be found in appendix 8. For the symbols used, see the 'List of symbols' at the end of the report.

Design considerations
The operating range of a distillation column is determined by weeping, flooding and entrainment. These three parameters are used in the design procedure[18, chapter 11], in which the following steps are taken:

- Determining the diameter
- Determining the interior
- Check for weeping
- Determine pressure drop per plate
- Check flooding/downcomer
- Check entrainment

External design
A column is normally divided into two sections, the top- or rectifying section and the bottom- or stripping section (with respect to the feed stage). As a consequence the overall-design exists of the stripper design and the rectifying design. The procedure and relationships are the same for both sections.

The first step is to determine the liquid-vapour flow factor[18], F_{LV} , see equation 8.3

$$F_{LV} = \frac{\phi_{mL}}{\phi_{mV}} \cdot \sqrt{\frac{\rho_V}{\rho_L}} \qquad [8.3]$$

Using this liquid-vapour flow factor and the chosen plate spacing (t_s), K_1 can be obtained[18, fig. 11.27]. This constant is used to determine the flooding velocity[18, p509], see equation 8.4.

$$u_f = K_1 * \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \qquad [8.4]$$

The upper limit of the vapour velocity determines the column diameter. In this case 90 % of the flooding velocity is taken as design value.

To calculate the column diameter an estimate of the net area is required. This follows easily from the mass flow and the design value for flooding velocity calculated above. The net area. A_{net} , must be corrected for the downcomer area in the column. Typically, 15 % of the total area is taken as initial design value. This leads to the following column diameter:

$$D_{col} = \sqrt{\frac{4}{\pi} \cdot A_{col}} \quad [8.5]$$

with:

$$A_{col} = \frac{A_{net}}{0.85} \quad [8.6]$$

The number of theoretical plates has to be corrected for plate inefficiencies, due to the fact that the thermodynamic equilibrium is not completely reached. This is corrected using the column efficiency, see equation 8.7.

$$N_{actual} = \frac{N_{theoretical}}{E_{col}} \quad [8.7]$$

The length of the column is estimated using equation 8.8

$$L_{col} = (N_{actual} - 1) \cdot t_s \cdot 1.15 \quad [8.8]$$

The next step is the design of a provisional plate. Some methods to estimate parameters like weir dimensions, active area, hole size, number of holes and the hole pitch are given in [18, sec. 11.13].
With the results calculated and estimated above one still has to justify the practical operability of the column. At first weeping is considered, then plate pressure drop, downcomer back-up and finally entrainment is checked.

Weeping
The lower limit of the operating range occurs when liquid leakage through the plate holes becomes excessive. The minimum design vapour velocity through the holes is given by equation[18, p513] 8.9:

$$u_{h,min} = \frac{(K_2 - 0.9 * (25.4 - D_h))}{\sqrt{\rho_V}} \quad [8.9]$$

The hole diameter, d_h ,is chosen arbitrarily in the provisional plate design. K_2 is a constant, depending on the depth of clear liquid on the plate [18, fig. 11.30].
The clear liquid depth equals the weir height h_w plus the crest of liquid over the weir h_{ow}. The latter can be estimated using the Francis weir formula for segmental downcomers:

$$h_{ow} = 750 * \left(\frac{\phi_{mL}}{\rho_L * l_w} \right)^{\frac{2}{3}} \quad [8.10]$$

The weir length, l_w, is estimated from [18, p515]. Weeping will not occur if vapour velocity exceeds the required minimum vapour velocity. Otherwise the provisional plate design has to be adjusted.

Plate pressure drop

The total plate pressure drop h_t, expressed in mm liquid, is as follows:

$$h_t = h_{ow} + h_w + h_d + h_r \quad [8.11]$$

The dry plate pressure drop, h_d, is given by:

$$h_d = 51 * \frac{\rho_V}{\rho_L} * \left(\frac{u_h}{C_o} \right)^2 \quad [8.12]$$

C_o is called the orifice coefficient and is a plate property; u_h is the hole velocity. The residual head, h_r, is given by:

$$h_r = \frac{12500}{\rho_L} \quad [8.13]$$

Downcomer

The level of liquid and froth in the downcomer should be well below the top of the outlet weir on the plate above to prevent flooding. The downcomer liquid back up (h_b) is defined as follows:

$$h_b = h_{ow} + h_w + h_t + h_{dc} \quad [8.14]$$

The head loss in the downcomer, h_{dc}, results from resistance to flow in the downcomer and can be estimated using the following equation:

$$h_{dc} = 166 * \left(\frac{\phi_{m.L}}{\rho_L * A_{ap}} \right)^2 \quad [8.15]$$

To account for the froth density in the downcomer, the liquid back up (h_b), calculated from equation 8.14, should satisfy the following condition:

$$h_b < 0.5 * (t_s + h_w) \quad [8.16]$$

Finally, the residence time in the downcomer must be sufficient for the entrained vapour to disengage from the liquid stream. A residence time of at least of 3 seconds is recommended. The residence time is given by:

$$t_s = \frac{A_d * h_b * \rho_L}{\phi_{m.L}} \quad [8.17]$$

Entrainment check

With the entrainment check the amount of liquid flowing up with the vapour is calculated. The flow factor F_{LV} is correlated with the fractional entrainment and the percentage flooding. Typically, the fractional entrainment should fall below 0.1 in order to neglect entrainment effects on plate efficiency.

A special case was the design of the distillation column C204, which could not be designed in an ordinary way because no thermodynamic model was available. Data form a similar column, was provided by Shell, and can be found in table 31.

Table 30, data from a similar column as C204

Feed	(kg/hr)	2000
L	(m)	19
D	(m)	0.9
A		0.64
R	(m3/hr)	5.2

The ratio of the provided feed and our feed was used to determine the surface area needed in our column. The length is kept the same. Using the same equations as used for the other columns an estimation was made for the column dimensions.

The general results of the column design are listed in table 32. A more detailed result can be found in appendix 6 'Design Specification Sheets'.

In the design of the columns weeping almost occurred; due to this the minimum vapour velocity is high. Columns were modelled using sieve plates, in future design it is recommended to use valve plates.

Table 31, summary column design

		C101	C201	C202	C203	C204	C205
Diameter							
-Top	[m]	2.34	2.67	2.51	2.60	1.34	2.68
-Bottom	[m]	2.61	3.03	2.84	n.a.	n.a.	2.68
Column height	[m]	31.05	31.05	39.68	23.29	19.00	18.98
Plate spacing	[m]	0.60	27.00	0.60	0.45	0.45	0.60
Downcomer area	[m2]	0.80	0.58	0.95	0.80	n.a.	0.85
Total plate pressure drop							
-Top	[Pa]	1168	2368	1122	1135	1000	1192
-Bottom	[Pa]	1160	2301	1776	n.a.	1000	1192
Number of trays (actual)	[-]	45	45	58	45	35	28

Absorber

Sizing an absorber starts similar to sizing a distillation column. The F_{lv} is determined, see equation 8.3. Using this liquid-vapour flow factor two constants are determined[18, fig.11.44], K_4 at flooding and K_4 at a certain pressure drop per meter packed height. Using equation 8.18 the gas mass flow-rate per unit column cross-sectional area can be calculated.

$$V_W^* = \left[\frac{K_4 \cdot \rho_v \cdot (\rho_L - \rho_v)}{13.1 \cdot F_p \cdot \left(\mu_L / \rho_L \right)^{0.1}} \right]^{1/2} \qquad [8.\ 18]$$

V_w^* = gas mass flow - rate per unit column cross - sectional area, kg/m^2s

F_p = packing factor, m^{-1}

μ_L = liquid viscosity, Ns/m^2

Dividing the actual mass flow rate through the V^*_w will result in the required area, and thus lead to the column diameter. This is summarised in equation 8.19.

$$D_{absorber} = \sqrt{\frac{4}{\pi} \cdot \frac{\phi_{m,gas}}{V_w^*}} \qquad [8.\ 19]$$

The length of the column depends on the packing used in the column. To obtain a reasonable column diameter, a structured packing was chosen (Mellapak). For an absorber with structured packings, the length of the column can be calculated using equation 8.20.

$$L_{absorber} = \mathrm{N}_{\mathrm{theoretical}} \cdot HETP \cdot 1.15 \qquad [8.\ 20]$$

The HETP will depend on the packing, but a conservative estimate is said[18, p538] to be 0.5m. This results in the figures as summarised in table 32, for a more detailed summary can be found in appendix 6 'Design Specification Sheets'.

Table 32, summary of the general results of the absorber

		C102
Diameter	[m]	2.78
Length	[m]	8.63
Packing	[-]	Mellapak 250_x
Theoretical plates	[-]	15

Dryer column

The dryer has been designed using similar calculations used to calculate the dimensions of both reactors. The following assumptions were made:

- The zeolite used can absorb 15 wt-% water.
- The zeolite is regenerated to 3 wt-% water.
- The zeolite bed has a weight of 50 tons.
- Density of the zeolite is 750 kg/m^3

The amount of water, which has to be removed was taken from the stream summary, appendix 2, stream <002>. The weight of the zeolite bed then determines the on-stream time of the first dryer, before the second dryer has to take over and the first dryer can be regenerated. Regeneration has to take place within the on-stream time of the second dryer. The regeneration time for each dryer was set to 6 hours, leaving two hours for other activities or allowing the dryer to be put on-stream earlier in case more water has to be removed. Theoretically, a dryer should be on-stream for 8 hours.

Because the weight and density of the catalyst have been set, the volume of the zeolite bed and thus the size of the dryer are easily calculated. L/D has been set to 4. For a summary of the results see appendix 6.

Vessels

The phase separator and accumulator vessels have been designed in the following way:
horizontal cylindrical vessels were used because of the large volume flows
the relation L/D was taken at 4
the liquid level was supposed to be at half the vessel height
it was demanded that the pumps after the vessel would not run dry for 15 minutes when the supply would stop; this 15 minutes is the required holdup time for half the vessel volume.

From these data and the throughput, the volume, L and D of the vessel were calculated:

$$V\ [m^3] = 2 * (\text{throughput}\ [m^3/\text{min}]) * (\text{holdup time}\ [\text{min}])$$

Now that the volume is known, we can obtain the length and diameter of the vessel – at the given L/D=4 – by using the Goal Seeker in Excel.

Exceptions made to above guidelines:

- for V102, L/D=2.5 was taken because of the large gas flow which needs a large area rather than a large volume
- for V101, V202 and V205 (all accumulators), the calculated volumes were too big; a required holdup of 5 instead of 15 minutes was taken. As a consequence, tripping systems will have to be installed on pumps P101, P205 and P215.

Heat exchangers, reboilers, condensers

Heat exchangers were designed according to Coulson & Richardson [18,chapter 12]. Results can be found in the design sheets, appendix 6.2.

The general equation for heat transfer across a surface is:

$$Q = UA\Delta T_m \qquad [8.21]$$

The logarithmic mean temperature is calculated from the temperature differences between the hot and cold medium:

$$\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \dfrac{(T_1 - t_2)}{(T_2 - t_1)}} \qquad [8.22]$$

To find the true mean temperature difference, the logarithmic mean temperature is multiplied with the temperature correction factor (F_t).

$$\Delta T_m = F_t \Delta T_{lm} \qquad [8.23]$$

This factor F_t is correlated as a function of two dimensionless temperature ratios, R and S.

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} \qquad [8.24]$$

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} \qquad [8.25]$$

Graphs were used to determine F_t [18, fig. 12.19 and 12.20]. Number of tube and shell passes were chosen according to the results of ratios R and S.
Heat duties for the exchangers were taken from process simulation. Average heat capacity for the streams was calculated with:

$$Q = \phi_M c_P (T_{out} - T_{in}) \qquad [8.26]$$

Overall heat transfer coefficients U were estimated with tabulated typical values [18, table 12.1]. Necessary heat transfer areas were calculated by completing formula 8.21 with Q, ΔT_m and estimated U. The results are summarised in appendix 6 'Design Specification Sheets'.

Furnace

For the furnace design certain assumptions have been made:

- 80% of the combustion heat is transferred into the radiant section
- No convection takes place, so pressure drop can be neglected
- the burner is fed with fuel gas and 20% excess of air
- Thermal efficiency is 80%. (This is a common value, the thermal efficiency for modern heaters lies between 80 and 90%)
- A heat flux of 30 kW/m^2 is used in order to make an estimate of the required tube area
- The heat of combustion of the fuel gas is 40 MJ/kg

The tube area is calculated with the next equation:

$$\mathrm{A}_{tubes} = \frac{\Delta \mathrm{H}_{flow}}{Q_{rad}} \qquad [8.\ 27]$$

The number of tubes can now be calculated:

$$\mathrm{N}_{tubes} = \frac{\Phi_{v,gas}}{\mathrm{A}_{\perp} v_{gas}} \qquad [8.\ 28]$$

Tube sizes normally used are between 75 and 150 mm diameter, 150 mm is assumed. Typical tube velocities are 1 to 2 m/s, 1.5 m/s will be taken.
So, the tube pitch is 150 mm. The length of the tubes can be calculated. Next, the number of tube passes can be calculated, so that the length and width of the furnace chamber are of the same magnitude.

The amount of fuel gas required is then calculated:

$$\Phi_{m,fuel} = \frac{\Delta H_{flow}}{0.8 * \Delta H_{comb}} \qquad [8.29]$$

The results are summarised in appendix 6, in the 'Design Specification Sheets'

Compressors

Within the chemical industry three types of compressors are normally used. The choice will depend on the capacity and compression ratio. The three types:

- Reciprocal compressor
- Centrifugal compressor
- Axial flow compressor

The operating window of the reciprocating compressor enables it to be used at sub atmospheric conditions. The axial compressor is able to handle large volumetric flow-rates. The exact range of each type can be found in lit. [6 &18, fig. 10.60].

In order to determine the number of stages(n) of the compressor, the following constraint has to be met:

$$\left(\frac{P_{out}}{P_{in}}\right)^{1/n} \langle 5 \qquad [8.30]$$

In table 33 a summary is given of the chosen compressors. The results also are summarised in appendix 6, in the 'Design Specification Sheets'

Table 33, summary of the compressors

Compressor number	Type[5]	Number of stage	Compression ratio stage	Efficiency [%]	Work done [kW]
K01	Reciprocal	1	3.35	72	2400
K02	Centrifugal	3	3.98	72	8015
K03	Centrifugal			72	4488
K04	Axial flow	1	1.17	72	900

Pumps

With capacities that range from 0.25 up to 1000 m^3/h and the typical head is 10 up to 50 m of water (per stage), centrifugal pumps will normally be the first choice for

[5] The reasons for these choices can be found in chapter 5.

pumping process fluids. The power required for pumping an incompressible fluid is given by[18, p427]:

$$Power = \frac{\Delta P Q_p}{\eta_p} \times 100 \qquad [8.\ 31]$$

ΔP = pressure difference, Pa

Q_p = flow rate, m^3/s

η_p = pump efficiency, --

The pump efficiency will depend on the capacity [18, fig. 10.62].

The transport of a bottom product is not possible without a bottom pump. For the design of these pumps a pressure drop of two bar is assumed. Most pumps are simple, standard centrifugal pumps, except for pumps P102, P103 and P203. Pumps P102 and P103 have to be multi stage centrifugal pumps to meet the desired head. The volume flow rate of P203 is very small which will become a problem for a normal centrifugal pump. For a summary of the pumps see table 34 and for a more detailed report see appendix 6.

Table 34, summary of the chosen pumps

Pump number	P_{in}	P_{out}	ϕ_v	η_p	Theoretical Power	Actual Power
	[Bara]	[Bara]	[m^3/h]	[%]	[kW]	[kW]
P101	35.00	36.8	99.36	77%	5.52	7.17
P102	1.00	34.0	31.24	70%	28.64	40.91
P103	1.00	29.00	94.20	76%	73.26	96.40
P201	1.40	3.39	418.48	85%	23.25	27.35
P202	1.00	3.00	72.80	75%	4.04	5.39
P203	1.00	3.00	0.02	40%	0.00	0.00
P204	1.50	3.50	375.96	84%	20.89	24.87
P205	0.70	2.66	219.72	80%	12.21	15.26
P206	0.70	2.66	54.93	75%	3.05	4.07
P207	0.7	2.67	32.42	70%	1.80	2.57
P208	1.30	3.25	27.41	70%	1.52	2.18
P209	1.00	3.00	4.45	50%	0.25	0.49
P210	1.00	3.00	131.46	80%	7.30	9.13
P211	1.20	3.17	3.60	45%	0.20	0.44
P212	0.80	2.82	3.96	45%	0.22	0.49
P213	0.80	2.82	0.86	40%	0.05	0.12
P214	1.10	3.11	28.68	70%	1.59	2.28
P215	0.80	2.77	52.58	75%	2.92	3.89
P216	0.80	2.77	26.29	70%	1.46	2.09

Storage tanks

IPA is produced continuously and is then transported with ships. This implies the need for an IPA storage facility. IPA in stream <222> first enters a run down tank where it's specifications are analysed. If the IPA is below the required specifications it is pumped into the off-spec tank, where it is stored until it can be mixed with IPA which is above specifications. In this way IPA of a constant quality can be sold. The main storage tank is capable of storing a quarter of the annual IPA production (= 44 kton).
The assumption has been made that 2 % of the annual production will be off spec. Analysis of the IPA in a run down tank takes a 24 hours. Therefore, four run down tanks are required to be able to store IPA on a continuous basis.

Table 35, Storage tanks and their dimensions

	Run down tank	Off spec. tank	On spec. tank	Sales tank
Number of tanks	4	1	1	1
Max. V_{IPA} [m^3]	203	4901	54618	~~25618~~
V_{tank} [m^3]	224	4901	~~60080~~	28008
L [m]	4.68	14.51	27.27	21.1481
D [m]	7.80	20.74	52.96	41.06
L/D	0.60	0.70	0.52	0.52

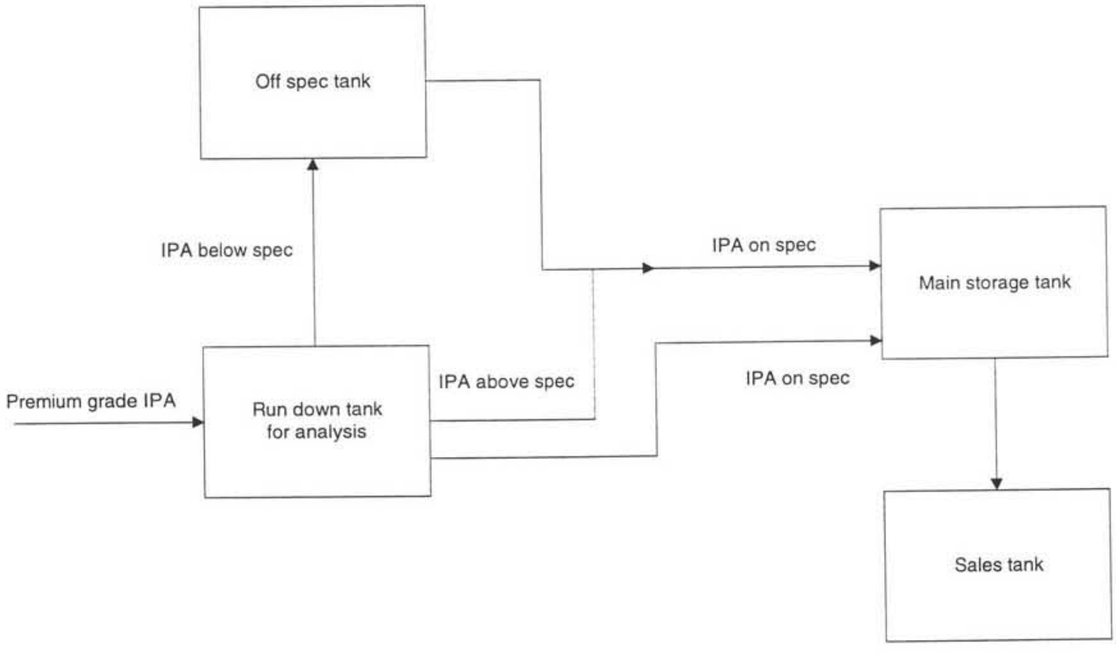


Figure 5, schematic overview of storage tanks

9 Wastes

Seven streams leave this process. Table 36 shows which streams are considered waste streams and which solutions have been found for them. Table 36 shows the future use of the other streams which leave this process and are not considered waste streams.

Table 36, Waste streams.

Waste stream	Ref.nr.	t/a	Remarks	Solution
To water treatment	<230>	30240	Contains cyclohexane	~~To water treatment~~
Water effluent	<002>	6720	Evaporated from dryer catalyst bed	Discharge in open air

Table 37, Other streams leaving the process.

Stream name	Ref.nr.	t/a	Classification	Future use
Light ends from C101	<116>	15960	Light ends	Fuel gas
Vent	<206>	1620	Light ends	Fuel gas
To NPA storage	<207>	~~168~~	By –product	Sale/Fuel gas
IPA product stream	<222>	174804	Product	Sale
Cokes	<001>	3612	Solids	Heating catalyst bed

10 Process Safety

10.1 HAZOP

A hazard and operability study (HAZOP)[18] has been carried out for the most critical units: both the reactor units and the C101 column, which operates under high pressure. The results can be found in Appendix 7, tables 7.1, 7.2 and 7.3.

The most important risks identified are:

- BLEVE potential at the top section of column C101, because liquid propylene is present here .
- Risk of autoignition of material inside R101, when oxygen leaking inside occurs
- Risk of runaway in R102, caused by propylene polymerisation, when an occurring temperature fluctuation is not adequately quenched with water.

10.2 Fire and Explosion Index

Also, for both reactor units and column C101, a Fire and Explosion Index (FIE20) has been calculated. The detailed calculation sheet can be found in Appendix 7.

In general, straightforward following the Dow guidelines led to the results. At some points, however, assumptions had to be made:

- the distance between the R101 reactor and the F101 furnace was estimated at about 30 m (105 ft)
- the distance between the C101 column and the F101 furnace was estimated at about 50 m (150 ft)
- the distance between the R102 reactor and the F101 furnace was estimated to be larger than 70 m (210 ft)
- although in mass there is not so much material present, a penalty of 0.10 is given at "G. Quantity of flammable material"
- for the "L. Rotating equipment" risk at R101, compressor K101 is taken into consideration; for R102, K102; for C101, K103.

The most important results are the following:

Table 38, FEI analysis basic results

Process Unit	Propylene Reactor	IPA Reactor	C101 Column
Unit Hazard Factor [-]	3.30	4.51	3.72
Material Factor [-]	21	21	21
Damage Factor [-]	0.61	0.70	0.63
F&E Index [-]	69.3	94.8	78.2
Exposure Radius [m]	17.7	24.3	20.0
Degree of Hazard	Moderate	Moderate	moderate

The F&E indices vary from 70 to 95; this classifies the degree of hazard for the process as moderate[21].

The numbers in the table must explained as follows: e.g. the propylene reactor has a damage factor of 0.61 and an exposure radius of 17.7 m. This means that, when a fire or explosion incident occurs, there is a 61% damage probability to objects within 17.7 m distance of the reactor (990 m^2 surrounding area).

11 Economy

With a price of 120 US$ for propane and 550 US$ for IPA there appears to be an interesting margin. In this chapter cost and income will be estimated in order to determine if the margin is as attractive as it may appear.

In the first section the total fixed capital will be estimated based on the equipment as has been designed. Using this fixed capital we are able to determine the operating costs, and thus the net cash flow. Using this cash flow, the economic criteria are calculated. At the end of the chapter a conclusion will be drawn whether the project is economically viable.

11.1 Calculation of Costs

The total investment cost is the sum of the fixed capital and working capital. Fixed capital is the total cost of the plant ready for start-up, which is a once-only cost. Working capital is the additional investment needed, over and above the fixed capital, to start the plant up and operate it to the point when income is earned.

Fixed Capital

The fixed capital can be estimated with the factorial method of Lang. The fixed capital cost of the project is given as a function of the total purchase equipment cost by the equation[18, p216]:

$$C_F = f_L * C_E \qquad [11.1]$$

C_F = Fixed capital

f_L = Lang factor, which depents on the type of the process

C_E = Total delivered cost of all major equipment

The contribution of different items to the total capital cost is calculated by multiplying the total Purchase Cost of Equipment (PCE) by an appropriate factor. Typical factors are found in the table below.

Table 39, Lang Factors

Item	f_o	Lang factor
Direct costs		
Equipment erection	f_1	0.4
Piping	f_2	0.7
Instrumentation	f_3	0.2
Electrical	f_4	0.1
Buildings, process	f_5	0.15
Storage	f_6	0.15
Ancillary buildings	f_7	0.15
$(f_1+\ldots+f_7)$ + 1		1.85 → 2.85
Indirect costs		
Design and engineering	f_8	0.3
Contractor's fee	f_9	0.05
Contingency	f_{10}	0.1
$(f_8+f_9+f_{10})$ + 1		0.45 → 1.45

The total Physical Plant Cost (PPC) can be calculated based on the direct costs, as can be seen in equation 11.2.

$$PPC = [1 + f_1 + ... + f_7] \cdot PCE = 2.85 \cdot PCE \qquad [11.2]$$

The Fixed Capital (FC) should include the direct costs and the indirect costs that are made, see equation 11.3.

$$FC = [1 + f_8 + f_9 + f_{10}] \cdot PPC = 1.45 \cdot PPC = 4.13 \cdot PCE \qquad [11.\ 2]$$

The purchase costs of equipment can be found, in appendix 9, to be 16.47 million US$, these calculations were based on the DACE[22] price guide and Coulson & Richardson's [~~18~~, chapter 6]. The catalyst is also placed in this table, this because of traditional reasons. There are no placed-spares taken into account due to company policy.
23 installed.

Now the total fixed costs can be calculated:

$$C_F = f_L \cdot C_E = 4.13 \cdot PCE = 68.07 \text{ M\$}. \qquad [11.\ 3]$$

Working capital

23

The working capital can be estimated with a typical[~~18~~, p218] figure for petrochemical plants, which is 15 per cent of the fixed capital. In table 40 the total investment is given, by adding the working capital to fixed capital

Table 40, summary of fixed, working capital and the total investment.

		Results [M$]
Fixed capital [million $]	C_F	68.07
Working capital	$0.15^* C_F$	10.21
Total investment	$1.15^* C_F$	78.28

Operating Costs

The operating costs (or production costs), the costs of producing the product after the plant has been built, are divided in direct and indirect costs. The direct costs can be divided in variable and fixed cost. In the block scheme a summary is given of the cost that are included[~~18~~].
23

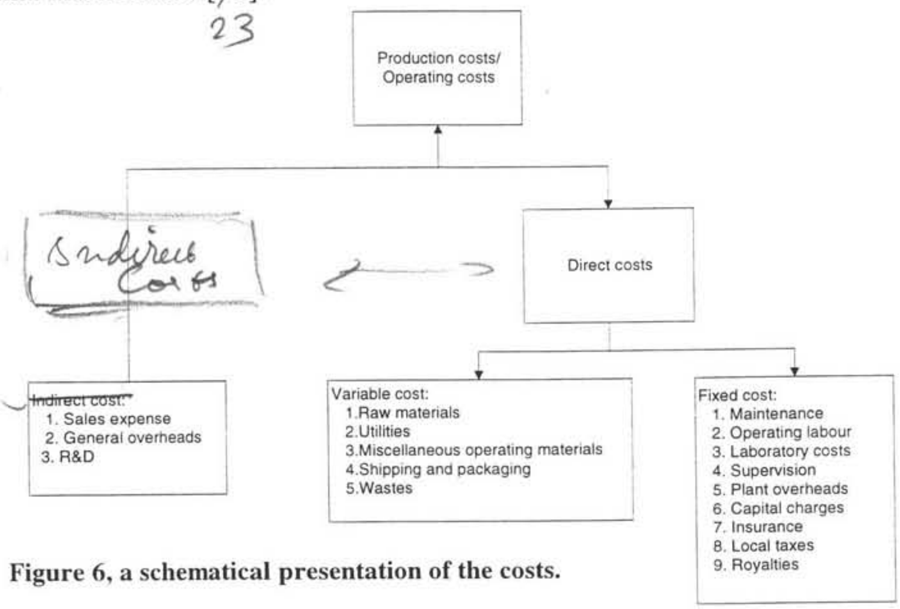


Figure 6, a schematical presentation of the costs.

10%

Variable costs

In appendix 9 the build up of these variable cost are given, which are summarised in table ~~40~~. As a rough guide the miscellaneous will be taken as ten per cent of the total maintenance costs. Storage tanks have been included in the fixed capital and therefore no other shipping and packing cost have been taken into account.

41

Fixed costs and Indirect costs

To estimate the fixed and the indirect costs, standard guidelines are used[~~18~~] see table 39.
23

In table 41 a summary is given of the variable, fixed and indirect production costs (this can also be found in appendix 9, table 18).

Table 41, summary of production costs.

	Costs [M$/a]	Remarks
Variable costs		
Raw materials	17.72	
Miscellaneous materials	0.68	10% of maintenance
Utilities	31.81	
Waste	0.05	
Subtotal A	50.27	
Fixed costs		
Maintenance	6.81	10% of fixed capital
Operating labour	2.00	
Laboratory costs	0.50	25% of operating labour
Supervision	0.40	20% of operating labour
Plant overheads	2.00	100% of operating labour
Capital charges	10.21	15% of fixed capital
Insurance	0.68	1% of fixed capital
Local taxes	1.36	2% of fixed capital
Royalties	0.68	1% of fixed capital
Subtotal B	24.64	
Direct prod costs (A+B)	**74.91**	
Sales expense	18.73	25 % of direct prod. costs
General overheads		
Research and development		
Indirect costs (C)	**18.73**	
Annual prod costs (A+B+C)	93.63	
Production cost [$/ton]	535.65	

%

30%

11.2 Economic criteria

The economic evaluation parameters are the rate of return (ROR), the pay-out-time (POT) and the discount cash-flow rate of return (DCFRR).

Rate of return

The ROR is the ratio of annual profit to investment and is based on the average income over the life of the project and the original investment. This can be calculated using equation 11.4:

$$ROR = \frac{\sum CF_i}{LP \cdot IC_0} \qquad [11.4]$$

CF_i = Cash flow generated in the year i.

LP = Life of the project.

IC_0 = Total investment valued at the first year.

The cash flow is the difference between the annual production costs and the annual income and can be found in appendix 9, table 20. It is assumed that the invested capital will be spent linear over the two construction years. As mentioned before the plant life has been set to 22 years including a construction period.

Pay-out-time
The POT is the time required after the start of the project to reach break even. The POT can be calculated using equation 11.5.

$$POT = \frac{1}{ROR} \qquad [11.5]$$

Discount cash-flow-rate of return
The DCFRR is the interest rate at which the cumulative net present worth at the end is zero. It is a measure of the maximum rate that the project can pay and still break even at the end of the project.

In appendix 9 calculations can be found that are needed to determine the economic evaluation parameters. These parameters are summarised in table 42:

Table 42, results of economic evaluation parameters

	Results
ROR	0.05
POT	19.46
DCFRR	1.14 %

11.3 Conclusion

The utilities are the most important cost drivers of the process[6] which implies that it may be interesting to investigate whether the use or costs of utilities can be reduced. The total fixed capital was based on the equipment. Heat exchanger and compressors are the main cost drivers concerning the equipment, this is due to the large energy changes within the process. A deeper investigation into these two cost drivers may result in a more interesting concept.

Although the economic criteria will probably benefit from an investigation into the main cost drivers, alternative use of the capital will decide whether the project is attractive. These figures are based on the conservative estimate of a plant life of 22 years. Changing this plant life to 30 years including construction will result in a increase of the ROR, DCFRR (resp. 0.2% and 1.5%) and a reduction of the POT of 3.8 months.

[6] A decrease of 50% of the electrical costs will bring the POT to 12 years and the DCFFR above 6%.

12 Conclusions and Recommendations

The objective of this design was to integrate a propane dehydrogenation unit and a IPA production unit. This was done in order to save unit operations and energy/utilities. Several commercial processes for dehydrogenation and propylene hydration have been investigated. Finally the option combining the Catofin process with the Veba process has been selected (see chapter Process Options & Selection). In this combination the usual separation units after the propylene reactor are not needed; the PP-splitter and several other unit operations are left out.

The new design has been simulated using Aspen Plus 10. Due to inaccuracies in the thermodynamic models the units for the azeotropic distillation were modeled as ideal separators. Performance of the distillation could be optimized.

Conversion of the propylene reactor was satisfying. Product of the dehydrogenation section contains 84.6 wt% propylene, enough to leave the PP splitter out (minimum 70 wt%). IPA is produced on specification, containing at most 0.05% NPA and 0.1% water.

Heat integration was also performed in this design. Utilities were saved mainly by steam generation and feed-product exchangers. Nonetheless, utility consumption is still quite high. Heat integration could be further analysed.

Finally, the economic analysis has been made. This design has a rate on return of 5% (discount cash-flow-rate of return of 1.14%) and a pay-out-time of 19.5 years.

Based on the outcome of this report, the following recommendations are made:

1) Utilities are the cost drivers for this process. Reduction in use of electricity could improve the economic profitability. Performance of the purification section could be optimised by using more accurate thermodynamic models in the design and further heat integration analysis.
2) Propylene and propane is lost in Unit 200. This is due to the fact that some unit operations are simulated at sub-optimal conditions of temperature and pressure. Appendix 12B shows a recommended flow scheme. Adjusted temperatures and pressures for some of the unit operations after the IPA reactor should be used in future design (sizing and cost estimation was already based on these adjusted figures).
3) The fixed bed used in the propylene reactor is a common cylindrical fixed bed. Because of the small pressure drop allowed, the catalyst bed has a fairly large cross-sectional area. Another option would be the use of a radial flow fixed bed. Attention should be paid to the packing of the bed. If it is not done correctly, the resistance of the bed may be lower at a certain spot and more gas will pass the bed at that spot. This speeds up catalyst deactivation at that spot and conversions drops. Calculations have to determine if the use of a radial flow fixed bed would be preferable.
4) Attention should be paid to the diameter of the catalyst used in the propylene reactor. It has been set at 5.3 mm for this design. However, it should be investigated if transport-limitation within the catalyst sphere does not appear.

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List of Symbols

Symbols are listed in the order of appearance in the report.

Chapter 4:

Reaction kinetics

K_p	Equilibrium constant	Pa
ξ	Conversion	--
P	Reaction pressure	Pa
ν_I	Stoiciometric coefficient	--
R	Gas constant	J/mol/K
T	Temperature	K

Chapter 8:

Reactor design

$m^3_{cat, required}$	Amount of catalyst needed	m^3
GHSV	Gas Hourly Space Velocity	$m^3_{gas}/m^3_{cat}/h$
Δp	Pressure drop	Bara
ρ	Density	kg/m^3
L	Height of catalyst bed	m
υ_0	Superficial gasvelocity	m/s
d_p	Particle diameter	m
ε	Void fraction	--
ν	kinematic viscosity	Ns/m^2

Columns sizing and design

F_{LV}	Liquid-vapour flow factor	-
ϕ_{mL}	Liquid mass flowrate	kg/s
ϕ_{mV}	Vapour mass flowrate	kg/s
ρ_V	Vapour density	kg/m^3
ρ_L	Liquid density	kg/m^3
u_f	Flooding vapour velocity	m/s
K_1	Flooding velocity constant	m/s
D_{col}	Column diameter	m
A_{col}	Column cross-sectional area	m^2
A_{net}	Net area for vapour-liquid disengagement	m^2
N_{actual}	Actual number of trays	--
$N_{theoretical}$	Theoretical number of trays	--
E_{col}	Column efficiency	--
L_{col}	Length of the column	m
t_s	Residence time	s
$u_{h,min}$	Minimum vapour velocity through holes	m/s
K_2	Constant depending on clear liquid depth	$kg^{1/2}/m^{5/2}/s$
D_h	Hole diameter	m
h_{ow}	Weir crest	mm liquid
l_w	Weir length	m
h_t	Total plate drop	mm liquid
h_w	Weir height	mm

h_d	Dry plate drop	m liquid
h_r	Residual head	mm liquid
u_h	Vapour velocity through holes	m/s
C_0	Orifice coefficient	-
h_b	Downcomer back-up	mm liquid
h_{dc}	Head loss in downcomer	mm liquid
A_{ap}	Clearance area under apron	m^2
A_d	Downcomer cross-sectional diameter	m

Separation Theory (absorber)

Vw*	Gas mass flowrate per unit column cross-sect. area	kg/m^2s
F_p	Packing factor	m^{-1}
μ_L	Liquid viscosity	Ns/m^2
ρ_V	Vapour density	kg/m^3
ρ_L	Liquid density	kg/m^3
K_4	Constant	--
$D_{absorber}$	Absorberdiameter	m
$\phi_{M,gas}$	Mass flow	kg/s
$L_{absorber}$	Length of the absorber	m
HETP	Height Equivalent of a Theoretical Plate	m

Heat Exchangers

Q	Heat transferred per unit time	W
U	Overall heat transfer coefficient	W/m^2 °C
A	Heat transfer area	m^2
ΔT_M	Mean temperature difference	°C
ΔT_{lm}	Log mean temperature difference	°C
F_t	Temperature correction factor	-
R	Dimensionless temperature ratio	-
T_1	Inlet shell-side fluid temperature	°C
T_2	Outlet shell-side fluid temperature	°C
t_1	Inlet tube-side temperature	°C
t_2	Outlet tube-side temperature	°C
S	Dimensionless temperature ratio	-
ϕ_M	Mass flow	kg/s
C_p	Heat capacity	J/kg °C
T_{out}	Outgoing temperature	°C
T_{in}	Ingoing temperature	°C
ΔH_{vap}	Heat of vaporization	kJ/kg

Air Coolers

t_1	Inlet air temperature	°C
t_2	Outlet air temperature	°C
U	Overall heat transfer coefficient	W/m^2 °C
T_2	Outlet temperature process stream	°C

T_1	Inlet temperature process stream	°C

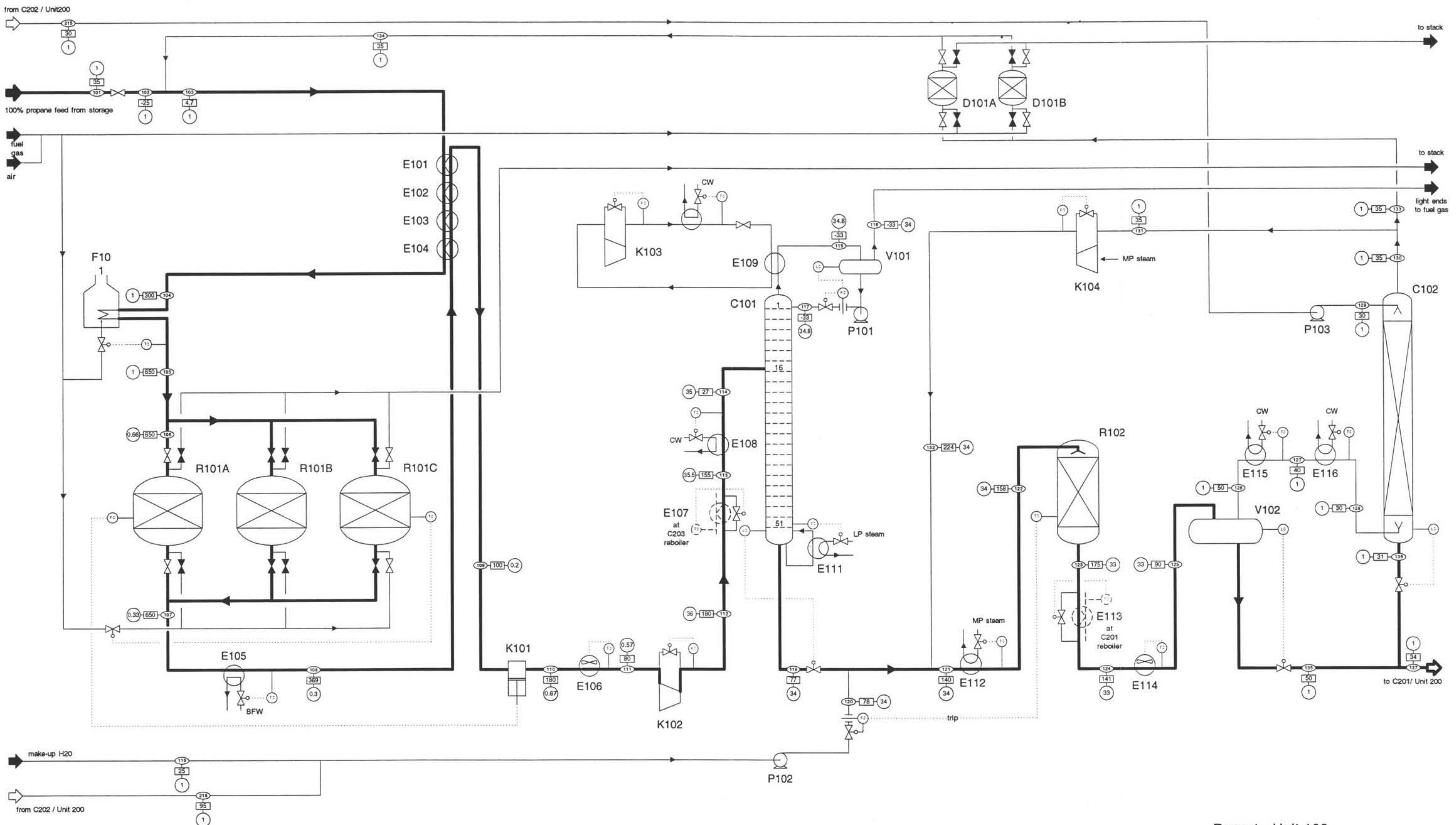
Furnace

A_{tubes}	Total tube area	m^2
ΔH_{flow}	Heat change of flow	J/s
Q_{rad}	Heat flux to tubes	kW/m^2
N_{tubes}	Number of tubes	-
$\phi_{v,gas}$	Flow through tubes	m^3/s
$A_{\perp}$	Cross section tubes	m^2
v_{gas}	Velocity in the tube	m/s
$\phi_{m,fuel}$	Flow of fuel oil	kg/s
ΔH_{comb}	Heat of combustion	MJ/kg

Pumps

ΔP	Pressure differential across the pump	bar
Q_p	Flow rate	m^3/h
η_p	Pump efficiency	%

APPENDICES: TABLE OF CONTENTS



Page 1 - Unit 100
Reaction section

Process Equipment Summary - Unit 100

C101	deethanizer	E109	C101 condenser	K103	C101 heat pump compressor
C102	washing column	E110	E101 condenser heat pump	K104	R102 recycle compressor
D101A/B	dryer	E111	C101 reboiler	P101	C101 reflux pump
E101	R101 feed-effluent A	E112	R102 feed preheater	P102	R102 water feed pump
E102	R101 feed-effluent B	E113	C201 reboiler	P103	C102 absorbent feed pump
E103	R101 feed-effluent C	E114	R102 effluent cooler	R101A/B/C	propylene reactor
E104	R101 feed-effluent D	E115	C102 feed cooler A	R102	IPA reactor
E105	R101 effluent-steam	E116	C102 feed cooler B	V101	C101 accumulator
E106	K101 effluent	F101	R101 feed furnace	V102	HP separator
E107	C203 reboiler	K101	R101 vacuum compressor		
E108	C101 feed cooler	K102	R102 HP compressor		

Designers

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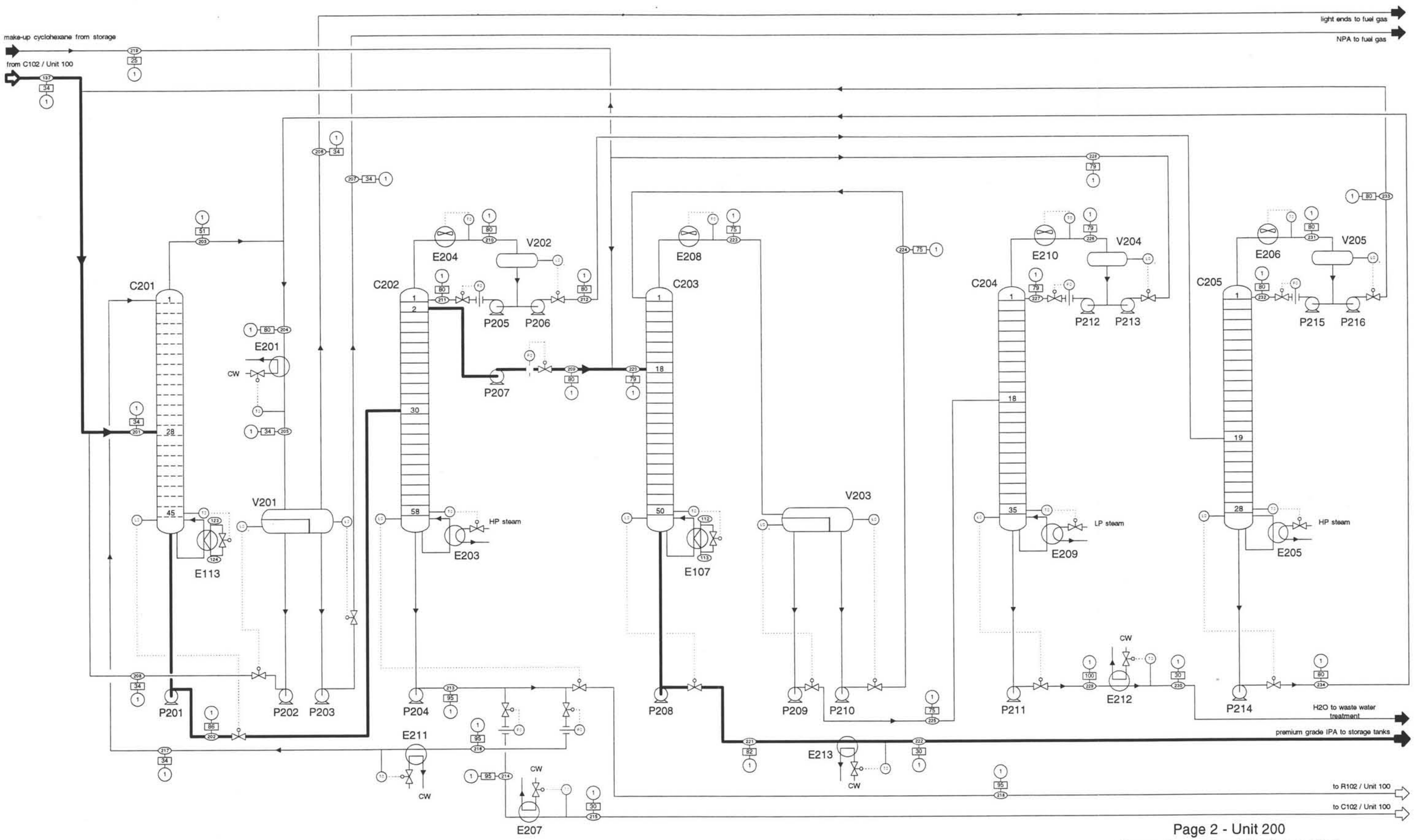
Process Flow Scheme

Project	: Isopropylalcohol from propane by successively dehydrogenation and hydration
Project ID. Number	: CPD3242
Completion Date	: December 22nd 1999

CONFIDENTIAL

Stream number — Temperature (C) — Pressure (Bara)



Page 2 - Unit 200
Separation and purification section

Process Equipment Summary - Unit 200

C201	light ends column	E207	recycle to C101 cooler	P205	C202 reflux pump	P216	C205 transfer pump
C202	IPA CBM column	E208	C203 condenser	P206	C202 top transfer pump	V201	light ends column phase separator
C203	IPA drying column	E209	C204 reboiler	P207	C202 product transfer pump	V202	C202 accumulator
C204	cyclohexane recovery column	E210	C204 condenser	P208	C203 bottom pump	V203	drying column phase separator
C205	NPA recovery column	E211	C201 top feed cooler	P209	V202 inorganic phase pump	V204	C204 accumulator
E201	V201 effluent cooler	E212	C204 bottom cooler	P210	V202 organic phase pump	V205	C205 accumulator
E202	[deleted]	E213	C203 product cooler	P211	C204 bottom pump		
E203	C202 reboiler	P201	C201 bottom pump	P212	C204 reflux pump		
E204	C202 condenser	P202	V201 inorganic phase pump	P213	C204 transfer pump		
E205	C205 reboiler	P203	V201 organic phase pump	P214	C205 bottom pump		
E206	C205 condenser	P204	C202 bottom pump	P215	C205 reflux pump		

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Process Flow Scheme

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CONFIDENTIAL

Stream number · Temperature (C) · Pressure (Bara)

Better.

PROCESS
STREAM SUMMARY

STREAM nr.:		<001>	OUT	<002>	OUT	<101>	✓ IN
Name:		COKES		WATER EFFLUENT		PROPANE FEED	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLENE	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	397.00	17468
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	0.00	0	0.00	0	0.00	0
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	0.00	0	44.46	800	0.00	0
CARBON	12.00	36.00	432	0.00	0	0.00	0
Total		36.00	432	44.46	800	397.00	17468
Enthalpy (MW)		6.46		-3.52		-13.16	
Phase		S		L		L	
Pressure (bara)		0.50		1.00		15.00	
Temperature C		650		35		35	

STREAM nr.:		<001>	Out	<002>	Out	<101>	
Name:		COKES		WATER EFFLUENT		PROPANE FEED	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	100.0	100.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	0.0	0.0	100.0	100.0	0.0	0.0
CARBON	12.00	100.0	100.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		6.46		-3.52		-13.16	
Phase		S		L		L	
Pressure (bara)		0.50		1.00		15.00	
Temperature C		650		35		35	

PROCESS STREAM SUMMARY

~~1~~ 105

STREAM nr.:		<102>		<103>		<104>	
Name:		FROM VALVE		TO E101		TO FURNACE	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	1.34	3	1.34	3
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.14	4	0.14	4
ETHANE	30.00	0.00	0	0.80	24	0.80	24
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	589.37	24754	589.37	24754
PROPANE	44.00	397.00	17468	517.43	22767	517.43	22767
ACETONE	58.00	0.00	0	0.38	22	0.38	22
IPA	60.00	0.00	0	18.14	1088	18.14	1088
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	7.76	791	7.76	791
WATER	18.00	0.00	0	0.00	0	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		397.00	17468	1135.35	49452	1135.35	49452
Enthalpy (MW)		-6.30		-8.55		-6.93	
Phase		V/L		V/L		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		-25		5		300	

STREAM nr.:		<102>		<103>		<104>	
Name:		FROM VALVE		TO E101		TO FURNACE	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.1	0.0	0.1	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.1	0.0	0.1	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	51.9	50.1	51.9	50.1
PROPANE	44.00	100.0	100.0	45.6	46.0	45.6	46.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	1.6	2.2	1.6	2.2
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.7	1.6	0.7	1.6
WATER	18.00	0.0	0.0	0.0	0.0	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-6.30		-8.55		-6.93	
Phase		V/L		V/L		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		-25		5		300	

104

PROCESS STREAM SUMMARY

STREAM nr.:		<105>		<106>		<107>	
Name:		FROM FURNACE		TO R101		R101 EFFLUENT	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	1.34	3	1.34	3	405.82	812
METHANE	16.00	0.00	0	0.00	0	18.26	292
ETHYLENE	28.00	0.14	4	0.14	4	7.72	216
ETHANE	30.00	0.80	24	0.80	24	9.39	282
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	589.37	24754	589.37	24754	955.82	40145
PROPANE	44.00	517.43	22767	517.43	22767	122.11	5373
ACETONE	58.00	0.38	22	0.38	22	0.38	22
IPA	60.00	18.14	1088	18.14	1088	18.14	1088
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	7.76	791	7.76	791	7.76	791
WATER	18.00	0.00	0	0.00	0	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		1135.35	49452	1135.35	49452	1545.40	49020
Enthalpy (MW)		7.99		11.15		22.75	
Phase		V		V		V	
Pressure (bara)		1.00		0.66		0.33	
Temperature C		650		650		650	

STREAM nr.:		<105>		<106>		<107>	
Name:		FROM FURNACE		TO R101		R101 EFFLUENT	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.1	0.0	0.1	0.0	26.3	1.7
METHANE	16.00	0.0	0.0	0.0	0.0	1.2	0.6
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.5	0.4
ETHANE	30.00	0.1	0.0	0.1	0.0	0.6	0.6
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	51.9	50.1	51.9	50.1	61.8	81.9
PROPANE	44.00	45.6	46.0	45.6	46.0	7.9	11.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	1.6	2.2	1.6	2.2	1.2	2.2
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.7	1.6	0.7	1.6	0.5	1.6
WATER	18.00	0.0	0.0	0.0	0.0	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		7.99		11.15		22.75	
Phase		V		V		V	
Pressure (bara)		1.00		0.66		0.33	
Temperature C		650		650		650	

**PROCESS
STREAM SUMMARY**

STREAM nr.:		<108>		<109>		<110>	
Name:		FROM E105		TO K101		FROM K101	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	405.82	812	405.82	812	405.82	812
METHANE	16.00	18.26	292	18.26	292	18.26	292
ETHYLENE	28.00	7.72	216	7.72	216	7.72	216
ETHANE	30.00	9.39	282	9.39	282	9.39	282
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	955.82	40145	955.82	40145	955.82	40145
PROPANE	44.00	122.11	5373	122.11	5373	122.11	5373
ACETONE	58.00	0.38	22	0.38	22	0.38	22
IPA	60.00	18.14	1088	18.14	1088	18.14	1088
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	7.76	791	7.76	791	7.76	791
WATER	18.00	0.00	0	0.00	0	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		1545.40	49020	1545.40	49020	1545.40	49020
Enthalpy (MW)		10.25		1.07		3.47	
Phase		V		V		V	
Pressure (bara)		0.30		0.20		0.67	
Temperature C		369		100		180	

STREAM nr.:		<108>		<109>		<110>	
Name:		FROM E105		TO K101		FROM K101	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	26.3	1.7	26.3	1.7	26.3	1.7
METHANE	16.00	1.2	0.6	1.2	0.6	1.2	0.6
ETHYLENE	28.00	0.5	0.4	0.5	0.4	0.5	0.4
ETHANE	30.00	0.6	0.6	0.6	0.6	0.6	0.6
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	61.8	81.9	61.8	81.9	61.8	81.9
PROPANE	44.00	7.9	11.0	7.9	11.0	7.9	11.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	1.2	2.2	1.2	2.2	1.2	2.2
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.5	1.6	0.5	1.6	0.5	1.6
WATER	18.00	0.0	0.0	0.0	0.0	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		10.25		1.07		3.47	
Phase		V		V		V	
Pressure (bara)		0.30		0.20		0.67	
Temperature C		369		100		180	

PROCESS
STREAM SUMMARY

STREAM nr.:		<111>		<112>		<113>	
Name:		FROM E106		FROM K102		FROM E107	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	405.82	812	405.82	812	405.82	812
METHANE	16.00	18.26	292	18.26	292	18.26	292
ETHYLENE	28.00	7.72	216	7.72	216	7.72	216
ETHANE	30.00	9.39	282	9.39	282	9.39	282
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	955.82	40145	955.82	40145	955.82	40145
PROPANE	44.00	122.11	5373	122.11	5373	122.11	5373
ACETONE	58.00	0.38	22	0.38	22	0.38	22
IPA	60.00	18.14	1088	18.14	1088	18.14	1088
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	7.76	791	7.76	791	7.76	791
WATER	18.00	0.00	0	0.00	0	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		1545.40	49020	1545.40	49020	1545.40	49020
Enthalpy (MW)		0.51		3.09		2.26	
Phase		V		V		V	
Pressure (bara)		0.57		36.00		35.50	
Temperature C		80		180		155	

STREAM nr.:		<111>		<112>		<113>	
Name:		FROM E106		FROM K102		FROM E107	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	26.3	1.7	26.3	1.7	26.3	1.7
METHANE	16.00	1.2	0.6	1.2	0.6	1.2	0.6
ETHYLENE	28.00	0.5	0.4	0.5	0.4	0.5	0.4
ETHANE	30.00	0.6	0.6	0.6	0.6	0.6	0.6
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	61.8	81.9	61.8	81.9	61.8	81.9
PROPANE	44.00	7.9	11.0	7.9	11.0	7.9	11.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	1.2	2.2	1.2	2.2	1.2	2.2
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.5	1.6	0.5	1.6	0.5	1.6
WATER	18.00	0.0	0.0	0.0	0.0	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		0.51		3.09		2.26	
Phase		V		V		V	
Pressure (bara)		0.57		36.00		35.50	
Temperature C		80		180		155	

PROCESS
STREAM SUMMARY

STREAM nr.:		<114>		<115>		<116> V	OUT
Name:		TO C101		FROM TOP C101		FROM ACC. C101	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	405.82	812	405.82	812	405.82	812
METHANE	16.00	18.26	292	44.99	720	18.26	292
ETHYLENE	28.00	7.72	216	38.13	1068	7.58	212
ETHANE	30.00	9.39	282	63.05	1892	8.62	259
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	955.82	40145	788.46	33115	7.16	301
PROPANE	44.00	122.11	5373	72.13	3174	0.56	25
ACETONE	58.00	0.38	22	0.00	0	0.00	0
IPA	60.00	18.14	1088	0.00	0	0.00	0
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	7.76	791	0.00	0	0.00	0
WATER	18.00	0.00	0	0.00	0	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		1545.40	49020	1412.59	40780	448.00	1900
Enthalpy (MW)		-4.57		-16.38		-0.76	
Phase		V/L		V/L		V	
Pressure (bara)		35.00		34.80		34.00	
Temperature C		27		1		-33	

STREAM nr.:		<114>		<115>		<116>	
Name:		TO C101		FROM TOP C101		FROM ACC. C101	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	26.3	1.7	28.7	2.0	90.6	42.7
METHANE	16.00	1.2	0.6	3.2	1.8	4.1	15.4
ETHYLENE	28.00	0.5	0.4	2.7	2.6	1.7	11.2
ETHANE	30.00	0.6	0.6	4.5	4.6	1.9	13.6
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	61.8	81.9	55.8	81.2	1.6	15.8
PROPANE	44.00	7.9	11.0	5.1	7.8	0.1	1.3
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	1.2	2.2	0.0	0.0	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.5	1.6	0.0	0.0	0.0	0.0
WATER	18.00	0.0	0.0	0.0	0.0	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-4.57		-16.38		-0.76	
Phase		V/L		V/L		V	
Pressure (bara)		35.00		34.80		34.00	
Temperature C		27		1		-33	

PROCESS
STREAM SUMMARY

STREAM nr.:		<117>		<118>		<119>	IN
Name:		REFLUX C101		BOTTOM C101		WATER MAKEUP	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	26.73	428	0.00	0	0.00	0
ETHYLENE	28.00	30.55	855	0.13	4	0.00	0
ETHANE	30.00	54.43	1633	0.77	23	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	781.30	32815	948.67	39844	0.00	0
PROPANE	44.00	71.57	3149	121.55	5348	0.00	0
ACETONE	58.00	0.00	0	0.38	22	0.00	0
IPA	60.00	0.00	0	18.14	1088	0.00	0
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	7.76	791	0.00	0
WATER	18.00	0.00	0	0.00	0	594.39	10699
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		964.59	38880	1097.40	47120	594.39	10699
Enthalpy (MW)		-15.62		-3.08		-35.54	
Phase		L		L		L	
Pressure (bara)		34.80		34.00		1.00	
Temperature C		-33		77		25	

STREAM nr.:		<117>		<118>		<119>	
Name:		REFLUX C101		BOTTOM C101		WATER MAKEUP	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	2.8	1.1	0.0	0.0	0.0	0.0
ETHYLENE	28.00	3.2	2.2	0.0	0.0	0.0	0.0
ETHANE	30.00	5.6	4.2	0.1	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	81.0	84.4	86.4	84.6	0.0	0.0
PROPANE	44.00	7.4	8.1	11.1	11.4	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	1.7	2.3	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.7	1.7	0.0	0.0
WATER	18.00	0.0	0.0	0.0	0.0	100.0	100.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-15.62		-3.08		-35.54	
Phase		L		L		L	
Pressure (bara)		34.80		34.00		1.00	
Temperature C		-33		77		25	

PROCESS STREAM SUMMARY

PROCESS STREAM

STREAM nr.:		<120>		<121>		<122>	
Name:		FROM P102		TO E112		TO R102	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	12.10	24	12.10	24
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	1.39	39	1.39	39
ETHANE	30.00	0.00	0	7.96	239	7.96	239
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	6251.04	262544	6251.04	262544
PROPANE	44.00	0.00	0	1205.63	53048	1205.63	53048
ACETONE	58.00	0.00	0	3.77	218	3.77	218
IPA	60.00	0.00	0	181.12	10867	181.12	10867
NPA	60.00	0.00	0	0.01	0	0.01	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	77.02	7856	77.02	7856
WATER	18.00	2330.00	41940	2730.06	49141	2730.06	49141
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		2330.00	41940	10470.09	383976	10470.09	383976
Enthalpy (MW)		-174.52		-214.70		-200.70	
Phase		L		V/L		V	
Pressure (bara)		34.00		34.00		34.00	
Temperature C		78		140		158	

STREAM nr.:		<120>		<121>		<122>	
Name:		FROM P102		TO E112		TO R102	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.1	0.0	0.1	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.1	0.1	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	59.7	68.4	59.7	68.4
PROPANE	44.00	0.0	0.0	11.5	13.8	11.5	13.8
ACETONE	58.00	0.0	0.0	0.0	0.1	0.0	0.1
IPA	60.00	0.0	0.0	1.7	2.8	1.7	2.8
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.7	2.0	0.7	2.0
WATER	18.00	100.0	100.0	26.1	12.8	26.1	12.8
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-174.52		-214.70		-200.70	
Phase		L		V/L		V	
Pressure (bara)		34.00		34.00		34.00	
Temperature C		78		140		158	

PROCESS STREAM SUMMARY

Eff. R102 !

STREAM nr.:		<123>		<124>		<125>	
Name:		TO E113		FROM E113		TO V102	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	13.40	27	13.40	27	13.40	27
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	1.39	39	1.39	39	1.39	39
ETHANE	30.00	7.96	239	7.96	239	7.96	239
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	5899.40	247775	5899.40	247775	5899.40	247775
PROPANE	44.00	1205.63	53048	1205.63	53048	1205.63	53048
ACETONE	58.00	5.07	294	5.07	294	5.07	294
IPA	60.00	528.30	31698	528.30	31698	528.30	31698
NPA	60.00	0.51	30	0.51	30	0.51	30
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	78.34	7991	78.34	7991	78.34	7991
WATER	18.00	2379.75	42835	2379.75	42835	2379.75	42835
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		10119.75	383976	10119.75	383976	10119.75	383976
Enthalpy (MW)		-185.24		-205.42		-231.25	
Phase		V		V/L		V/L	
Pressure (bara)		32.86		32.86		32.86	
Temperature C		175		141		90	

STREAM nr.:		<123>		<124>		<125>	
Name:		TO E113		FROM E113		TO V102	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.1	0.0	0.1	0.0	0.1	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.1	0.1	0.1	0.1	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	58.3	64.5	58.3	64.5	58.3	64.5
PROPANE	44.00	11.9	13.8	11.9	13.8	11.9	13.8
ACETONE	58.00	0.1	0.1	0.1	0.1	0.1	0.1
IPA	60.00	5.2	8.3	5.2	8.3	5.2	8.3
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.8	2.1	0.8	2.1	0.8	2.1
WATER	18.00	23.5	11.2	23.5	11.2	23.5	11.2
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-185.24		-205.42		-231.25	
Phase		V		V/L		V/L	
Pressure (bara)		32.86		32.86		32.86	
Temperature C		175		141		90	

PROCESS
STREAM SUMMARY

STREAM nr.:		<126>		<127>		<128>	
Name:		FROM V102		TO E116		TO C102	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	13.40	27	13.40	27	13.40	27
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	1.39	39	1.39	39	1.39	39
ETHANE	30.00	7.96	239	7.96	239	7.96	239
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	5898.90	247754	5898.90	247754	5898.90	247754
PROPANE	44.00	1205.53	53043	1205.53	53043	1205.53	53043
ACETONE	58.00	4.96	288	4.96	288	4.96	288
IPA	60.00	503.55	30213	503.55	30213	503.55	30213
NPA	60.00	0.46	28	0.46	28	0.46	28
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	78.32	7989	78.32	7989	78.32	7989
WATER	18.00	1077.73	19399	1077.73	19399	1077.73	19399
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		8792.19	359017	8792.19	359017	8792.19	359017
Enthalpy (MW)		-116.90		-127.48		-130.45	
Phase		V		V/L		V/L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		50		40		30	

STREAM nr.:		<126>		<127>		<128>	
Name:		FROM V102		TO E116		TO C102	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.2	0.0	0.2	0.0	0.2	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.1	0.1	0.1	0.1	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	67.1	69.0	67.1	69.0	67.1	69.0
PROPANE	44.00	13.7	14.8	13.7	14.8	13.7	14.8
ACETONE	58.00	0.1	0.1	0.1	0.1	0.1	0.1
IPA	60.00	5.7	8.4	5.7	8.4	5.7	8.4
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.9	2.2	0.9	2.2	0.9	2.2
WATER	18.00	12.3	5.4	12.3	5.4	12.3	5.4
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-116.90		-127.48		-130.45	
Phase		V		V/L		V/L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		50		40		30	

PROCESS
STREAM SUMMARY

STREAM nr.:		<129>		<130>		<131>	
Name:		TOP FEED C102		TOP C102		TO K104	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	13.40	27	12.06	24
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	1.39	39	1.25	35
ETHANE	30.00	0.00	0	7.96	239	7.16	215
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	5893.72	247536	5304.35	222783
PROPANE	44.00	0.00	0	1204.33	52990	1083.89	47691
ACETONE	58.00	0.00	0	3.75	218	3.38	196
IPA	60.00	0.00	0	181.38	10883	163.24	9795
NPA	60.00	0.00	0	0.01	0	0.01	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	77.56	7911	69.81	7120
WATER	18.00	5233.15	94197	444.64	8003	400.17	7203
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		5233.15	94197	7828.13	327847	7045.31	295062
Enthalpy (MW)		-399.05		-52.35		-47.11	
Phase		L		V		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		30		35		35	

STREAM nr.:		<129>		<130>		<131>	
Name:		TOP FEED C102		TOP C102		TO K104	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.2	0.0	0.2	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.1	0.1	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	75.3	75.5	75.3	75.5
PROPANE	44.00	0.0	0.0	15.4	16.2	15.4	16.2
ACETONE	58.00	0.0	0.0	0.0	0.1	0.0	0.1
IPA	60.00	0.0	0.0	2.3	3.3	2.3	3.3
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	1.0	2.4	1.0	2.4
WATER	18.00	100.0	100.0	5.7	2.4	5.7	2.4
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-399.05		-52.35		-47.11	
Phase		L		V		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		30		35		35	

PROCESS
STREAM SUMMARY

STREAM nr.:		<132>		<133>		<134>	
Name:		IPA RECYCLE		TO DRYER		OVERALL RECYCL	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	12.06	24	1.34	3	1.34	3
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	1.25	35	0.14	4	0.14	4
ETHANE	30.00	7.16	215	0.80	24	0.80	24
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	5304.35	222783	589.37	24754	589.37	24754
PROPANE	44.00	1083.89	47691	120.43	5299	120.43	5299
ACETONE	58.00	3.38	196	0.38	22	0.38	22
IPA	60.00	163.24	9795	18.14	1088	18.14	1088
NPA	60.00	0.01	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	69.81	7120	7.76	791	7.76	791
WATER	18.00	400.17	7203	44.46	800	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		7045.31	295062	782.81	32785	738.35	31984
Enthalpy (MW)		-19.30		-5.23		-2.25	
Phase		V		V		V	
Pressure (bara)		34.00		1.00		1.00	
Temperature C		224		35		35	

STREAM nr.:		<132>		<133>		<134>	
Name:		IPA RECYCLE		TO DRYER		OVERALL RECYCL	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.2	0.0	0.2	0.0	0.2	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.1	0.1	0.1	0.1	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	75.3	75.5	75.3	75.5	79.8	77.4
PROPANE	44.00	15.4	16.2	15.4	16.2	16.3	16.6
ACETONE	58.00	0.0	0.1	0.0	0.1	0.1	0.1
IPA	60.00	2.3	3.3	2.3	3.3	2.5	3.4
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	1.0	2.4	1.0	2.4	1.1	2.5
WATER	18.00	5.7	2.4	5.7	2.4	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-19.30		-5.23		-2.25	
Phase		V		V		V	
Pressure (bara)		34.00		1.00		1.00	
Temperature C		224		35		35	

PROCESS
STREAM SUMMARY

STREAM nr.:		<135>		<136>		<137>	
Name:		BOTTOM HP SEPA		BOTTOM ABSORB		TO UNIT 200	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.01	0	0.01	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.50	21	5.18	217	5.68	239
PROPANE	44.00	0.10	4	1.20	53	1.30	57
ACETONE	58.00	0.11	6	1.21	70	1.32	76
IPA	60.00	24.71	1483	322.17	19330	346.88	20813
NPA	60.00	0.04	3	0.46	27	0.50	30
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.02	2	0.76	77	0.78	80
WATER	18.00	1301.62	23429	5866.24	105592	7167.86	129021
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		1327.11	24949	6197.21	125368	7524.33	150316
Enthalpy (MW)		-104.77		-477.15		-583.28	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		50		31		34	

STREAM nr.:		<135>		<136>		<137>	
Name:		BOTTOM HP SEPA		BOTTOM ABSORB		TO UNIT 200	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.1	0.1	0.2	0.1	0.2
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.1	0.0	0.1
IPA	60.00	1.9	5.9	5.2	15.4	4.6	13.8
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.1	0.0	0.1
WATER	18.00	98.1	93.9	94.7	84.2	95.3	85.8
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-104.77		-477.15		-583.28	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		50		31		34	

PROCESS STREAM SUMMARY

STREAM nr.:		<201>		<202>		<203>	
Name:		FEED TO C201		BOTTOM C201		TOP C201	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.01	0	0.00	0	0.01	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	5.68	239	0.00	0	5.68	239
PROPANE	44.00	1.30	57	0.00	0	1.30	57
ACETONE	58.00	1.32	76	0.00	0	1.31	76
IPA	60.00	939.33	56360	939.33	56360	0.00	0
NPA	60.00	0.50	30	0.50	30	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.78	80	0.00	0	0.78	80
WATER	18.00	7487.83	134781	17753.83	319569	1.24	22
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		8436.75	191623	18693.66	375959	10.33	474
Enthalpy (MW)		-630.31		-1423.90		-0.21	
Phase		L		L		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		34		86		51	

STREAM nr.:		<201>		<202>		<203>	
Name:		FEED TO C201		BOTTOM C201		TOP C201	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.1	0.1	0.0	0.0	55.0	50.3
PROPANE	44.00	0.0	0.0	0.0	0.0	12.6	12.1
ACETONE	58.00	0.0	0.0	0.0	0.0	12.7	16.1
IPA	60.00	11.1	29.4	5.0	15.0	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	7.6	16.8
WATER	18.00	88.8	70.3	95.0	85.0	12.0	4.7
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-630.31		-1423.90		-0.21	
Phase		L		L		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		34		86		51	

PROCESS
STREAM SUMMARY

STREAM nr.:		<204>		<205>		~~<206>~~	OUT
Name:		TO E201		TO V201		VENT	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.01	0	0.01	0	0.01	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	5.68	239	5.68	239	5.68	239
PROPANE	44.00	1.30	57	1.30	57	1.30	57
ACETONE	58.00	1.31	76	1.31	76	1.31	76
IPA	60.00	306.49	18389	306.49	18389	0.00	0
NPA	60.00	0.32	19	0.32	19	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.78	80	0.78	80	0.78	80
WATER	18.00	175.56	3160	175.56	3160	0.00	0
CARBON	12.00	0.00	0	0.00	0		0
Total		491.45	22020	491.45	22020	9.09	452
Enthalpy (MW)		-99.71		-102.36		-0.14	
Phase		V/L		V/L		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		34		34	

STREAM nr.:		<204>		<205>		<206>	
Name:		TO E201		TO V201		VENT	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.1	0.1
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	1.2	1.1	1.2	1.1	62.5	52.8
PROPANE	44.00	0.3	0.3	0.3	0.3	14.3	12.7
ACETONE	58.00	0.3	0.3	0.3	0.3	14.5	16.9
IPA	60.00	62.4	83.5	62.4	83.5	0.0	0.0
NPA	60.00	0.1	0.1	0.1	0.1	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.2	0.4	0.2	0.4	8.6	17.6
WATER	18.00	35.7	14.4	35.7	14.4	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-99.71		-102.36		-0.14	
Phase		V/L		V/L		V	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		34		34	

PROCESS
STREAM SUMMARY

STREAM nr.:		<207>	OUT	<208>		<209>	
Name:		TO NPA STORAGE		TOP FEED C201		SIDE STREAM C20	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	0.00	0	306.49	18389	346.88	20813
NPA	60.00	0.32	19	0.00	0	0.19	11
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	0.00	0	175.56	3160	199.10	3584
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		0.32	19	482.05	21549	546.17	24408
Enthalpy (MW)		-0.02		-112.14		-41.34	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		34		34		80	

STREAM nr.:		<207>		<208>		<209>	
Name:		TO NPA STORAGE		TOP FEED C201		SIDE STREAM C20	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	63.6	85.3	63.5	85.3
NPA	60.00	100.0	100.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	0.0	0.0	36.4	14.7	36.5	14.7
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-0.02		-112.14		-41.34	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		34		34		80	

PROCESS
STREAM SUMMARY

STREAM nr.:		<210>		<211>		<212>	
Name:		OVERHEAD C202		REFLUX C202		DITILLATE C202	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	1934.67	116080	1342.22	80533	592.45	35547
NPA	60.00	0.94	56	0.62	37	0.32	19
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	1042.57	18766	723.84	13029	318.73	5737
CARBON	12.00	0.00	0	0.00	0		0
Total		2978.19	134903	2066.69	93600	911.50	41303
Enthalpy (MW)		-246.68		-171.16		-75.53	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		80		80	

STREAM nr.:		<210>		<211>		<212>	
Name:		OVERHEAD C202		REFLUX C202		DITILLATE C202	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	65.0	86.0	64.9	86.0	65.0	86.1
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	35.0	13.9	35.0	13.9	35.0	13.9
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-246.68		-171.16		-75.53	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		80		80	

PROCESS
STREAM SUMMARY

STREAM nr.:		<213>		<214>		<215>	
Name:		BOTTOM C202		TO E207		TO PUMP P103	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	0.00	0	0.00	0	0.00	0
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	17236.00	310248	5233.15	94197	5233.15	94197
CARBON	12.00		0		0	0.00	0
Total		17236.00	310248	5233.15	94197	5233.15	94197
Enthalpy (MW)		-531.47		-392.49		-399.05	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		95		95		30	

STREAM nr.:		<213>		<214>		<215>	
Name:		BOTTOM C202		TO E207		TO PUMP P103	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	100.0	100.0	100.0	100.0	100.0	100.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-531.47		-392.49		-399.05	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		95		95		30	

PROCESS
STREAM SUMMARY

STREAM nr.:		<216>		<217>		<218>	
Name:		TO E211		TO TOP C201		WATER RECYCLE	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	0.00	0	0.00	0	0.00	0
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	10266.00	184788	10266.00	184788	1735.61	31241
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		10266.00	184788	10266.00	184788	1735.61	31241
Enthalpy (MW)		-799.95		-812.85		-138.98	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		95		34		95	

STREAM nr.:		<216>		<217>		<218>	
Name:		TO E211		TO TOP C201		WATER RECYCLE	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	100.0	100.0	100.0	100.0	100.0	100.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-799.95		-812.85		-138.98	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		95		34		95	

PROCESS
STREAM SUMMARY

<209> 24408

STREAM nr.:		<219>	IN	<220>		<221>	
Name:		MAKE UP		FEED C203		HOT IPA	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	0.00	0	358.01	21481	346.88	20813
NPA	60.00	0.00	0	0.19	11	0.19	11
CYCLOHEXANE	84.00	0.25	21	2.65	223	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	0.00	0	199.10	3584	1.16	21
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		0.25	21	559.96	25299	348.23	20845
Enthalpy (MW)		-0.01		-46.23		-30.22	
Phase		L		V/L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		25		79		82	

STREAM nr.:		<219>		<220>		<221>	
Name:		MAKE UP		FEED C203		HOT IPA	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	0.0	0.0	63.9	84.9	99.6	99.8
NPA	60.00	0.0	0.0	0.0	0.0	0.1	0.1
CYCLOHEXANE	84.00	100.0	100.0	0.5	0.9	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	0.0	0.0	35.6	14.2	0.3	0.1
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-0.01		-46.23		-30.22	
Phase		L		V/L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		25		79		82	

PROCESS
STREAM SUMMARY

STREAM nr.:		<222>	OUT	<223>		<224>	
Name:		IPA PRODUCT		TOP C203		TOP FEED C203	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	346.88	20813	38.41	2305	27.28	1637
NPA	60.00	0.19	11	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	1152.65	96823	1150.00	96600
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	1.16	21	197.94	3563	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		348.23	20845	1389.01	102690	1177.28	98237
Enthalpy (MW)		-34.98		-62.33		-50.67	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		30		75		75	

STREAM nr.:		<222>		<223>		<224>	
Name:		IPA PRODUCT		TOP C203		TOP FEED C203	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	99.6	99.8	2.8	2.2	2.3	1.7
NPA	60.00	0.1	0.1	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	83.0	94.3	97.7	98.3
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	0.3	0.1	14.3	3.5	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-34.98		-62.33		-50.67	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		30		75		75	

PROCESS
STREAM SUMMARY

STREAM nr.:		<225>		<226>		<227>	
Name:		FEED C204		OVERHEAD C204		REFLUX C204	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	11.13	668	61.93	3716	50.87	3052
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	2.65	223	13.36	1123	10.98	922
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	197.94	3563	0.00	0	0.00	0
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		211.73	4454	75.29	4838	61.85	3974
Enthalpy (MW)		-16.65		-5.93		-4.86	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		75		79		79	

STREAM nr.:		<225>		<226>		<227>	
Name:		FEED C204		OVERHEAD C204		REFLUX C204	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	5.3	15.0	82.3	76.8	82.3	76.8
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	1.3	5.0	17.7	23.2	17.7	23.2
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	93.5	80.0	0.0	0.0	0.0	0.0
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-16.65		-5.93		-4.86	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		75		79		79	

PROCESS STREAM SUMMARY

STREAM nr.:		<228>		<229>		<230> ✓	OUT
Name:		C204 DISTILLATE		TO E212		TO WATER TREAT	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	11.13	668	0.00	0	0.00	0
NPA	60.00	0.00	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	2.40	202	0.25	21	0.25	21
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	0.00	0	197.94	3563	197.94	3563
CARBON	12.00	0.00	0	0.00	0	0.00	0
Total		13.54	870	198.19	3584	198.19	3584
Enthalpy (MW)		-1.06		-15.60		-15.80	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		79		100		30	

STREAM nr.:		<228>		<229>		<230>	
Name:		C204 DISTILLATE		TO E212		TO WATER TREAT	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	82.3	76.8	0.0	0.0	0.0	0.0
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	17.7	23.2	0.1	0.6	0.1	0.6
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	0.0	0.0	99.9	99.4	99.9	99.4
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-1.06		-15.60		-15.80	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		79		100		30	

PROCESS
STREAM SUMMARY

STREAM nr.:		<231>		<232>		<233>	
Name:		OVERHEAD C205		REFLUX C205		DISTILLATE C205	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	857.88	51473	571.92	34315	285.96	17158
NPA	60.00	0.01	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	433.24	7798	288.83	5199	144.41	2599
CARBON	12.00	0.00	0	0.00	0		0
Total		1291.13	59272	860.75	39514.33	430.38	19757
Enthalpy (MW)		-95.48		-63.65		-31.83	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		80		80	

STREAM nr.:		<231>		<232>		<233>	
Name:		OVERHEAD C205		REFLUX C205		DISTILLATE C205	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	66.4	86.8	66.4	86.8	66.4	86.8
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	33.6	13.2	33.6	13.2	33.6	13.2
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-95.48		-63.65		-31.83	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		80		80	

PROCESS
STREAM SUMMARY

<234> ?

STREAM nr.:		<231>		<232>		<233>	
Name:		OVERHEAD C205		REFLUX C205		DISTILLATE C205	
COMP	Mw	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr
HYDROGEN	2.00	0.00	0	0.00	0	0.00	0
METHANE	16.00	0.00	0	0.00	0	0.00	0
ETHYLENE	28.00	0.00	0	0.00	0	0.00	0
ETHANE	30.00	0.00	0	0.00	0	0.00	0
ETHANOL	46.00	0.00	0	0.00	0	0.00	0
PROPYLEN	42.00	0.00	0	0.00	0	0.00	0
PROPANE	44.00	0.00	0	0.00	0	0.00	0
ACETONE	58.00	0.00	0	0.00	0	0.00	0
IPA	60.00	857.88	51473	571.92	34315	285.96	17158
NPA	60.00	0.01	0	0.00	0	0.00	0
CYCLOHEXANE	84.00	0.00	0	0.00	0	0.00	0
N-HEXANE	86.00	0.00	0	0.00	0	0.00	0
DIPE	102.00	0.00	0	0.00	0	0.00	0
WATER	18.00	433.24	7798	288.83	5199	144.41	2599
CARBON	12.00	0.00	0	0.00	0		0
Total		1291.13	59272	860.75	39514.33	430.38	19757
Enthalpy (MW)		-95.48		-63.65		-31.83	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		80		80	

STREAM nr.:		<231>		<232>		<233>	
Name:		OVERHEAD C205		REFLUX C205		DISTILLATE C205	
COMP	Mw	mol%	wt%	mol%	wt%	mol%	wt%
HYDROGEN	2.00	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	16.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	28.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	30.00	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	42.00	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	44.00	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	58.00	0.0	0.0	0.0	0.0	0.0	0.0
IPA	60.00	66.4	86.8	66.4	86.8	66.4	86.8
NPA	60.00	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	84.00	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	86.00	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	102.00	0.0	0.0	0.0	0.0	0.0	0.0
WATER	18.00	33.6	13.2	33.6	13.2	33.6	13.2
CARBON	12.00	0.0	0.0	0.0	0.0	0.0	0.0
Total		100.0	100.0	100.0	100.0	100.0	100.0
Enthalpy (MW)		-95.48		-63.65		-31.83	
Phase		L		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature C		80		80		80	

Stream <234> ?

SUMMARY OF UTILITIES

EQUIPMENT		UTILITIES													REMARKS
		Heating					Cooling				Power				
Nr.	Name	Load	Consumption (t/h)				Load	Consumption (t/h)			Actual	Consumption (t/h, kWh/h)			
			Steam			Fuel		Cooling	Air	BFW	Load	Steam (t/h)		Electr.	
		kW	LP	MP	HP	Gas	kW	Water			kW	HP	MP	kWh/h	
E105	R101 effluent cooler						12498.00			17.17					Generates MP steam
E106	K101 effluent cooler						2986.00		351.36		7.99			7.99	Air cooler
E108	C101 feed cooler						6830.00	392.04							
E109	C101 condenser						5394.00	86.04							Medium: propylene
E110	C101 cond. heat pump						11599.00	665.64							
E111	C101 reboiler	6240.00	9.24												Self generated
E112	R102 feed heater	14061.00		24.48											
E114	R102 effluent cooler						25824.00		3038.04		98.55			98.55	Air cooler
E115	C102 feed cooler A						36459.00	2093.35							
E116	C102 feed cooler B						2967.50	511.15							
E201	V201 feed cooler						2615.00	150.12							
E203	C202 reboiler	58308.00			99.72										
E204 c	C202 condenser						50799.00		5976.36		498.15			498.15	Air cooler
E205	C205 reboiler	46950.00			80.28										
E206	C205 condenser						46950.00		5523.48		383.67			383.67	Air cooler
E207	Recycle to C101 cooler						6809.00	390.96							
E208	C203 top cooler						690.00		81.36		8.21			8.21	Air cooler
E209	C204 reboiler	5100.00	7.56												Self generated
E210	C204 condenser						5000.00		588.24		51.40			51.40	Air cooler
E211	C201 Top feed cooler						12899.30	2221.89							
E212	C204 bottom cooler						197.00	33.84							
E213	C203 product cooler						4765.00	820.08							
F101	Furnace catofin inlet	14918.00				1.69									
	TRANSPORT	145577.00	16.80	24.48	180.00	1.69	235281.80	7365.11	15558.84	17.17	1047.97			1047.97	

Designers	:	M.E. Brons	M.P. Hoff	Project ID Number :	**CPD3242**
		J.H.M. Jansen	E.P.E. Seveke	Completion Date :	December 16th 1999

SUMMARY OF UTILITIES

EQUIPMENT		UTILITIES													REMARKS
		Heating					Cooling				Power				
Nr.	Name	Load	Consumption (t/h)				Load	Consumption (t/h)			Actual Load	Consumption (t/h, kWh/h)			
			Steam			Fuel Gas		Cooling Water	Air	BFW		Steam (t/h)		Electr.	
		kW	LP	MP	HP		kW				kW	HP	MP	kWh/h	
	TRANSPORT	145577.00	16.80	24.48	180.00	1.69	235281.80	7365.11	15558.84	17.17	1047.97			1047.97	
R101	Regeneration Catofin	6612.59				0.75									
D101	Regeneration dryer	617.17				0.07									
K101	R101 vacuum compressor										2400.00			2400.00	
K102	R102 HP compressor										8015.00			8015.00	
K103	Refrigeration compressor										6233.88			6233.88	
K104	R102 recycle compressor										900.00		17.17		Self generated
P101	C101 reflux pump										7.17			7.17	
P102	R102 Water feed pump										40.91			40.91	
P103	C102 absorbent feed pump										96.40			96.40	
P201	C201 bottom pump										27.35			27.35	
P202	V201 inorganic phase pump										5.39			5.39	
P203	V201 organic phase pump										0.00			0.00	Duty = 2.3 W
P204	C202 bottom pump										24.87			24.87	
P205	C202 reflux pump										15.26			15.26	
P206	C202 top transfer pump										4.07			4.07	
P207	C202 product transfer pump										2.57			2.57	
P208	C203 bottom pump										14.95			14.95	
P209	V203 inorganic phase pump										0.49			0.49	
P210	V203 organic phase pump										9.13			9.13	
P211	C204 bottom pump										0.44			0.44	
P212	C204 reflux pump										0.49			0.49	
P213	C204 transfer pump										0.12			0.12	
P214	C205 bottom pump										2.28			2.28	
P215	C205 reflux pump										3.89			3.89	
P216	C205 transfer pump										2.09			2.09	
	TOTAL	**152807**	**17**	**24**	**180**	**3**	**235282**	**7365**	**15559**	**17**	**18855**	**0**	**17**	**17955**	
ility required		152807	0	24	180	3	235282	7365	15559	17	18855	0	0	17955	

Designers :	M.E. Brons	M.P. Hoff	Project ID Number	:	**CPD3242**
	J.H.M. Jansen	E.P.E. Seveke	Completion Date	:	December 16th 1999

Overall Component Mass Balance & Stream Heat Balance

STREAM nr.:		<001>	OUT	<002>	OUT	<101>	IN
Name:		COKES		WATER EFFLUENT		PROPANE FEED	
COMP	Mw	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s
HYDROGEN	2.00	0.00	0.00	0.00	0.00	0.00	0.00
METHANE	16.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHYLENE	28.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHANE	30.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHANOL	46.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPYLEN	42.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPANE	44.00	0.00	0.00	0.00	0.00	0.11	4.85
ACETONE	58.00	0.00	0.00	0.00	0.00	0.00	0.00
IPA	60.00	0.00	0.00	0.00	0.00	0.00	0.00
NPA	60.00	0.00	0.00	0.00	0.00	0.00	0.00
CYCLOHEXANE	84.00	0.00	0.00	0.00	0.00	0.00	0.00
N-HEXANE	86.00	0.00	0.00	0.00	0.00	0.00	0.00
DIPE	102.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	18.00	0.00	0.00	0.01	0.22	0.00	0.00
CARBON	12.00	0.01	0.12	0.00	0.00	0.00	0.00
Total		0.01	0.12	0.01	0.22	0.11	4.85
Enthalpy (kW)		6461		-3520		-13160	
Phase		S		L		L	
Pressure (bara)		0.50		1.00		15.00	
Temperature °C		650		35		35	

STREAM nr.:		<116>	OUT	<119>	IN	<120>	OUT
Name:		FROM ACC. C101		WATER MAKEUP		FROM P102	
COMP	Mw	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s
HYDROGEN	2.00	0.11	0.23	0.00	0.00	0.00	0.00
METHANE	16.00	0.01	0.08	0.00	0.00	0.00	0.00
ETHYLENE	28.00	0.00	0.06	0.00	0.00	0.00	0.00
ETHANE	30.00	0.00	0.07	0.00	0.00	0.00	0.00
ETHANOL	46.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPYLEN	42.00	0.00	0.08	0.00	0.00	0.00	0.00
PROPANE	44.00	0.00	0.01	0.00	0.00	0.00	0.00
ACETONE	58.00	0.00	0.00	0.00	0.00	0.00	0.00
IPA	60.00	0.00	0.00	0.00	0.00	0.00	0.00
NPA	60.00	0.00	0.00	0.00	0.00	0.00	0.00
CYCLOHEXANE	84.00	0.00	0.00	0.00	0.00	0.00	0.00
N-HEXANE	86.00	0.00	0.00	0.00	0.00	0.00	0.00
DIPE	102.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	18.00	0.00	0.00	0.17	10699.11	0.65	11.65
CARBON	12.00	0.00	0.00	0.00	0.00	0.00	0.00
Total		0.12	0.53	0.17	10699.11	0.65	11.65
Enthalpy (kW)		-763		-35537		-174522	
Phase		V		L		L	
Pressure (bara)		34.00		1.00		34.00	
Temperature °C		-33		25		78	

Overall Component Mass Balance & Stream Heat Balance-C'ntd

STREAM nr.:		<206>	OUT	<207>	OUT	<219>	IN
Name:		VENT		TO NPA STORAGE		MAKE UP	
COMP	Mw	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s
HYDROGEN	2.00	0.00	0.00	0.00	0.00	0.00	0.00
METHANE	16.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHYLENE	28.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHANE	30.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHANOL	46.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPYLEN	42.00	0.00	0.07	0.00	0.00	0.00	0.00
PROPANE	44.00	0.00	0.02	0.00	0.00	0.00	0.00
ACETONE	58.00	0.00	0.02	0.00	0.00	0.00	0.00
IPA	60.00	0.00	0.00	0.00	0.00	0.00	0.00
NPA	60.00	0.00	0.00	0.00	0.01	0.00	0.00
CYCLOHEXANE	84.00	0.00	0.00	0.00	0.00	0.00	0.01
N-HEXANE	86.00	0.00	0.00	0.00	0.00	0.00	0.00
DIPE	102.00	0.00	0.02	0.00	0.00	0.00	0.00
WATER	18.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON	12.00	0.00	0.00	0.00	0.00	0.00	0.00
Total		0.00	0.13	0.00	0.01	0.00	0.01
Enthalpy (kW)		-144		-17		-10	
Phase		V		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature °C		34		34		25	

STREAM nr.:		<222>	OUT	<230>	OUT
Name:		IPA PRODUCT		TO WATER TREATMENT	
COMP	Mw	kmol/s	kg/s	kmol/s	kg/s
HYDROGEN	2.00	0.00	0.00	0.00	0.00
METHANE	16.00	0.00	0.00	0.00	0.00
ETHYLENE	28.00	0.00	0.00	0.00	0.00
ETHANE	30.00	0.00	0.00	0.00	0.00
ETHANOL	46.00	0.00	0.00	0.00	0.00
PROPYLEN	42.00	0.00	0.00	0.00	0.00
PROPANE	44.00	0.00	0.00	0.00	0.00
ACETONE	58.00	0.00	0.00	0.00	0.00
IPA	60.00	0.10	5.78	0.00	0.00
NPA	60.00	0.00	0.00	0.00	0.00
CYCLOHEXANE	84.00	0.00	0.00	0.00	0.01
N-HEXANE	86.00	0.00	0.00	0.00	0.00
DIPE	102.00	0.00	0.00	0.00	0.00
WATER	18.00	0.00	0.01	0.05	0.99
CARBON	12.00	0.00	0.00	0.00	0.00
Total		0.10	5.79	0.06	1.00
Enthalpy (kW)		-34981		-15795	
Phase		L		L	
Pressure (bara)		1.00		1.00	
Temperature °C		30		30	

Overall Component Mass Balance & Stream Heat Balance-C'ntd

STREAM nr.:		<206>	OUT	<207>	OUT	<219>	IN
Name:		VENT		TO NPA STORAGE		MAKE UP	
COMP	Mw	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s
HYDROGEN	2.00	0.00	0.00	0.00	0.00	0.00	0.00
METHANE	16.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHYLENE	28.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHANE	30.00	0.00	0.00	0.00	0.00	0.00	0.00
ETHANOL	46.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPYLEN	42.00	0.00	0.07	0.00	0.00	0.00	0.00
PROPANE	44.00	0.00	0.02	0.00	0.00	0.00	0.00
ACETONE	58.00	0.00	0.02	0.00	0.00	0.00	0.00
IPA	60.00	0.00	0.00	0.00	0.00	0.00	0.00
NPA	60.00	0.00	0.00	0.00	0.01	0.00	0.00
CYCLOHEXANE	84.00	0.00	0.00	0.00	0.00	0.00	0.01
N-HEXANE	86.00	0.00	0.00	0.00	0.00	0.00	0.00
DIPE	102.00	0.00	0.02	0.00	0.00	0.00	0.00
WATER	18.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON	12.00	0.00	0.00	0.00	0.00	0.00	0.00
Total		0.00	0.13	0.00	0.01	0.00	0.01
Enthalpy (kW)		-144		-17		-10	
Phase		V		L		L	
Pressure (bara)		1.00		1.00		1.00	
Temperature °C		34		34		25	

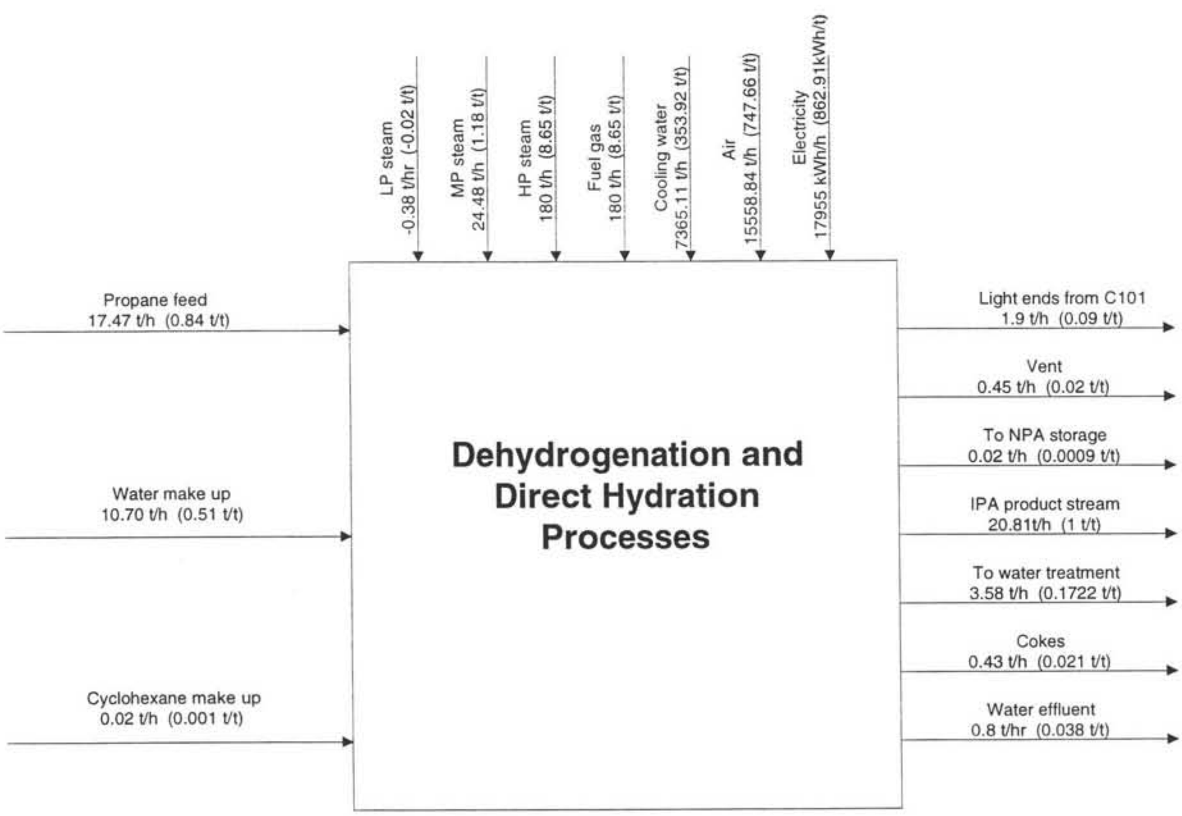
STREAM nr.:		<222>	OUT	<230>	OUT
Name:		IPA PRODUCT		TO WATER TREATMENT	
COMP	Mw	kmol/s	kg/s	kmol/s	kg/s
HYDROGEN	2.00	0.00	0.00	0.00	0.00
METHANE	16.00	0.00	0.00	0.00	0.00
ETHYLENE	28.00	0.00	0.00	0.00	0.00
ETHANE	30.00	0.00	0.00	0.00	0.00
ETHANOL	46.00	0.00	0.00	0.00	0.00
PROPYLEN	42.00	0.00	0.00	0.00	0.00
PROPANE	44.00	0.00	0.00	0.00	0.00
ACETONE	58.00	0.00	0.00	0.00	0.00
IPA	60.00	0.10	5.78	0.00	0.00
NPA	60.00	0.00	0.00	0.00	0.00
CYCLOHEXANE	84.00	0.00	0.00	0.00	0.01
N-HEXANE	86.00	0.00	0.00	0.00	0.00
DIPE	102.00	0.00	0.00	0.00	0.00
WATER	18.00	0.00	0.01	0.05	0.99
CARBON	12.00	0.00	0.00	0.00	0.00
Total		0.10	5.79	0.06	1.00
Enthalpy (kW)		-34981		-15795	
Phase		L		L	
Pressure (bara)		1.00		1.00	
Temperature °C		30		30	

Process Yields - Streams							
Name	Ref. Stream	kg/s IN	kg/s OUT	t/h IN	t/h OUT	t/t IPA Product IN	t/t IPA Product OUT
Propane feed	<101>	4,85		17,47		0,84	
Make up water	<119>	2,97		10,70		0,51	
Make up cyclo hexane	<219>	0,01		0,02		0,001	
Light ends from C101	<116>		0,53		1,90		0,09
Vent	<206>		0,13		0,45		0,02
To NPA storage	<207>		0,01		0,02		0,001
IPA product stream	<222>		5,78		20,81		1,00
To water treatment	<230>		1,00		3,58		0,17
Cokes	<001>		0,12		0,43		0,02
Water effluent	<002>		0,22		0,80		0,04
Total		7,830	7,778	28,188	28,000	1,3543	1,3453

830
778
052

Δ = 188 kg/h

Process Yields - Utilities							
Name	Ref. Stream	kg/s	kW	t/h	kWh/h	t/t IPA Product	kWh/t IPA Prod.
LP Steam (1)	-	-0,11		-0,38		-0,02	
MP Steam	-	6,80		24,48		1,18	
HP Steam	-	50,00		180,00		8,65	
Fuel Gas	-	0,70		2,51		0,12	
Cooling Water	-	2045,86		7365,11		353,92	
Air	-	4321,90		15558,84		747,66	
Electricity	-		17955,00		17955,00		862,81
Remarks: (1) LP steam is net produced in the process							



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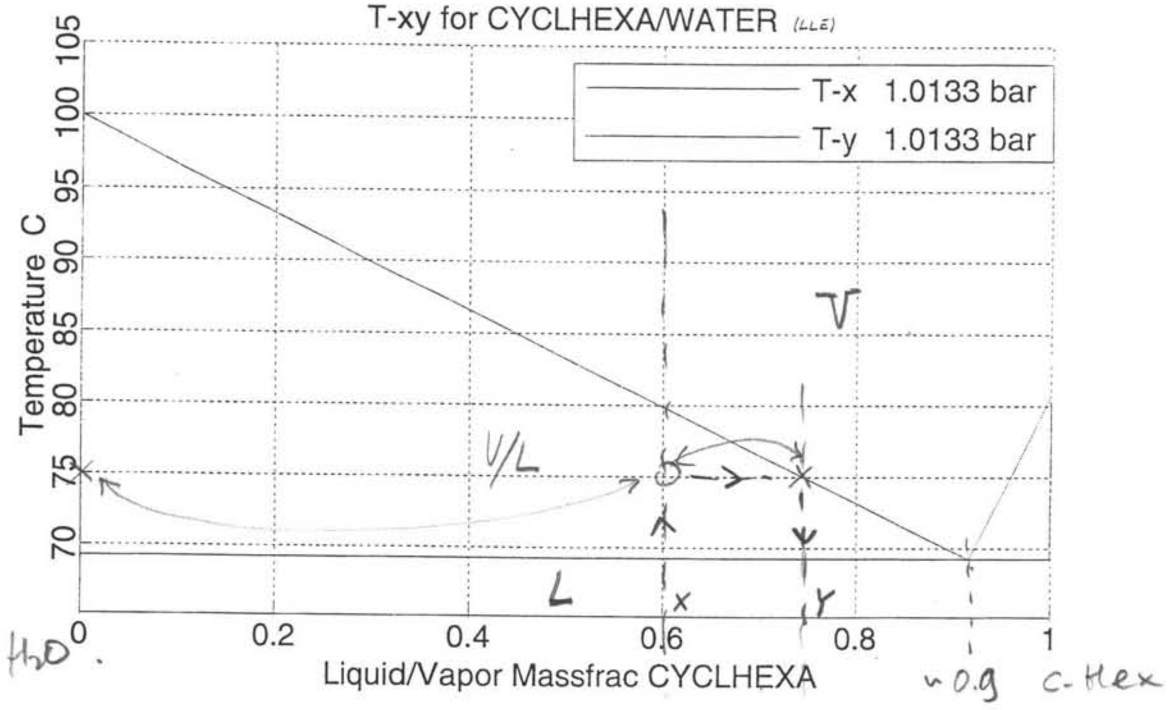
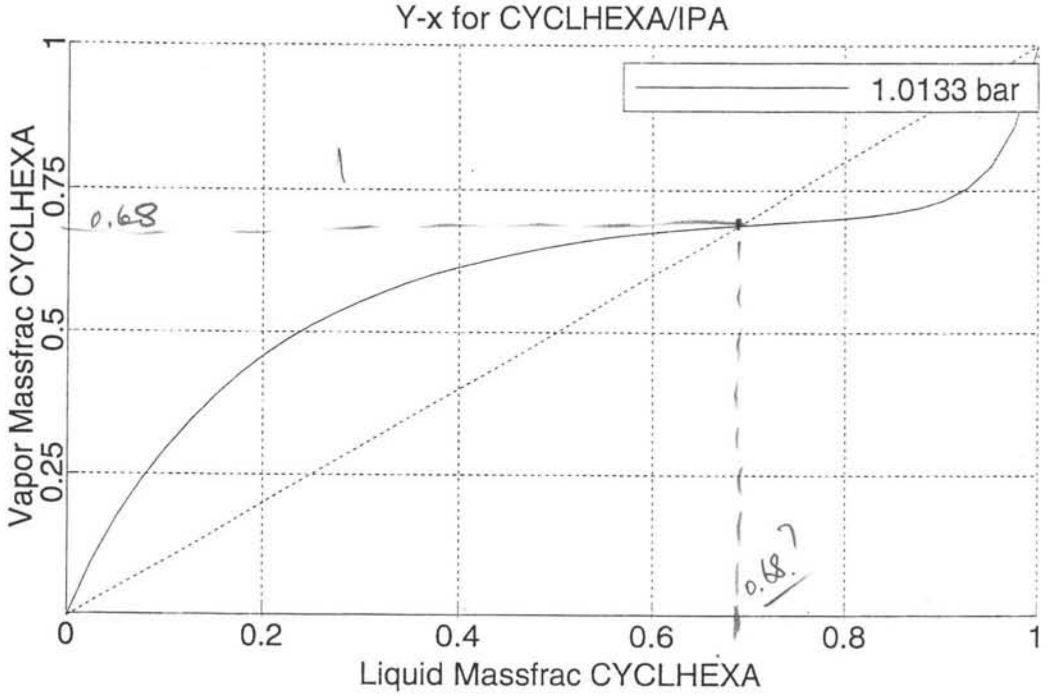


REF: DECHEMA

		REF	TEMP DEG C	SOLUBILITIES MOLE PCT OF (1)IN(2)	(2)IN(1)	MODEL PARAMETERS IN KELVIN (PROCESSED DATA) AND LITERATURE CITATION (FOR ORIGINAL DATA) UNIQUAC A12	A21	NRTL(ALPHA=.2) A12	A21
(1) C6H12	CYCLOHEXANE					R = 4.0464	Q = 3.240		
(2) H2O	WATER					R = 0.9200	Q = 1.400		
		*	25.0	0.120E-02	0.580E-01	1247.3	540.36	1806.5	2840.8
		1	14.00		0.2335E-01	TARASSENKOV D.N., POLOSHINZEVA E.N.			
		1	19.00		0.4670E-01	BER.DTSCH.CHEM.GES. 65(1932)184			
		1	28.50		0.7004E-01				
		1	32.50		0.9336E-01				
		1	38.00		0.1447				
		1	53.00		0.2332				
		2	20.00		0.4669E-01	BLACK C., JORIS G.G., TAYLOR H.S. J.CHEM.PHYS. 16(1948)537			
		3	16.00	0.1337E-02		DURAND R. C.R.HEBD.SEANCES ACAD.SCI. 226(1948)409			
		4	25.00	0.1177E-02		MCAULIFFE C. NATURE(LONDON) 200(1963)1092			
		5	10.00		0.3129E-01	ENGLIN B.A. ET AL.			
		5	20.00		0.5697E-01	KHIM.TEKHNOL.TOPL.MASEL (1965)9,42			
		5	30.00		0.9057E-01				
		5	40.00		0.1479				
		5	50.00		0.2285				
		6	25.00		0.2594E-01	JOHNSON J.R., CHRISTIAN S.D., AFFSPRUNG H.E. J.CHEM.SOC. A(1966)77			
		7	25.00		0.3728E-01	RODDY J.W., COLEMAN C.F. TALANTA 15(1968)1281			

ASPEN



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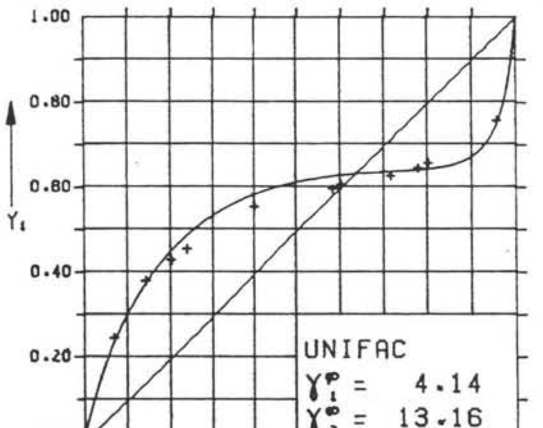

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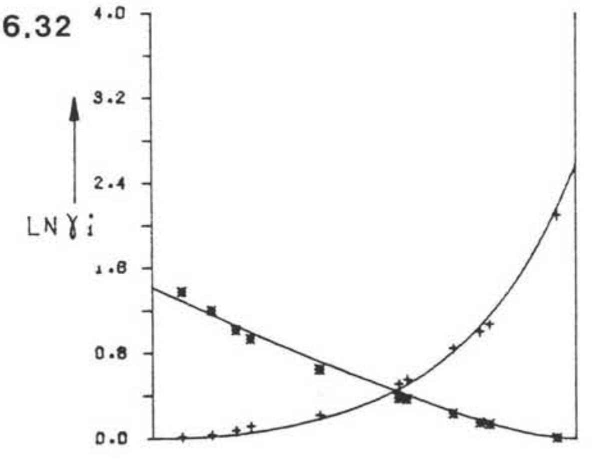
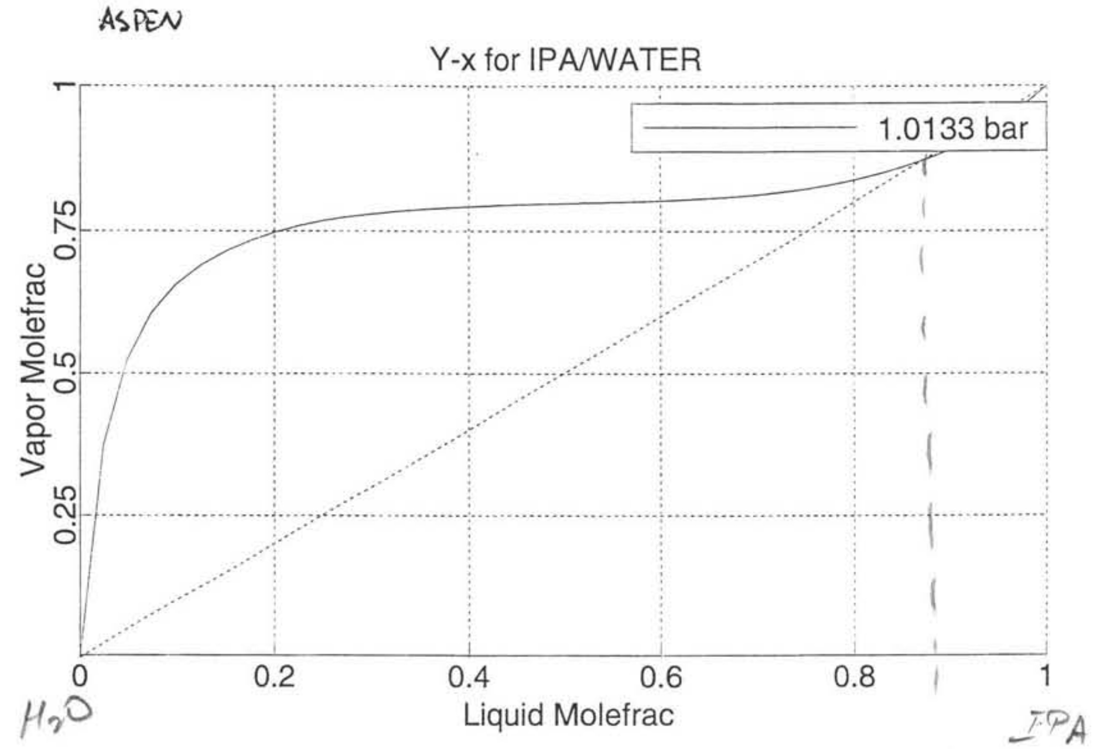
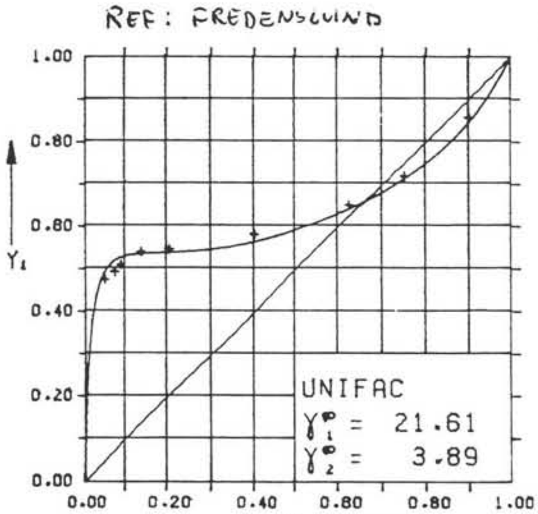
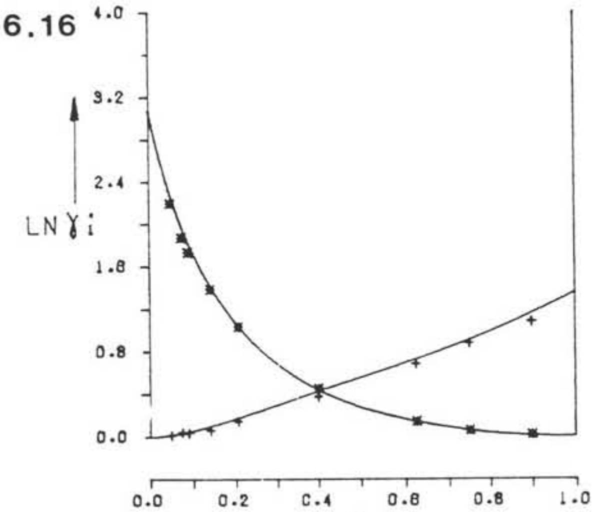


Cyclohexane(1)-2-Propanol(2) at 760 mm Hg
L.A.J. Verhoeye, J.Chem.Eng. Data, 13(1968)462.

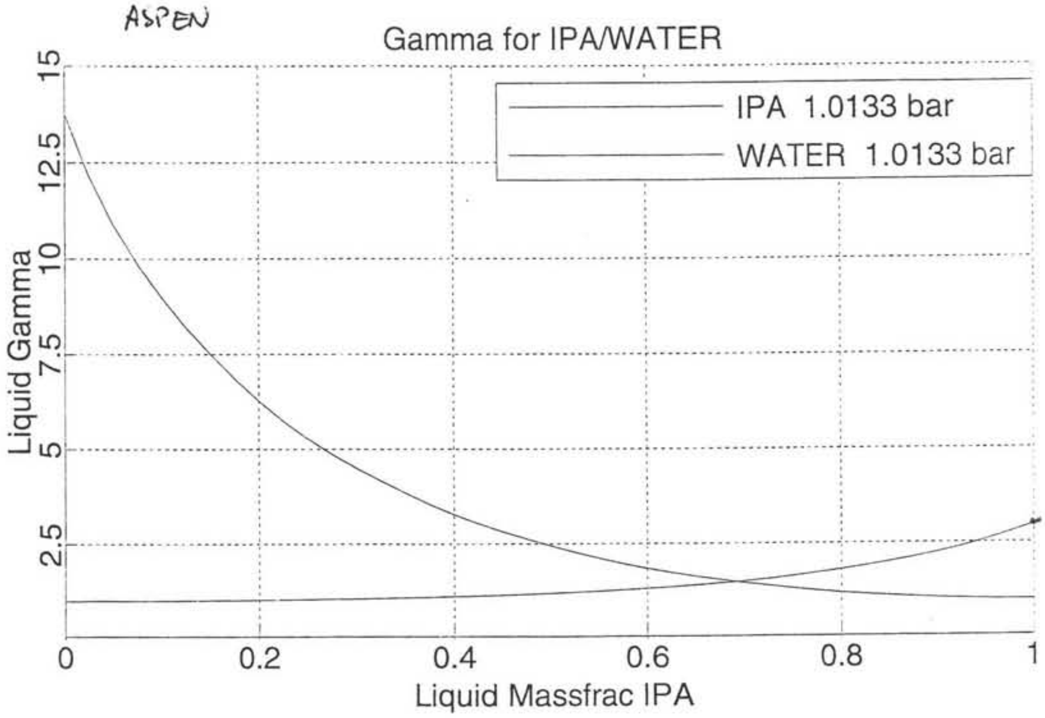






2-Propanol(1)-Water(2) at 760 mm Hg
L.A.J. Verhoeye, J.Chem.Eng. Data, 13(1968)462.

ASPEN



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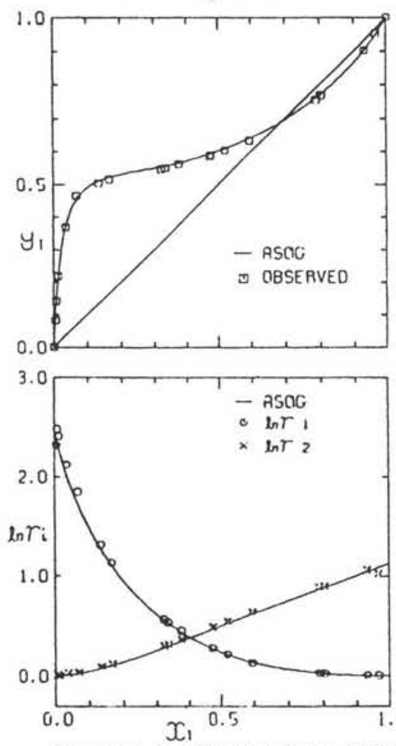
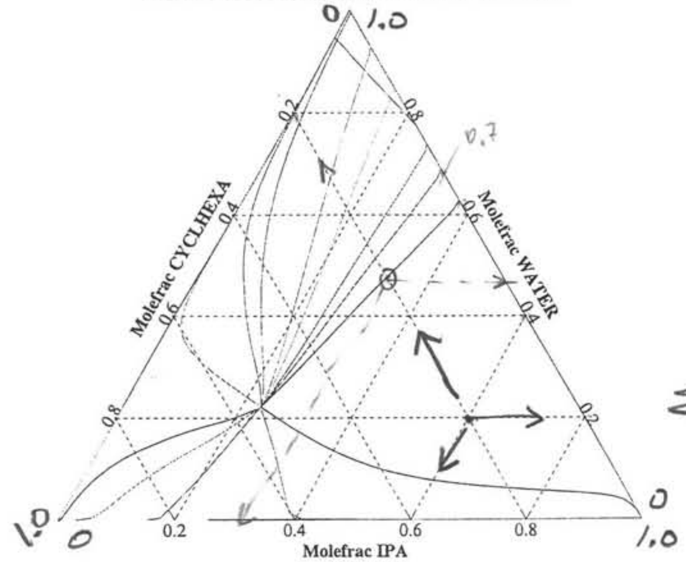


x_1	y_1	t(°C)	Δy_1	Δt
0.000	0.000	100.00		
0.005	0.081	97.57	-0.013	0.19
0.007	0.141	96.20	0.007	-0.01
0.013	0.219	93.66	0.010	-0.22
0.036	0.369	87.84	0.004	-0.65
0.068	0.465	84.28	0.014	-0.96
0.133	0.504	82.84	-0.005	-0.12
0.165	0.515	82.52	-0.006	0.01
0.320	0.546	81.52	-0.009	-0.01
0.334	0.549	81.45	-0.008	-0.02
0.375	0.562	81.19	-0.004	-0.09
0.472	0.586	80.77	-0.007	-0.10
0.476	0.589	80.73	-0.005	-0.13
0.520	0.603	80.58	-0.006	-0.12
0.595	0.633	80.52	-0.007	0.04
0.788	0.755	80.46	-0.005	-0.05
0.802	0.768	80.55	-0.003	-0.01
0.930	0.901	81.32	-0.001	-0.07
0.966	0.952	81.85	0.003	0.07
1.000	1.000	82.39		
		MEAN	0.006	0.16

KOJIMA.K.,K.OCHI.Y.NAKAZAWA:KAGAKUKOUGAKU.VOL.32.441(1966).

ASPEN

Residue curve for IPA/WATER/CYCLHEXA



$\Sigma = 1.0$.

REF : DECHEMA

(1) H2O WATER

(2) C3H8O 2-PROPANOL

(3) C6H12 CYCLOHEXANE

VERHOEYE L.A.J.
J.CHEM.ENG.DATA 13(1968)462

TEMPERATURE = 25.0 DEG C TYPE OF SYSTEM = 1

EXPERIMENTAL TIE LINES IN MOLE PCT

LEFT PHASE			RIGHT PHASE		
(1)	(2)	(3)	(1)	(2)	(3)
93.903	5.999	0.098	0.462	2.631	96.908
87.677	12.074	0.249	1.336	12.679	85.985
83.905	15.799	0.295	2.174	18.245	79.581
76.401	21.812	1.787	5.676	23.966	70.358
70.379	26.845	2.776	8.022	27.409	64.565
60.317	33.443	6.240	13.168	31.909	54.923

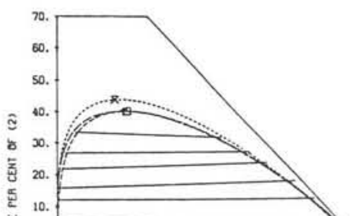
SPECIFIC MODEL PARAMETERS IN KELVIN

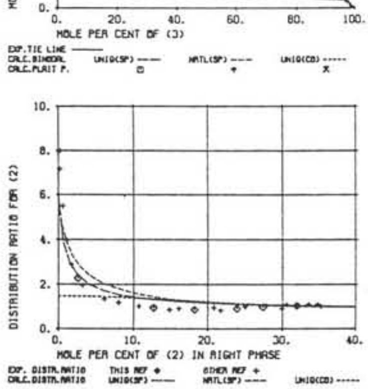
	UNIQUAC		NRTL(ALPHA=.2)	
I J	AIJ	AJI	AIJ	AJI
1 2	-726.89	-244.80	-1600.8	-541.64
1 3	927.06	1009.2	1669.5	874.98
2 3	-214.76	-432.15	-9.0112	-1805.8

R1 = 0.9200 R2 = 2.7791 R3 = 4.0464
Q1 = 1.400 Q2 = 2.508 Q3 = 3.240

MEAN DEV. BETWEEN CALC. AND EXP. CONC. IN MOLE PCT

UNIQUAC (SPECIFIC PARAMETERS)	1.58
NRTL (SPECIFIC PARAMETERS)	2.12
UNIQUAC (COMMON PARAMETERS)	1.87





614
$C_3H_8O-C_6H_{12}$

Reaction kinetics for the propylene reactor

The compositions (wt%) of the by-products are summarised in table 5.2-1. Values are as described in literature [I] at four temperatures (hydrogen is not included).

(in reactor effluent)

Table 5.2-1. By-product composition as reported in literature [I].

Temperature[°C]	500	550	600	650
Ethylene [wt%]	0.04	0.12	0.40	1.00
Ethane [wt%]	0.09	0.40	2.00	2.10
Methane [wt%]	0.18	0.80	3.00	3.50

Total. 0.31 1.32 5.40 6.60

Circumstances as described in literature and those discussed in this report are not exactly the same, but they probably are the best estimate. Operation temperature of the dehydrogenation reaction was set on 650°C because high conversion can then be achieved.

The figures of table 5.2-1 are used to calculate the selectivity of the reactions. Next a brief summary of the manner of calculation is discussed per reaction (see chapter 4):

Reaction 5:
Ethane can only be formed by reaction (5).

Reaction ~~4~~ 3:
Because ethane can only be produced by reaction (5), it can be seen that the total amount of ethylene formed by reaction ~~(4)~~ (3) equals the amount of ethylene in the product stream plus the amount of ethylene used to produce ethane.

Reaction ~~3~~ 4:
Because of the foregoing, one knows the amount of methane produced by reaction ~~(4)~~ (3). From the product composition, table 5.2-1, the total amount of methane can be found. From these two facts it can be concluded how much propane is converted by reaction ~~(3)~~ (4).

Reaction 2:
From the product, table 5.2-1, and from the amount of coke~~s~~ formed by reaction (4) it can be calculated how much coke~~s~~ is formed by reaction (2).

These calculations lead to the values as reported in table 5.2-2.

Table 5.2-2. Summary of the calculated selectivitiess, conversions and yields.

	Selectivity	Conversion	Yield
Reaction 1	0.879	0.900	0.791
Reaction 2	0.067	0.900	0.061
Reaction 3	0.003	0.900	0.003
Reaction 4	0.051	0.900	0.045

An interesting question is whether the amount of cokes formed is enough to make up for the heat consumed by the reactions. This calculation will be shown here.

Table 5.2-3. Summary of the propane conversion.

	kmol/h
Propane in ~~C~~ R101	517.43
Propane out ~~C~~ R101	122.11
Amount of propane converted	395.32

Table 5.2-4. Calculation of the heat of reaction in the propylene reactor.

		Enthalpy of reaction [kJ/mol]	Enthalpy of reaction [kJ/kg]	Selectivity [--]	Energy consumed [kJ/mol]
1	Propane<=>Propylene+Hydrogen	124.7	2834	0.6286	78
2	Propane<=>Propylene+Hydrogen	104.7	2380	0.2708	28
3	Propane => 2CH4 + C	-44.1	-1002	0.0061	0
4	Propane => Methane + Ethylene	-7.2	-164	0.0946	-1
5	Ethylene H2 <=> Ethane	-46.3	-1654	0.0046	0
	Total energy consumed				106

Minimal amount heat needed [MJ/h]		41734
Cokes formed [kmol/h]	36.00	
heat of combustion of cokes [kJ/mol]	-393.5	
Efficiency [%]	80	
Energy released by combustion [MJ/h]		-17709
Heat needed [MJ/h]		24024
Heat of combustion of fuel gas [MJ/kg]	40	
Efficiency [%]	80	
Fuel gas needed [kg/h]		750

In a number of reports it has been described that regeneration and heating by burning of the cokes is preferred over the use of fuel gas, due to efficiency reasons. Therefore it may be interesting to know the minimal amount of cokes needed for the reactor to be self-sufficient. This can be calculated be done by solving the calculation below:

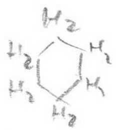
Minimal amount heat needed [MJ/h]		41734
Energy released by combustion of cokes [MJ/h]		-41734
Heat needed [MJ/h]		0
heat of combustion of cokes [kJ/mol]	-393.5	
Efficiency [%]	80	
Cokes formed [kmol/h]		106.06

Therefore it can be concluded that if the reported amount of cokes is formed, a substantial amount of fuel gas is needed (751 kg/h). Only if (at least) three times as much cokes is formed no fuel gas is needed.

(I) G.F. Froment and K.C. Waugh et al., Reaction kinetics and the development of catalytic processes: Proceedings of the international symposium, Brugge, Belgium, April 19-21, 1999 [paper: Kinetic Based Deactivation Modelling of an Isothermal Propane Dehydrogenation reactor].

PURE COMPONENT PROPERTIES

Component Name		Technological Data					Medical Data		Notes
Design	Systematic	Formula	Molecular Weight g/mol	Boiling Point (1) °C	Melting Point (1) °C	Density Liquid (2) kg/m^3	MAC Value mg/m^3	LD50 Oral g	
Paraffins									
Methane	Methane	CH4	16.04	-161.5	-182.5	160.0			(3)
Ethane	Ethane	C2H6	30.07	-88.6	-182.8	544.6			(3)
Propane	Propane	C3H8	44.10	-42.1	-187.6	585.3		none	(4)
IPA	2-Propanol	C3H8O	60.10	82.3	-89.5	785.5			(5)
NPA	1-propanol	C3H8O	60.10						
Olefins									(6)
Ethylene	Ethylene	C2H4	28.05	-103.7	-169.0	567.8			(7)
Propylene	1-Propene	C3H6	42.08	-47.6	-185.2	505.0	none	none	
Other hydrocarbons									
Acetone	Acetone	C3H6O	58.08	56.3	-94.7	913.1			(8)
Cyclohexane	Cyclohexane	C6H12	84.16	80.7	6.5	789.4			(9)
Ethers									
DIPE	propane-2,2-oxybis	C6H14O	102.18	68.5	-86.8	724.1		8470 mg/kg	
Inorganic Compounds									
Carbon	Charcoal, activated	C	12.01				none	none	
Hydrogen	Hydrogen	H2	2.02	-252.9	-259.3	88.0	none	none	
Water	Water	H2O	18.02	100.0	0.0	997.0		none	(10)



Notes:
(1) At 101.3 kPa
(2) Density at 15.56 °C, unless specified otherwise
(3) Density at -89 °C
(4) Density at -45 °C from H2O at 4 °C
(5) Density at 20 °C
(6) Density at -104 °C
(7) Density at 25 °C
(8) Density at 20 °C from H2O at 4 °C
(8) Density at -94.7 °C
(9) Density at 6.54 °C
(10) Values at indicated purity

n.a. = not available

Designers :	M.E. Brons	M.P. Hoff	Project ID Number :	**CPD3242**
	J.H.M. Jansen	E.P.E.Seveke	Completion Date :	December 22nd 1999

Delft, University of Technology
Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. : NAME :	R101 Propylene reactor	R102 IPA reactor	C101 Deethanizer	C102 Washing column
	Horizontal	Horizontal	Tray column	Absorber
Temperature[1] [°C][1] :	650/650	158/175	-33/77	35/31
Pressure [bara][1] :	0.66/0.33	34/32.86	34.8/35.4	29.0/29.1[3]
Volume [m^3] :	100.93	142.63	45.67	45.67
Diameter [m] :	4.62	3.57	2.6	2.73
L or H [m] :	6.01	14.27	35.5	8.6
Internals				
- Tray Type :	n.a.	n.a.	Sieve	0
- Tray Number :	n.a.	n.a.	51	15 (theor.)
- Fixed Packing				
Type :	n.a.	n.a.	n.a.	n.a.
Shape :	n.a.	n.a.	n.a.	Structured
-Catalyst				
Type :	Chromium oxide	H3PO4	n.a.	n.a.
Shape :	Spherical	Spherical	n.a.	n.a.
Number				
- Series :	-	-	-	-
- Parallel :	3	1	1	1
Materials of Construction[2]				
-Column :	C.S.[2]	C.S.	C.S.	C.S.
-Tray :				

Remarks:
1. Top / bottom temperature or pressure.
2. CS is carbon steel.
3. Based on recommended flow scheme

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

Delft, University of Technology
Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. : NAME :	C201 Light ends column	C202 IPA CBM column	C203 IPA drying column	C204 Cyclohexane recovery column
	Tray column	Tray column	Tray column	Tray column
Temperature[1] [°C][1] :	51/86	86/95	75/82.3	78.6/100
Pressure [bara][1] :	0.34/1.39	0.66/1.50	0.80/1.25	0.82/1.17
Volume [m³] :	0	0	0	0
Diameter [m] :	3.03	2.84	2.6	1.34
L or H [m] :	31.1	39.7	23.29	19
Internals				
- Tray Type :	Sieve	Sieve	Sieve	Sieve
- Tray Number :	45	58	50	35
- 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.
Number				
- Series :	-	-	-	-
- Parallel :	1	1	1	1
Materials of Construction[2]				
-Column :	C.S.[2]	C.S.	C.S.	C.S.
-Tray :				

Remarks:

4. Top / bottom temperature or pressure.
5. CS is carbon steel.

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. : NAME :	C205 NPA recovery column	V101 C101 accumulator	V102 HP Separator	V201 Light ends column phase separator
	Tray column	Horizontal	Horizontal	Horizontal
Temperature[1] [°C][1] :	80/80	-33	90	34
Pressure [bara][1] :	0.77/1.11	34.8	32/30	1
Volume [m³] :	0	16.83	31.91	46.63
Diameter [m] :	2.68	1.75	2.5	2.5
L or H [m] :	19	7	6.5	9.5
Internals				
- Tray Type :	Sieve	n.a.	n.a.	n.a.
- Tray Number :	28	n.a.	n.a.	n.a.
- 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.
Number				
- Series :	-	-	-	-
- Parallel :	1	1	1	1
Materials of Construction[2]				
-Column :	C.S.[2]	C.S.	C.S.	C.S.
-Tray :				

Remarks:

6. Top / bottom temperature or pressure.
7. CS is carbon steel.

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. : NAME :	V202 C202 accumulator	V203 Drying column phase separator	V204 C204 accumulator	V205 C205 accumulator
	Horizontal	Horizontal	Horizontal	Horizontal
Temperature[1] [°C][1] :	80	75	90	80
Pressure [bara][1] :	0.66	1	0.82	0.77
Volume [m^3] :	47.86	65.34	2.39	13.07
Diameter [m] :	2.5	2.75	0.9	1.6
L or H [m] :	9.75	11	3.75	6.5
Internals				
- Tray Type :	n.a.	n.a.	n.a.	n.a.
- Tray Number :	n.a.	n.a.	n.a.	n.a.
- 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.
Number				
- Series :	-	-	-	-
- Parallel :	1	1	1	1
Materials of Construction[2]				
-Column :	C.S.[2]	C.S.	C.S.	C.S.
-Tray :				

Remarks:

8. Top / bottom temperature or pressure.
9. CS is carbon steel.

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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Process Systems Engineering

REACTORS, COLUMNS & VESSELS - SUMMARY

EQUIPMENT NR. :	T201 A/B/C/D	T202	T203	T204	D101
NAME :	Run down tank	Off spec tank	On spec tank	Sales tank	Dryer in overall recycle
	Vertical	Vertical	Vertical	Vertical	Horizontal
Temperature[1] [°C][1] :	30	30	30	30	35
Pressure [bara][1] :	1	1	1	1	15[3]
Volume [m^3] :	224	4902	60080	28008	76.68
Diameter [m] :	7.8	20.74	52.96	41.06	2.9
L or H [m] :	4.68	14.51	27.27	21.14	11.6
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.	Drying agent
Shape :	n.a.	n.a.	n.a.	n.a.	Spherical
-Catalyst					
Type :	n.a.	n.a.	n.a.	n.a.	
Shape :	n.a.	n.a.	n.a.	n.a.	
Number					
- Series :	-	-	-	-	-
- Parallel :	4	1	1	1	2
Materials of Construction[2]					
-Column :	C.S.[2]	C.S.	C.S.	C.S.	C.S.
-Tray :					

Remarks:
10. Top / bottom temperature or pressure.
11. CS is carbon steel
12. Based on recommended flow scheme

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

Delft, University of Technology
Process Systems Engineering

HEAT EXCHANGERS & FURNACE- SUMMARY

EQUIPMENT NR. : NAME :	E101 R101 Feed-Effluent A	E102 R101 Feed-Effluent B	E103 R101 Feed-Effluent C	E104 R101 Feed-Effluent D	E105 R101 Effluent - Steam
Type :	HTXR	HTXR	HTXR	HTXR	HTXR
Substance - Tubes : - Shell :	 Propylene Propane/ Propylene	 Propylene Propane/ Propylene	 Propylene Propane/ Propylene	 Propylene Propane/ Propylene	 Steam Propylene
Duty [kW] :	1625	2816	2562	2176	12498
Heat exchange area [m^2] :	92	204	217	189	62
Number of passes - Tubes : - Shell :	 2 1	 2 1	 2 1	 2 1	 2 1
Pressure [bara] - Tubes : - Shell :	 0,3 2,0	 0,3 2,0	 0,3 2,0	 0,3 2,0	 15,0 0,3
Temperature [°C] - Tubes In : Out : - Shell In : Out :	 147,6 100,0 4,7 57,0	 230,0 147,6 57,0 147,6	 305,0 230,0 147,6 230,0	 368,7 305,0 230,0 300,0	 100,0 300,0 650,0 368,7
Special materials of construction -Tubes : -Shell :	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.
Other :					

Remarks:

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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Process Systems Engineering

HEAT EXCHANGERS & FURNACE- SUMMARY

EQUIPMENT NR. NAME	: :	E106 K101 Effluent	E107 C203 Reboiler	E108 C101 Feed	E109 C101 Condenser	E110 C101 Condenser Heat pump
Type	:	Cooler	HTXR Reboiler	Cooler	Condenser	Cooler
Substance - Tubes - Shell	 : :	 Propylene Air	 Propylene IPA	 Propylene Cooling water	 H2/Light ends Propylene	 Propylene Cooling water
Duty [kW]	:	2986	825	6830	5394	11599
Heat exchange area [m^2]	:	50	49	522	1065	394
Number of passes - Tubes - Shell	 : :	 1 n.a.	 2 1	 4 2	 2 1	 2 1
Pressure [bara] - Tubes - Shell	 : :	 0,7 1,0	 36,0 1,0	 35.5 1,0	 34,8 1,0	 36,0 1,0
Temperature [°C] - Tubes In Out - Shell In Out	 : : : :	 180,0 80,0 30,0 60,0	 180,3 155,0 82,3 82,3	 155,0 26,9 25,0 40,0	 -33,0 -33,0 -47,6 -40,0	 144,8 40,0 25,0 40,0
Special materials of construction -Tubes -Shell	 : :	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.
Other	:					
Remarks:						

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

Delft, University of Technology
Process Systems Engineering

HEAT EXCHANGERS & FURNACE- SUMMARY

EQUIPMENT NR. NAME	: :	E111 C101 Reboiler	E112 R102 Feed heater	E113 C201 Reboiler	E114 R102 Effluent cooler	E115 C102 Feed cooler A
Type	:	Reboiler	Heater	Reboiler	Cooler	Cooler
Substance						
- Tubes	:	Propylene	Propylene/ IPA	IPA/ Propylene	IPA/ Propylene	IPA/ Propylene
- Shell	:	LP Steam	MP Steam	IPA/water	Air	Cooling water
Duty [kW]	:	6240	14061	20201	25824	36459
Heat exchange area [m^2]	:	139	475	945	614	1706
Number of passes						
- Tubes	:	2	2	2	1	4
- Shell	:	1	1	1	n.a.	2
Pressure [bara]						
- Tubes	:	35,4	34,0	32,9	32,9[1]	30,2[1]
- Shell	:	4,7	15,0	1,4	1,0	1,0
Temperature [°C]						
- Tubes In	:	77,1	139,7	175,0	141,3	90,0
Out	:	77,1	158,0	141,3	90,0	40,0
- Shell In	:	196,8	200,0	85,6	30,0	25,0
Out	:	100,0	175,0	85,6	60,0	40,0
Special materials of construction						
-Tubes	:	C.S.	C.S.	C.S.	C.S.	C.S.
-Shell	:	C.S.	C.S.	C.S.	C.S.	C.S.
Other	:					

Remarks:
1. Based on recommended flow scheme

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

Delft, University of Technology
Process Systems Engineering

HEAT EXCHANGERS & FURNACE- SUMMARY

EQUIPMENT NR. : NAME :	E116 C102 Feed cooler B	E201 V201 Feed cooler	E202 DELETED	E203 C202 Reboiler	E204 C202 Condenser
Type :	Cooler	Cooler		HTXR Reboiler	Condenser
Substance - Tubes : - Shell :	 IPA/ Propylene Cooling water	 Cooling water Water/IPA		 HP Steam Water	 IPA/water Air
Duty [kW] :	2968	2615		58308	50799
Heat exchange area [m^2] :	645	185		742	3103[2]
Number of passes - Tubes : - Shell :	 2 1	 4 2		 2 1	 1 n.a.
Pressure [bara] - Tubes : - Shell :	 29,7[1] 1,0	 1,0 1,0		 18,0 1,5	 0,7 1,0
Temperature [°C] - Tubes In : Out : - Shell In : Out :	 40,0 30,0 25,0 30,0	 25,0 40,0 79,8 34,3		 225,0 175,0 95 95	 80,0 80,0 30,0 60,0
Special materials of construction -Tubes : -Shell :	 C.S. C.S.	 C.S. C.S.		 C.S. C.S.	 C.S. C.S.
Other :					

Remarks:

1. Based on recommended flow scheme
2. In practice 6 air coolers would be placed

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

Delft, University of Technology
Process Systems Engineering

HEAT EXCHANGERS & FURNACE- SUMMARY

EQUIPMENT NR.	:	E205	E206	E207	E208	E209
NAME	:	C205 reboiler	C205 Condenser	Recycle to C101	C203 Condenser	C204 Reboiler
Type	:	Reboiler	Condenser	Cooler	Condenser	Reboiler
Substance						
- Tubes	:	HP Steam	IPA/water	Cooling water	Cyclohex-ane/water	LP Steam
- Shell	:	NPA	Air	Water	Air	Water
Duty [kW]	:	46950	46950	6809	690	5100
Heat exchange area [m^2]	:	496	2390[1]	473	51	380
Number of passes						
- Tubes	:	2	1	4	1	1
- Shell	:	1	n.a.	2	n.a.	2
Pressure [bara]						
- Tubes	:	18,0	0,8	1,0	0,8	4,7
- Shell	:	1,1	1,0	1,0	1,0	1,0
Temperature [°C]						
- Tubes In	:	225,0	80,0	25,0	74,7	196,8
Out	:	175,0	80,0	40,0	74,7	101,0
- Shell In	:	80,0	30,0	95,0	30,0	100,0
Out	:	80,0	60,0	30,0	60,0	100,0
Special materials of construction						
-Tubes	:	C.S.	C.S.	C.S.	C.S.	C.S.
-Shell	:	C.S.	C.S.	C.S.	C.S.	C.S.
Other	:					

Remarks:
1. In practice five air coolers would be placed

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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Process Systems Engineering

HEAT EXCHANGERS & FURNACE- SUMMARY

EQUIPMENT NR. : NAME :	E210 C204 Condenser	E211 C201 Top Feed Cooler	E212 C204 Bottom cooler	E213 C203 Product cooler	F101 R101 Feed Furnace
Type :	Condenser	Cooler	Cooler	Cooler	Furnace
Substance - Tubes : - Shell :	 Cyclohex- ane Air	 Cooling water Water	 Cooling water Water	 Cooling water IPA	 Propane/ Propylene Fuel gas
Duty [kW] :	5000	12899	197	4765	14918
Heat exchange area $[m^2]$:	320	634	11	350	497
Number of passes - Tubes : - Shell :	 1 n.a.	 4 2	 2 1	 2 1	 1 1
Pressure [bara] - Tubes : - Shell :	 0,8 1,0	 1,0 0,8	 1,0 1,2	 1,0 1,3	 2[1] 1
Temperature [°C] - Tubes In : Out : - Shell In : Out :	 78,6 78,6 30,0 60,0	 25,0 30,0 95.0 34.0	 25,0 30,0 100,0 30,0	 25,0 30,0 82,3 30,0	 300 650 900 900
Special materials of construction -Tubes : -Shell :	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.	 C.S. C.S.
Other :					

Remarks:
1. Based on recommended flow scheme

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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COMPRESSORS SUMMARY

EQUIPMENT NR. NAME	: :	K101 R101 vacuum compressor	K102 R102 HP compressor	K104 R102 recycle compressor
Type Number	: :	Reciprocating 1	Centrifugal 1	Axial flow 1
Medium transferred	:	Propane/Propylene/ Light ends	Propane/Propylene/ Light ends	Propane/Propylene/ Water/IPA
Capacity [m^3/s].	:	24.2	0.42	1.17
Density [kg/m^3]	:	0.056403907	3.253968254	69.90
Pressure In [bara] Out [bara] difference [bara]	 : : :	 0.2 0.67 0.47	 0.57 36 35.43	 29[2] 34 5
Temperature Out [°C]	 :	 180	 180	 224
Power [kW] - Theor. - Actual - Efficiency	 : : :	 1728.32 2400.44 72%	 5770.50 8014.58 72%	 648.00 900.00 72%
Number - Theor. - Actual	 : :	 1 2	 1 2	 1 2
Special Materials of Construction	:	C.S.	C.S.	C.S.

Remarks:
1. K103 in closed heat pump cycle is left out of design, but is included in the economical analysis.
2. Based on recommended flow scheme

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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PUMPS - SUMMARY

EQUIPMENT NR. NAME	: :	P101 C101 Reflux pump	P102 R102 Water feed pump	P103 C102 Absorbent feed pump[1]	P201 C201 Bottom pump
Type Number	: :	Centrifugal 1	Multi stage centrifugal 1	Centrifugal 1	Centrifugal 1
Medium transferred	:	Propane/Propylen e	Water	Water	Water/IPA
Capacity [m^3/s].	:	2.76E-02	8.68E-03	2.62E-02	1.16E-01
Density [kg/m^3]	:	391.0	1000.0	1000.0	898.1
Pressure In [bara] Out [bara] difference [bara]	 : : :	 34.8 36.8 2.0	 1.0 34.0 33.0	 1.0 29.0 28.0	 1.4 3.4 2.0
Temperature In (= Out) [°C]	 :	 -33	 77.6	 30	 86
Power [kW] - Theor. - Actual - Efficiency	 : : :	 5.52 7.17 77%	 28.64 40.91 70%	 73.26 96.40 76%	 23.25 27.35 85%
Number - Theor. - Actual	 : :	 1	 1	 1	 1
Special Materials of Construction	:	CS	CS	CS	CS

Remarks:

1. Based on recommended flow scheme

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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Process Systems Engineering

PUMPS - SUMMARY

EQUIPMENT NR. : NAME :	P202 V201 Inorganic phase pump	P203 V201 Organic phase pump	P204 C202 Bottom pump	P205 C202 Reflux pump
Type : Number :	Centrifugal 1	Centrifugal 1	Multi stage centrifugal 1	Centrifugal 1
Medium transferred :	Water/IPA	NPA	Water	IPA/NPA/Water
Capacity [m^3/s]. :	2.02E-02	4.61E-06	1.04E-01	6.10E-02
Density [kg/m^3] :	296.0	1156.0	1000.0	753.7
Pressure In [bara] : Out [bara] : difference [bara] :	 1.0 3.0 2.0	 1.0 3.0 2.0	 1.5 3.5 2.0	 0.7 2.7 2.0
Temperature In (= Out) [°C] :	 34	 34	 95	 80
Power [kW] - Theor. : - Actual : - Efficiency :	 4.04 5.39 75%	 9.2E-04 2.3E-03 40%	 20.89 24.87 84%	 12.21 15.26 80%
Number - Theor. : - Actual :	 1	 1	 1	 1
Special Materials of Construction :	CS	CS	CS	CS
Remarks:				

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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PUMPS - SUMMARY

EQUIPMENT NR. NAME	: :	P206 C202 Top transfer pump	P207 C202 Product transfer pump	P208 C203 Bottom pump	P209 V203 Inorganic phase pump
Type Number	: :	Centrifugal 1	Centrifugal 1	Centrifugal 1	Centrifugal 1
Medium transferred	:	IPA/NPA/Water	IPA/Water	IPA	Water/Cyclohexane
Capacity [m^3/s].	:	1.53E-02	9.00E-03	7.61E-03	1.24E-03
Density [kg/m^3]	:	753.7	755.2	761.9	1000.0
Pressure In [bara] Out [bara] difference [bara]	 : : :	 0.7 2.7 2.0	 0.7 2.7 2.0	 1.3 3.3 2.0	 1.0 3.0 2.0
Temperature In (= Out) [°C]	 :	 80	 80	 82.3	 75
Power [kW] - Theor. - Actual - Efficiency	 : : :	 3.05 4.07 75%	 1.80 2.57 70%	 1.52 2.18 70%	 0.25 0.49 50%
Number - Theor. - Actual	 : :	 1	 1	 1	 1
Special Materials of Construction	:	CS	CS	CS	CS

Remarks:

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PUMPS - SUMMARY

EQUIPMENT NR. : NAME :	P210 V203 Organic phase pump	P211 C204 Bottom pump	P212 C204 Reflux pump	P213 C204 Transfer pump
Type : Number :	Centrifugal 1	Centrifugal 1	Centrifugal 1	Centrifugal 1
Medium transferred :	Cyclhexane	Water/Cyclhexane	Cyclohexane/Water	Cyclohexane/Water
Capacity [m^3/s]. :	3.65E-02	1.00E-03	1.10E-03	2.40E-04
Density [kg/m^3] :	747.6	1000.0	1000.0	1000.0
Pressure In [bara] : Out [bara] : difference [bara] :	 1.0 3.0 2.0	 1.2 3.2 2.0	 0.8 2.8 2.0	 0.8 2.8 2.0
Temperature In (= Out) [°C] :	 75	 100	 78.6	 78.6
Power [kW] - Theor. : - Actual : - Efficiency :	 7.30 9.13 80%	 0.20 0.44 45%	 0.22 0.49 45%	 0.05 0.12 40%
Number - Theor. : - Actual :	 1	 1	 1	 1
Special Materials of Construction :	CS	CS	CS	CS
Remarks:				

Designers:	M.E. Brons J.H.M. Jansen	M.P. Hoff E.P.E. Seveke	Project ID-number Date	: **CPD 3242** : 22nd December 99

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PUMPS - SUMMARY

EQUIPMENT NR. NAME	: :	P214 C205 Bottom pump	P215 C205 Reflux pump	P216 C205 Transfer pump
Type Number	: :	Centrifugal 1	Centrifugal 1	Centrifugal 1
Medium transferred	:	IPA/Water/NPA	IPA/Water	IPA/Water
Capacity [m^3/s].	:	7.97E-03	1.46E-02	7.30E-03
Density [kg/m^3]	:	753.2	753.2	753.2
Pressure In [bara] Out [bara] difference [bara]	 : : :	 1.1 3.1 2.0	 0.8 2.8 2.0	 0.8 2.8 2.0
Temperature In (= Out) [°C]	 :	 80	 80	 80
Power [kW] - Theor. - Actual - Efficiency	 : : :	 1.59 2.28 70%	 2.92 3.89 75%	 1.46 2.09 70%
Number - Theor. - Actual	 : :	 1	 1	 1
Special Materials of Construction	 :	CS	CS	CS
Remarks:				

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VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	R101 A/B/C. — in series — in parallel: 3
NAME	:	Propylene reactor

General Data

Function		:	- Reaction
Type		:	- Fixed bed
Column Type		:	- Horizontal
Internals		:	- Fixed-bed
Heating / Cooling medium		:	- Closed
-Type		:	1. Cokes burn-off during catalyst regeneration 2. Heating with fuel gas during cat. reg. (0.21 kg/s)
-Quantity	[kg/s]	:	0.21 kg/s fuel gas (=minimum requirement)
-Press./Temp.'s	[bara/°C]	:	Not applicable
Vessel Diameter	[m]	:	4.62
Vessel Height	[m]	:	6.01
Vessel Total Volume	[m^3]	:	100.93
Vessel Material		:	CS
Pressure Drop	[bara]	:	0.33

Process Conditions

Stream Data[1]		Feed		Effluent		Coke forming	
Temperature	[°C]	650		650		650	
Pressure	[bara]	0.66		0.33		0.5	
Liquid Density	[kg/m^3]	-		-		-	
Liq. Mass Flow	[kg/s]	-		-		-	
Gas Density	[kg/m^3]	0.3		0.21		-	
Gas Mass Flow	[kg/s]	13.74		13.62		0.12 (solid!)	
Components (lumps) (*1*)		Mol%	Wt%	Mol%	Wt%	Mol%	Wt%
HYDROGEN		0.1	0.0	26.3	1.7	0.0	0.0
METHANE		0.0	0.0	1.2	0.6	0.0	0.0
ETHYLENE		0.0	0.0	0.5	0.4	0.0	0.0
ETHANE		0.1	0.0	0.6	0.6	0.0	0.0
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN		51.9	50.1	61.8	81.9	0.0	0.0
PROPANE		45.6	46.0	7.9	11.0	0.0	0.0
ACETONE		0.0	0.0	0.0	0.0	0.0	0.0
IPA		1.6	2.2	1.2	2.2	0.0	0.0
NPA		0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE		0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0
DIPE		0.7	1.6	0.5	1.6	0.0	0.0
WATER		0.0	0.0	0.0	0.0	0.0	0.0
CARBON		0.0	0.0	0.0	0.0	100.0	100.0

Vessel Internals

Catalyst				Beds			
Catalyst type		:	Chromium Oxide	Nr. of Beds		:	1
Catalyst shape		:	Spherical	V_{Bed}	[m^3]	:	70.65
Part. diameter	[mm]	:	5.3				
Catalyst density	[kg/m^3]	:	750				
Catalyst load	[kg]	:	31793				

Remarks:

1. Three reactors required due to regeneration every 7 minutes.
2. The vacuum has been set equivalent to 3.6 bar for cost estimate reasons.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: :	**CPD3242** December 22nd, '99

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Process Systems Engineering

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : R102			In Series : 1x	
NAME : IPA reactor			In Parallel : –	
General Data				
Function	: - Reaction			
Type	: - Fixed bed			
Column Type	: - Horizontal			
Internals	: - Fixed-bed			
Heating / Cooling medium	: - None			
-Type	:			
-Quantity [kg/s]	:			
-Press./Temp.'s [bara/°C]	:			
Vessel Diameter [m]	: 3.57			
Vessel Height [m]	: 14.27			
Vessel Total Volume [m^3]	: 142.63			
Vessel Material	: CS			
Pressure Drop [~~Bara~~]	: 1.14			
Process Conditions				
Stream Data[1]	Feed		Effluent	
Temperature [°C]	158		175	
Pressure [bara]	34		32.86	
Liquid Density [kg/m^3]	-		-	
Liq. Mass Flow [kg/s]	-		-	
Gas Density [kg/m^3]	45.37		37.16	
Gas Mass Flow [kg/s]	106.66		106.66	
Components (lumps) (*1*)	Mol%	Wt%	Mol%	Wt%
HYDROGEN	0.1	0.0	0.1	0.0
METHANE	0.0	0.0	0.0	0.0
ETHYLENE	0.0	0.0	0.0	0.0
ETHANE	0.1	0.1	0.1	0.1
ETHANOL	0.0	0.0	0.0	0.0
PROPYLEN	59.7	68.4	58.3	64.5
PROPANE	11.5	13.8	11.9	13.8
ACETONE	0.0	0.1	0.1	0.1
IPA	1.7	2.8	5.2	8.3
NPA	0.0	0.0	0.0	0.0
CYCLOHEXANE	0.0	0.0	0.0	0.0
N-HEXANE	0.0	0.0	0.0	0.0
DIPE	0.7	2.0	0.8	2.1
WATER	26.1	12.8	23.5	11.2
CARBON	0.0	0.0	0.0	0.0

Vessel Internals

Catalyst		Beds	
Catalyst type	: H_3PO_4 catalyst	V_{Bed} [m^3]	: 99.84
Catalyst shape	: Spherical		
Particle size [mm]	: 5.3		
Catalyst density [kg/m^3]	: 743		
Catalyst load [kg]	: 74183		

Remarks: Heat is released in the reactor by the reaction.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

DISTILLATION COLUMN - SPECIFICATION SHEET

EQUIPMENT NUMBER : C101
NAME : Deethanizer

General Data

Service		: - distillation			
Column Type		: - tray			
Tray Type[1]		: - sieve			
Tray Number Theoretical		: 36			
Actual		: 51			
Feed (Actual)		: 16			
Tray Distance (HETP)	[m]	: 0.600	Tray Material	:	CS[2]
Column Diameter	[m]	: 2.6	Column Material	:	CS[2]
Column Height	[m]	: 35.5			
Heating		: - reboiler			

Process Conditions

Stream Details		Feed	Top	Bottom	Reflux
Temp.	[°C]	27	-33	77	-33
Pressure	[bara]	35	34.8	35.4	34.8
Density	[kg/m³]	89.8	36.2	391	391
Mass Flow	[kg/s]	13.6	0.5	13.1	10.8

Composition	Feed Mol%	Feed Wt%	Top Mol%	Top Wt%	Bottom Mol%	Bottom Wt%	Reflux Mol%	Reflux Wt%
HYDROGEN	26.3	1.7	90.6	42.7	0.0	0.0	0.0	0.0
METHANE	1.2	0.6	4.1	15.4	0.0	0.0	2.6	1.1
ETHYLENE	0.5	0.4	1.7	11.2	0.0	0.0	3.2	2.2
ETHANE	0.6	0.6	1.9	13.6	0.1	0.0	5.6	4.2
ETHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	61.8	81.9	1.6	15.8	86.4	84.6	81.1	84.4
PROPANE	7.9	11.0	0.1	1.3	11.1	11.4	7.4	8.1
ACETONE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
IPA	1.2	2.2	0.0	0.0	1.7	2.3	0.0	0.0
NPA	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	0.5	1.6	0.0	0.0	0.7	1.7	0.0	0.0
WATER	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Column Internals

Trays			Packing		*Not Applicable*
Number of sieve holes		: 11649	Type		:
Active tray area	[m²]	: 3.75	Material		:
Weir Length	[m]	: 2.12	Volume	[m³]	:
Diameter of chute pipe / hole /	[mm]	: 5	Length	[m]	:
			Width	[m]	:
			Height	[m]	:

Remarks:

1. Tray numbering from top to bottom
2. CS = Stainless Steel
3. To prohibit weeping, the total hole area was minimised, this resulted in a strain on the down comer design. This can be avoided by using valves instead of ordinary sieve plates.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

ABSORBER- SPECIFICATION SHEET

EQUIPMENT NUMBER : C102								
NAME : Washing column								
General Data								
Service : - absorption								
Column Type : - packed								
Tray Type : - n.a.								
Tray Number Theoretical : 15								
Tray Distance (HETP) [m] : 0.50				Packing Material : CS				
Column Diameter [m] : 2.728				Column Material : CS				
Column Height [m] : 8.63								
Heating : - none								
Process Conditions								
Stream Details	Feed		Top		Bottom		Absorbent	
Temp. [°C]	30[3]		35		31		30	
Pressure [bara]	29.1[3]		29		29.1		29.1	
Density [kg/m^3]	1.8		1.7		983.9		947.4	
Mass Flow [kg/s]	99.7		91.1		34.8		26.2	
Composition	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%
HYDROGEN	0.2	0.0	0.2	0.0	0.0	0.0	0.0	0.0
METHANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	0.1	0.1	0.1	0.1	0.0	0.0	0.0	0.0
ETHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	67.1	69.0	75.3	75.5	0.1	0.2	0.0	0.0
PROPANE	13.7	14.8	15.4	16.2	0.0	0.0	0.0	0.0
ACETONE	0.1	0.1	0.0	0.1	0.0	0.1	0.0	0.0
IPA	5.7	8.4	2.3	3.3	5.2	15.4	0.0	0.0
NPA	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	0.9	2.2	1.0	2.4	0.0	0.1	0.0	0.0
WATER	12.3	5.4	5.7	2.4	94.7	84.2	100.0	100.0

Column Internals

Trays	*Not Applicable*	Packing	
Number of caps / sieve holes /	:	Type	: Structured Mellapak
Active tray area [m^2]	:	Material	: Steel
Weir Length [m]	:	Volume bed [m^3]	: 45.661
Diameter of		Length [m]	: n.a.
		Width [m]	: n.a.
chute pipe / hole / [mm]	:	Height [m]	: n.a.

Remarks:
1. The height of this column has been based on a typical HTU mentioned in Coulson & Richardson's of 0.5m.
2. The Mellapak has been chosen because of its low packing factor (8 m^{-1}).
3. Based on recommended flowscheme.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

DISTILLATION - SPECIFICATION SHEET

EQUIPMENT NUMBER	: C201		
NAME	: Light ends column		
General Data			
Service	: - distillation		
Column Type	: - tray		
Tray Type	: - sieve		
Tray Number Theoretical	: 36		
Actual	: 45		
Feed (Actual)	: 28 IPA rich phase		
	: 1 Water rich phase		
Tray Distance (HETP) [m]	: 0.60	Tray Material	: CS
Column Diameter [m]	: 3.03	Column Material	: CS
Column Height [m]	: 31.1		
Heating	: - reboiler		

Process Conditions

Stream Details	IPA rich feed		Water rich feed		Top		Bottom	
Temp. [°C]	34		34		51		86	
Pressure [bara]	1.00		1.00		0.34		1.39	
Density [kg/m^3]	917.7		984.8		1.711		898.1	
Mass Flow [kg/s]	53.2		51.3		0.1		104.4	
Composition	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%
HYDROGEN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	0.0	0.0	0.0	0.0	0.1	0.1	0.0	0.0
ETHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	0.1	0.1	0.0	0.0	57.6	53.4	0.0	0.0
PROPANE	0.0	0.0	0.0	0.0	13.2	12.8	0.0	0.0
ACETONE	0.0	0.0	0.0	0.0	8.6	11.0	0.0	0.0
IPA	11.1	29.4	0.0	0.0	0.0	0.0	5.0	15.0
NPA	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	0.0	0.0	0.0	0.0	7.9	17.8	0.0	0.0
WATER	88.8	70.3	100.0	100.0	12.6	5.0	95.0	85.0

Column Internals

Trays			Packing		*Not Applicable*
Number of sieve holes		: 33105	Type		:
			Material		:
Active tray area	[m^2]	: 6.01	Volume	[m^3]	:
Weir Length	[m]	: 2.12	Length	[m]	:
Diameter of			Width	[m]	:
chute pipe / hole /	[mm]	: 5	Height	[m]	:

Remarks:

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

Delft, University of Technology
Process Systems Engineering

DISTILLATION - SPECIFICATION SHEET

EQUIPMENT NUMBER	: C202			
NAME	: IPA CBM column			

General Data

Service		:	- distillation			
Column Type		:	- tray			
Tray Type		:	- sieve			
Tray Number	Theoretical	:	46			
	Actual	:	58			
	Feed (Actual)	:	30			
	Side stream	:	2			
Tray Distance (HETP)	[m]	:	0.60	Tray Material	:	CS
Column Diameter	[m]	:	2.84	Column Material	:	CS
Column Height	[m]	:	39.68			
Heating		:	- reboiler			

Process Conditions

Stream Details		Feed		Top		Bottom		Side stream		Reflux	
Temp.	[°C]	86		80		95		80		80	
Pressure	[bara]	1		0.66		1.50		0.67		0.66	
Density	[kg/m^3]	898.0		753.7		884.5		755.2		753.7	
Mass Flow	[kg/s]	104.5		11.5		86.2		6.8		46.0	
Composition		Mol %	Wt %	Mol %	Wt %	Mol %	Wt%	Mol %	Wt%	Mol %	Wt%
HYDROGEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
METHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
ETHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
PROPYLEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
PROPANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
ACETONE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
IPA		5.0	15.0	65.0	86.1	0.0	0.0	63.5	85.3	64.93	86.04
NPA		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.03	0.04
CYCLOHEXANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
DIPE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.00	0.00
WATER		95.0	85.0	35.0	13.9	100.0	100.0	36.5	14.7	35.04	13.92

Column Internals

<u>Trays</u>				<u>Packing</u>			*Not Applicable*
Number of				Type		:	
sieve holes		:	25465	Material		:	
Active tray area	[m^2]	:	4.44	Volume	[m^3]	:	
Weir Length	[m]	:	2.30	Length	[m]	:	
Diameter of				Width	[m]	:	
chute pipe / hole /	[mm]	:	5	Height	[m]	:	

Remarks:

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

DISTILLATION - SPECIFICATION SHEET

EQUIPMENT NUMBER	: C203		
NAME	: IPA drying column		

General Data

Service		:	- distillation			
Column Type		:	- tray			
Tray Type		:	- sieve			
Tray Number	Theoretical	:	40			
	Actual	:	50			
	Feed (Actual)	:	18 IPA rich phase			
		:	1 Cyclohexane rich phase			
Tray Distance (HETP)	[m]	:	0.45	Tray Material	:	CS
Column Diameter	[m]	:	2.60	Column Material	:	CS
Column Height	[m]	:	23.29			
Heating		:	- reboiler			

Process Conditions

Stream Details		IPA rich feed		Cyclohexane rich feed		Top		Bottom	
Temp.	[°C]	79		74.7		74.7		82.3	
Pressure	[bara]	1		0.80		0.80		1.25	
Density	[kg/m^3]	761.9		747.6		11.9		761.9	
Mass Flow	[kg/s]	7.0		27.3		28.5		5.8	
Composition		Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%
HYDROGEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
IPA		63.9	84.9	2.3	1.7	2.8	2.2	99.6	99.8
NPA		0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.1
CYCLOHEXANE		0.5	0.9	97.7	98.3	83.0	94.3	0.0	0.0
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
WATER		35.6	14.2	0.0	0.0	14.3	3.5	0.3	0.1

Column Internals

Trays				Packing		Not Applicable
Number of sieve holes		:	18926	Type		:
Active tray area	[m^2]	:	0.37	Material		:
Weir Length	[m]	:	2.11	Volume	[m^3]	:
Diameter of chute pipe / hole /	[mm]	:	5	Length	[m]	:
				Width	[m]	:
				Height	[m]	:

Remarks:

1. Due to the limited amount of information, the sizing of the column has been based on the top section.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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DISTILLATION - SPECIFICATION SHEET

EQUIPMENT NUMBER : C204
NAME : Cyclohexane recovery column

General Data

Service		: - distillation			
Column Type		: - tray			
Tray Type		: - sieve			
Tray Number	Theoretical	: 28			
	Actual	: 35			
	Feed (Actual)	: 18[2]			
Tray Distance (HETP)	[m]	: 0.45	Tray Material	:	CS
Column Diameter	[m]	: 1.34[1]	Column Material	:	CS
Column Height	[m]	: 19.00			
Heating		: - reboiler			

Process Conditions

Stream Details		Feed		Top		Bottom		Reflux	
Temp.	[°C]	52.4		78.6		100		78.6	
Pressure	[bara]	1		0.82[3]		1.17[3]		0.82[3]	
Density	[kg/m^3]	1000		1000		1000		1000	
Mass Flow	[kg/s]	1.24		0.24		1.00		1.10[4]	
Composition		Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%
HYDROGEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
IPA		5.3	15.0	82.3	76.8	0.0	0.0	82.3	76.8
NPA		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE		1.3	5.0	17.7	23.2	0.1	0.6	17.7	23.2
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
WATER		93.5	80.0	0.0	0.0	99.9	99.4	0.0	0.0

Column Internals

Trays				Packing		Not Applicable
Number of				Type		:
caps / sieve holes /		:		Material		:
Active tray area	[m^2]	:	([5])	Volume	[m^3]	:
Weir Length	[m]	:	([5])	Length	[m]	:
Diameter of				Width	[m]	:
chute pipe / hole /	[mm]	:	([5])	Height	[m]	:

Remarks due to insufficient information:
1. It has been assumed that ratio surfaces is equal to the ratio of the feed
2. Feed stage is assumed to be in the middle (stage 18).
3. Pressure drop per stage has been set to 1000 kPa.
4. Based on pump capacity.
5. Not enough data available.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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DISTILLATION - SPECIFICATION SHEET

EQUIPMENT NUMBER : C205							
NAME : NPA recovery column							
General Data							
Service : - distillation							
Column Type : - tray							
Tray Type : - sieve							
Tray Number Theoretical : 22							
Actual : 28							
Feed (Actual) : 19							
Tray Distance (HETP) [m] : 0.60				Tray Material : CS			
Column Diameter [m] : 2.682				Column Material : CS			
Column Height [m] : 18.98							
Heating : - reboiler							
Process Conditions							

Stream Details	Feed		Top		Bottom		Reflux	
Temp. [°C]	80		80		80		80	
Pressure [bara]	1		0.77		1.11		0.77	
Density [kg/m³]	753.7		753.2		754.9		753.2	
Mass Flow [kg/s]	11.5		5.5		6.0		11	
Composition	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%
HYDROGEN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
IPA	65.0	86.1	66.4	86.8	63.7	85.3	66.4	86.8
NPA	0.0	0.0	0.0	0.0	0.1	0.1	0.0	0.0
CYCLOHEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
WATER	35.0	13.9	33.6	13.2	36.2	14.6	33.6	13.2

Column Internals

Trays				Packing		Not Applicable
Number of				Type		:
sieve holes		:	35651	Material		:
Active tray area	[m²]	:	3.95	Volume	[m³]	:
Weir Length	[m]	:	2.17	Length	[m]	:
Diameter of				Width	[m]	:
chute pipe / hole /	[mm]	:	5	Height	[m]	:

Remarks:

1. During the sizing of the column it was found that figures from ASPEN couldn't be correct. Therefore it has been assumed that the reflux ratio is about two. This assumption does make the number of plates doubtful.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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DECANTER - SPECIFICATION SHEET

EQUIPMENT NUMBER : V102
NAME : HP separator

General Data

Function		: - Separation		
Column Type		: - Horizontal		
Internals		: - none		
Heating / Cooling medium		: - none		
-Type		: n.a.		
-Quantity	[kg/s]	: n.a.		
-Press./Temp.'s	[bara/°C]	: n.a.		
Vessel Diameter	[m]	: 2.5	Vessel Material	: CS
Vessel Height	[m]	: 6.5		
Vessel Tot. Volume	[m³]	: 31.9		

Process Conditions

Stream Data		Feed		Top		Bottom	
Temp.	[°C]	90.0		90.0[2]		90.0[2]	
Pressure	[bara]	32		30		30	
Density	[kg/m³]	68.0		1.54		953.4	
Mass Flow	[kg/s]	106.6		99.7		6.9	
Components		Mol%	Wt%	Mol %	Wt%	Mol%	Wt%
HYDROGEN		0.1	0.0	0.2	0.0	0.0	0.0
METHANE		0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0
ETHANE		0.1	0.1	0.1	0.1	0.0	0.0
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN		58.3	64.5	67.1	69.0	0.0	0.1
PROPANE		11.9	13.8	13.7	14.8	0.0	0.0
ACETONE		0.1	0.1	0.1	0.1	0.0	0.0
IPA		5.2	8.3	5.7	8.4	1.9	5.9
NPA		0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE		0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0
DIPE		0.8	2.1	0.9	2.2	0.0	0.0
WATER		23.5	11.2	12.3	5.4	98.1	93.9

Vessel Internals

<u>Coils</u>		*Not Applicable*	<u>Feed Insert</u>		
Number of tubes		:	-Type		: Tube
Tube shape		:	-Position		: Top
Tube diam	[m²]	:	-Diameter	[m]	: Determined by expert
Tube length	[m]	:	-Length	[m]	: -
Tube surf.	[m²]	:	-Material		: CS

Remarks:
1. Complete phase separation is not necessary in this vessel; nonetheless, the vapour velocity is probably still too high. This is because the sizing calculation has been carried out mainly with respect to the liquid phase. A much larger vessel would result if one would aim for a low vapour velocity; perhaps it would then be preferable to insert demisting pads or mesh, although this would increase the pressure drop.
2. Based on recommended flow scheme.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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DECANTER - SPECIFICATION SHEET

EQUIPMENT NUMBER : V201
NAME : Light ends column phase separator

General Data

Function		:	- Separation
Column Type		:	- Horizontal
Internals		:	- none
Heating / Cooling medium		:	- none
-Type		:	n.a.
-Quantity	[kg/s]	:	n.a.
-Press./Temp.'s	[bara/°C]	:	n.a.
Vessel Diameter	[m]	:	2.5 Vessel Material : CS
Vessel Height	[m]	:	9.5
Vessel Tot. Volume	[m^3]	:	46.6

Process Conditions

Stream Data		Feed		Top		Middle		Bottom	
Temp.	[°C]	34		34		34		34	
Pressure	[bara]	1		1		1		1	
Density	[kg/m^3]	0.84		1.71		1156		269	
Mass Flow	[kg/s]	6.12		0.12		0.01		5.99	
Components		Mol%	Wt%	Mol %	Wt%	Mol %	Wt%	Mol%	Wt%
HYDROGEN		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANE		0.0	0.0	0.1	0.1	0.0	0.0	0.0	0.0
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN		1.2	1.1	62.5	52.8	0.0	0.0	0.0	0.0
PROPANE		0.3	0.3	14.3	12.7	0.0	0.0	0.0	0.0
ACETONE		0.3	0.3	14.5	16.9	0.0	0.0	0.0	0.0
IPA		62.4	83.5	0.0	0.0	0.0	0.0	63.6	85.3
NPA		0.1	0.1	0.0	0.0	100	100	0.0	0.0
CYCLOHEXANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DIPE		0.2	0.4	8.6	17.6	0.0	0.0	0.0	0.0
WATER		35.7	14.4	0.0	0.0	0.0	0.0	36.4	14.7

Vessel Internals

Coils	*Not Applicable*		Feed Insert			
Number of tubes		:	-Type		:	Tube
Tube shape		:	-Position		:	Top
Tube diam	[m^2]	:	-Diameter	[m]	:	Determined by expert
Tube length	[m]	:	-Length	[m]	:	-
Tube surf.	[m^2]	:	-Material		:	CS

Remarks:

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

DECANTER - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	V203
NAME	:	Drying column phase separator

General Data

Function		:	- Separation
Column Type		:	- Horizontal
Internals		:	- none
Heating / Cooling medium		:	- none
-Type		:	n.a.
-Quantity	[kg/s]	:	n.a.
-Press./Temp.'s	[bara/°C]	:	n.a.
Vessel Diameter	[m]	:	2.75 Vessel Material : CS
Vessel Height	[m]	:	11
Vessel Tot. Volume	[m^3]	:	65.3

Process Conditions

Stream Data		Feed		Top		Bottom	
Temp.	[°C]	75		75		75	
Pressure	[bara]	1		1		1	
Density	[kg/m^3]	787		752		973	
Mass Flow	[kg/s]	28.5		27.3		1.2	
Components		Mol%	Wt%	Mol %	Wt%	Mol%	Wt%
HYDROGEN		0.0	0.0	0.0	0.0	0.0	0.0
METHANE		0.0	0.0	0.0	0.0	0.0	0.0
ETHYLENE		0.0	0.0	0.0	0.0	0.0	0.0
ETHANE		0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL		0.0	0.0	0.0	0.0	0.0	0.0
PROPYLEN		0.0	0.0	0.0	0.0	0.0	0.0
PROPANE		0.0	0.0	0.0	0.0	0.0	0.0
ACETONE		0.0	0.0	0.0	0.0	0.0	0.0
IPA		2.8	2.2	2.3	1.7	5.3	15.0
NPA		0.0	0.0	0.0	0.0	0.0	0.0
CYCLOHEXANE		83.0	94.3	97.7	98.3	1.3	5.0
N-HEXANE		0.0	0.0	0.0	0.0	0.0	0.0
DIPE		0.0	0.0	0.0	0.0	0.0	0.0
WATER		14.3	3.5	0.0	0.0	93.5	80.0

Vessel Internals

Coils		*Not Applicable*	Feed Insert			
Number of tubes		:	-Type		:	Tube
Tube shape		:	-Position		:	Side
Tube diam	[m^2]	:	-Diameter	[m]	:	Determined by expert
Tube length	[m]	:	-Length	[m]	:	-
Tube surf.	[m^2]	:	-Material		:	CS

Remarks:

3. In the stream tables, there is still some vapor in the feed stream. This is not meant to be, and therefore the density of an all liquid feed stream is used to design the vessel.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER : T201 A/B/C/D		
NAME : Run down tank		
General Data		
Function : - Storage		
Column Type : - Vertical		
Internals : - None		
Heating / Cooling medium : - None		
-Type :		
-Quantity [kg/s] :		
-Press./Temp.'s [bara/°C] :		
Vessel Diameter [m] : 7.80		
Vessel Height [m] : 4.68		
Vessel Total Volume [m^3] : 224		
Vessel Material : CS		
Pressure Drop [Bara] :		
Process Conditions		
Stream Data[1]	Feed	Effluent
Temperature [°C]	30	30
Pressure [bara]	1	1
Liquid Density [kg/m^3]	761.9	761.9
Liq. Mass Flow [kg/s]	-	-
Gas Density [kg/m^3]	n.a.	n.a.
Gas Mass Flow [kg/s]	0	0
Remarks:		

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number : **CPD3242** Date : December 22nd, '99

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Process Systems Engineering

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	T202
NAME	:	Off spec tank
General Data		
Function	:	- Storage
Column Type	:	- Vertical 1
Internals	:	- None
Heating / Cooling medium	:	- None
-Type	:	
-Quantity [kg/s]	:	
-Press./Temp.'s [bara/°C]	:	
Vessel Diameter [m]	:	20.74
Vessel Height [m]	:	14.51
Vessel Total Volume [m^3]	:	4901
Vessel Material	:	CS
Pressure Drop [Bara]	:	

Process Conditions

Stream Data[1]		Feed	Effluent
Temperature	[°C]	30	30
Pressure	[bara]	1	1
Liquid Density	[kg/m^3]	761.9	761.9
Liq. Mass Flow	[kg/s]	-	-
Gas Density	[kg/m^3]	n.a.	n.a.
Gas Mass Flow	[kg/s]	0	0

Remarks:

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: :	**CPD3242** December 22nd, '99

Delft, University of Technology
Process Systems Engineering

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	T203
NAME	:	On spec tank

General Data

Function		:	- Storage
Column Type		:	- Vertical
Internals		:	- None
Heating / Cooling medium		:	- None
-Type		:	
-Quantity	[kg/s]	:	
-Press./Temp.'s	[bara/°C]	:	
Vessel Diameter	[m]	:	52.96
Vessel Height	[m]	:	27.27
Vessel Total Volume	[m^3]	:	60080
Vessel Material		:	CS
Pressure Drop	[Bara]	:	

Process Conditions

Stream Data[1]		Feed	Effluent
Temperature	[°C]	30	30
Pressure	[bara]	1	1
Liquid Density	[kg/m^3]	761.9	761.9
Liq. Mass Flow	[kg/s]	-	-
Gas Density	[kg/m^3]	n.a.	n.a.
Gas Mass Flow	[kg/s]	0	0

Remarks:

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

Delft, University of Technology
Process Systems Engineering

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	T204
NAME	:	Sales tank

General Data

Function		:	- Storage
Column Type		:	- Vertical
Internals		:	- None
Heating / Cooling medium		:	- None
-Type		:	
-Quantity	[kg/s]	:	
-Press./Temp.'s	[bara/°C]	:	
Vessel Diameter	[m]	:	14.06
Vessel Height	[m]	:	21.15
Vessel Total Volume	[m^3]	:	28008
Vessel Material		:	CS
Pressure Drop	[Bara]	:	

Process Conditions

Stream Data[1]		Feed	Effluent
Temperature	[°C]	30	30
Pressure	[bara]	1	1
Liquid Density	[kg/m^3]	761.9	761.9
Liq. Mass Flow	[kg/s]	-	-
Gas Density	[kg/m^3]	n.a.	n.a.
Gas Mass Flow	[kg/s]	0	0

Remarks:

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

Delft, University of Technology
Process Systems Engineering

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER	: D101 A/B
NAME	: Dryer in overall recycle stream

General Data

Function		: - Dryer
Type		: - Fixed bed
Column Type		: - Horizontal
Internals		: - Fixed-bed
Heating / Cooling medium		: - None
-Type		:
-Quantity	[kg/s]	:
-Press./Temp.'s	[bara/°C]	:
Vessel Diameter	[m]	: 2.9
Vessel Height	[m]	: 11.60
Vessel Total Volume	[m^3]	: 76.68
Vessel Material		: CS
Pressure Drop	[Bara]	: 0.26

Process Conditions

Stream Data[1]		Feed		Effluent	
Temperature	[°C]	35		35	
Pressure	[bara]	15[3]		1	
Liquid Density	[kg/m^3]	-		-	
Liq. Mass Flow	[kg/s]	-		-	
Gas Density	[kg/m^3]	1.65		1.71	
Gas Mass Flow	[kg/s]	9.11		8.88	
Components (lumps) *(1)*		Mol%	Wt%	Mol%	Wt%
HYDROGEN		0.2	0.0	0.2	0.0
METHANE		0.0	0.0	0.0	0.0
ETHYLENE		0.0	0.0	0.0	0.0
ETHANE		0.1	0.1	0.1	0.1
ETHANOL		0.0	0.0	0.0	0.0
PROPYLEN		75.3	75.5	79.8	77.4
PROPANE		15.4	16.2	16.3	16.6
ACETONE		0.0	0.1	0.1	0.1
IPA		2.3	3.3	2.5	3.4
NPA		0.0	0.0	0.0	0.0
CYCLOHEXANE		0.0	0.0	0.0	0.0
N-HEXANE		0.0	0.0	0.0	0.0
DIPE		1.0	2.4	1.1	2.5
WATER		5.7	2.4	0.0	0.0
CARBON		0.0	0.0	0.0	0.0

Vessel Internals

Catalyst			Beds		
Catalyst type		: Drying agent	V_{Bed}	[m^3]	: 66.67
Catalyst shape		: Spherical			
Particle size	[mm]	: 5.3			
Catalyst density	[kg/m^3]	: 750			
Catalyst load	[kg]	: 50000			

Remarks:

1. Dryer has to be regenerated 3 times per day.
2. Regeneration duty is 679.01 kJ/s, total duty is 14667 MJ and regeneration time is 6 hr.
3. Based on recommended flow scheme.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E101				
NAME : R101 Feed-Effluent A				
General Data				
Service		: - Heat exchanger		
Type		: - Floating head		
Position		: - Horizontal		
Capacity	[kW]	:	1625	(Calc.)
Heat Exchanging Area	[m^2]	:	92	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	200	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	93	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.95	
Corrected LMTD	[°C]	:	88	

Process Conditions

			Shell Side	Tube Side
Medium		:	Propane/Propylene	Propylene
Mass Stream	[kg/s]	:	13.7	13.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	2.26	2.51
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	4.7	147.6
Temperature OUT	[°C]	:	57.0	100.0
Pressure	[bara]	:	2.0	0.3
Material		:	C.S.	C.S.

Remarks:
1.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E102				
NAME : R101 Feed-Effluent B				
General Data				
Service	: - Heat exchanger			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	2816	(Calc.)
Heat Exchanging Area	[m^2]	:	204	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot {}^oC$]	:	200	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	86	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.80	
Corrected LMTD	[°C]	:	69	

Process Conditions

			Shell Side	Tube Side
Medium		:	Propane/Propyle ne	Propylene
Mass Stream	[kg/s]	:	13.7	13.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	2.26	2.51
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	57.0	230.0
Temperature OUT	[°C]	:	147.6	147.6
Pressure	[bara]	:	2.0	0.3
Material		:	C.S.	C.S.

Remarks:
2.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E103				
NAME : R101 Feed-Effluent C				
General Data				
Service	: - Heat exchanger			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	2562	(Calc.)
Heat Exchanging Area	[m^2]	:	217	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	200	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	79	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.75	
Corrected LMTD	[°C]	:	59	

Process Conditions				
			Shell Side	Tube Side
Medium		:	Propane/Propylene	Propylene
Mass Stream	[kg/s]	:	13.7	13.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	2.26	2.51
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	147.6	305.0
Temperature OUT	[°C]	:	230.0	230.0
Pressure	[bara]	:	2.0	0.3
Material		:	C.S.	C.S.

Remarks:
3.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER	: E104			
NAME	: R101 Feed-Effluent D			
General Data				
Service	: - Heat exchanger			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	2176	(Calc.)
Heat Exchanging Area	[m^2]	:	189	(Calc.)
Overall Heat Transfer Coefficient	[W/m^2·°C]	:	200	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	72	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.80	
Corrected LMTD	[°C]	:	57	

Process Conditions

			Shell Side	Tube Side
Medium		:	Propane/Propylene	Propylene
Mass Stream	[kg/s]	:	13.7	13.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	2.26	2.51
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	230.0	368.7
Temperature OUT	[°C]	:	300.0	305.0
Pressure	[bara]	:	2.0	0.3
Material		:	C.S.	C.S.

Remarks:
4.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: :	**CPD3242** December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E105				
NAME : R101 Effluent - Steam				
General Data				
Service	: - Heat exchanger			
Type	: - Fixed Tube Sheets			
Position	: - Horizontal			
Capacity	[kW]	:	12498	(Calc.)
Heat Exchanging Area	[m^2]	:	62	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	700	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	308	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.94	
Corrected LMTD	[°C]	:	289	

Process Conditions

			Shell Side	Tube Side
Medium		:	Propylene	Steam
Mass Stream	[kg/s]	:	13.6	4.8
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot °C$]	:	3.26	
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	650.0	100.0
Temperature OUT	[°C]	:	368.7	300.0
Pressure	[bara]	:	0.3	15.0
Material		:	C.S.	C.S.

Remarks:
5.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E106					
NAME : K101 Effluent					
General Data					
Service	: - Cooler				
Type	: - Finned tubes				
Position	: - Horizontal				
Capacity	[kW]	:	2986		(Calc.)
Heat Exchanging Area	[m^2]	:	50		(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2\cdot°C$]	:	750		(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	80		
Passes Tube Side		:	1		
Passes Shell Side		:	n.a.		
Correction Factor LMTD (min 0.75)		:	1.00		
Corrected LMTD	[°C]	:	80		
Process Conditions					
			Shell Side	Tube Side	
Medium		:	Air	Propylene	
Mass Stream	[kg/s]	:	97.6	13.6	
Mass Stream to Evaporize	[kg/s]	:			
Mass Stream to Condense	[kg/s]	:			
Average Specific Heat	[kJ/kg·°C]	:	1.02	2.19	
Heat of Evap./Condensation	[kJ/kg]	:			
Temperature IN	[°C]	:	30.0	180.0	
Temperature OUT	[°C]	:	60.0	80.0	
Pressure	[bara]	:	1.0	0.7	
Material		:	C.S.	C.S.	
Remarks:					
6. Fan duty: 7.99 kW					

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22[nd], '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E107				
NAME : C203 Reboiler				
General Data				
Service	: - Reboiler			
Type	: - Thermosyphon			
Position	: - Vertical			
Capacity	[kW]	:	825	(Calc.)
Heat Exchanging Area	[m^2]	:	49	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	200	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	85	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	85	

Process Conditions

			Shell Side	Tube Side
Medium		:	IPA	Propylene
Mass Stream	[kg/s]	:		13.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		2.40
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	82.3	180.3
Temperature OUT	[°C]	:	82.3	155.0
Pressure	[bara]	:	1.0	36.0
Material		:	C.S.	C.S.

Remarks:
7.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER	: E108			
NAME	: C101 Feed			
General Data				
Service	: - Cooler			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	6830	(Calc.)
Heat Exchanging Area	[m^2]	:	522	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	500	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	28	
Passes Tube Side		:	4	
Passes Shell Side		:	2	
Correction Factor LMTD (min 0.75)		:	0.95	
Corrected LMTD	[°C]	:	26	

Process Conditions

			Shell Side	Tube Side
Medium		:	Cooling water	Propylene
Mass Stream	[kg/s]	:	108.9	13.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot °C$]	:	4.18	3.90
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	25.0	155.0
Temperature OUT	[°C]	:	40.0	26.9
Pressure	[bara]	:	1.0	35.5
Material		:	C.S.	C.S.

Remarks:
8.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E109				
NAME : C101 Condenser				
General Data				
Service	: - Condenser			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	5394	(Calc.)
Heat Exchanging Area	[m^2]	:	1065	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	500	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	10	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.99	
Corrected LMTD	[°C]	:	10	

Process Conditions

			Shell Side	Tube Side
Medium		:	Propylene	H2/Light ends
Mass Stream	[kg/s]	:	23.9	11.3
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	29.70	
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	-47.6	-33.0
Temperature OUT	[°C]	:	-40.0	-33.0
Pressure	[bara]	:	1.0	34.8
Material		:	C.S.	C.S.

Remarks:
9.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

Delft. University of Technology
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E110				
NAME : C101 Condenser Heat pump				
General Data				
Service	: - Cooler			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	11599	(Calc.)
Heat Exchanging Area	[m^2]	:	394	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	46	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.85	
Corrected LMTD	[°C]	:	39	

Process Conditions

			Shell Side	Tube Side
Medium		:	Cooling water	Propylene
Mass Stream	[kg/s]	:	184.9	23.9[1]
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	4.18	4.60
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	25.0	144.8
Temperature OUT	[°C]	:	40.0	40.0
Pressure	[bara]	:	1.0	36.0
Material		:	C.S.	C.S.

Remarks:
10. Closed heat pump cycle not included in stream summary.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E111				
NAME : C101 Reboiler				
General Data				
Service	:	- Reboiler		
Type	:	- Thermosyphon		
Position	:	- Vertical		
Capacity	[kW]	:	6240	(Calc.)
Heat Exchanging Area	[m^2]	:	139	(Calc.)
Overall Heat Transfer Coefficient	[W/m^2·°C]	:	800	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	59	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.96	
Corrected LMTD	[°C]	:	56	

Process Conditions

			Shell Side	Tube Side
Medium		:	LP Steam[1]	Propylene
Mass Stream	[kg/s]	:	2.6	
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	196.8	77.1
Temperature OUT	[°C]	:	100.0	77.1
Pressure	[bara]	:	4.7	35.4
Material		:	C.S.	C.S.

Remarks:

11. From compressor K104; self produced.

Delft. University of Technology
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E112				
NAME : R102 Feed heater				
General Data				
Service	: - Heater[1]			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	14061	(Calc.)
Heat Exchanging Area	[m^2]	:	475	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot {}^{\circ}C$]	:	800	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	39	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.96	
Corrected LMTD	[°C]	:	37	

Process Conditions				
			Shell Side	Tube Side
Medium		:	MP Steam	Propylene/IPA
Mass Stream	[kg/s]	:	6.8	106.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot {}^{\circ}C$]	:		7.20
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	200.0	139.7
Temperature OUT	[°C]	:	175.0	158.0
Pressure	[bara]	:	12.0	34.0
Material		:	C.S.	C.S.

Remarks:
12. Product-feed exchanger was unfeasible, because of small temperature difference.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

Delft. University of Technology
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E113				
NAME : C201 Reboiler				
General Data				
Service	: - Reboiler			
Type	: - Thermosyphon			
Position	: - Vertical			
Capacity	[kW]	:	20201	(Calc.)
Heat Exchanging Area	[m^2]	:	945	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	300	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	71	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	71	

Process Conditions			Shell Side	Tube Side
Medium		:	IPA/water	IPA/Propylene
Mass Stream	[kg/s]	:		106.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		5.60
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	85.6	175.0
Temperature OUT	[°C]	:	85.6	141.3
Pressure	[bara]	:	1.4	32.9
Material		:	C.S.	C.S.

Remarks:
13.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E114				
NAME : R102 Effluent cooler				
General Data				
Service		:	- Cooler	
Type		:	- Finned tubes	
Position		:	- Horizontal	
Capacity	[kW]	:	25824	(Calc.)
Heat Exchanging Area	[m^2]	:	614	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	600	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	70	
Passes Tube Side		:	1	
Passes Shell Side		:	n.a.	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	70	

Process Conditions

			Shell Side	Tube Side
Medium		:	Air	IPA/Propylene
Mass Stream	[kg/s]	:	843.9	106.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	1.02	4.70
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	30.0	141.3
Temperature OUT	[°C]	:	60.0	90.0
Pressure	[bara]	:	1.0	32.9[2]
Material		:	C.S.	C.S.

Remarks:
14. Fan duty: 98.55 kW
15. Based on recommended flow scheme

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E115				
NAME : C102 Feed cooler A				
General Data				
Service	: - Cooler			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	36459	(Calc.)
Heat Exchanging Area	[m^2]	:	1706	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	29	
Passes Tube Side		:	4	
Passes Shell Side		:	2	
Correction Factor LMTD (min 0.75)		:	0.98	
Corrected LMTD	[°C]	:	28	

Process Conditions			Shell Side	Tube Side
Medium		:	Cooling water	IPA/Propylene
Mass Stream	[kg/s]	:	581.5	99.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	4.18	7.3
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	25.0	90.0[1]
Temperature OUT	[°C]	:	40.0	40.0
Pressure	[bara]	:	1.0	30.2[1]
Material		:	C.S.	C.S.

Remarks:
16. Based on recommended flow scheme

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E116				
NAME : C102 Feed cooler B				
General Data				
Service	: - Cooler			
Type	: - Floating head			
Position	: - Horizontal			
Capacity	[kW]	:	2967	(Calc.)
Heat Exchanging Area	[m^2]	:	645	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot {}^oC$]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	7	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	0.85	
Corrected LMTD	[°C]	:	6	

Process Conditions

			Shell Side	Tube Side
Medium		:	Cooling water	IPA/Propylene
Mass Stream	[kg/s]	:	142.0	99.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot {}^oC$]	:	4.18	2.98
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	25.0	40.0
Temperature OUT	[°C]	:	30.0	30.0
Pressure	[bara]	:	1.0	29.7[1]
Material		:	C.S.	C.S.

Remarks:
17. Based on recommended flow scheme

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E201			
NAME : V201 Feed cooler			
General Data			
Service	: - Cooler		
Type	: - Fixed Tube Sheets		
Position	: - Horizontal		
Capacity	[kW]	: 2615	(Calc.)
Heat Exchanging Area	[m^2]	: 185	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	: 750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	: 21	
Passes Tube Side		: 4	
Passes Shell Side		: 2	
Correction Factor LMTD (min 0.75)		: 0.90	
Corrected LMTD	[°C]	: 19	

Process Conditions

			Shell Side	Tube Side
Medium		:	Water/IPA	Cooling water
Mass Stream	[kg/s]	:	6.0	41.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot °C$]	:	9.60	4.18
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	79.8	25.0
Temperature OUT	[°C]	:	34.3	40.0[1]
Pressure	[bara]	:	1.0	1.0
Material		:	C.S.	C.S.

Remarks:
18. Temperature cross: in practice this exchanger would be split in two.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	E202
NAME	:	DELETED
General Data		

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number : **CPD3242** Date : December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E203				
NAME : C202 Reboiler				
General Data				
Service	: - Reboiler			
Type	: - Thermosyphon			
Position	: - Vertical			
Capacity	[kW]	:	58308	(Calc.)
Heat Exchanging Area	[m^2]	:	742	(Calc.)
Overall Heat Transfer Coefficient	[W/m^2·°C]	:	800	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	98	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	98	

Process Conditions

			Shell Side	Tube Side
Medium		:	Water	HP Steam
Mass Stream	[kg/s]	:		27.7
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	95	225.0
Temperature OUT	[°C]	:	95	175.0
Pressure	[bara]	:	1.5	18.0
Material		:	C.S.	C.S.

Remarks:
19.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E204				
NAME : C202 Condenser				
General Data				
Service	: - Condenser			
Type	: - Finned tubes			
Position	: - Horizontal			
Capacity	[kW]	:	50799	(Calc.)
Heat Exchanging Area	[m^2]	:	3103[2]	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	500	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	33	
Passes Tube Side		:	1	
Passes Shell Side		:	n.a.	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	33	

Process Conditions

			Shell Side	Tube Side
Medium		:	Air	IPA/water
Mass Stream	[kg/s]	:	1660.1	
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	1.02	
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	30.0	80.0
Temperature OUT	[°C]	:	60.0	80.0
Pressure	[bara]	:	1.0	0.7
Material		:	C.S.	C.S.

Remarks:
20. Fan duty: 498.15 kW
21. In practice 6 air coolers would be placed

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E205				
NAME : C205 reboiler				
General Data				
Service	: - Reboiler			
Type	: - Thermosyphon			
Position	: - Vertical			
Capacity	[kW]	:	46950	(Calc.)
Heat Exchanging Area	[m^2]	:	496	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot {}^oC$]	:	800	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	118	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	118	

Process Conditions				
			Shell Side	Tube Side
Medium		:	NPA	HP Steam
Mass Stream	[kg/s]	:		22.3
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	80.0	225.0
Temperature OUT	[°C]	:	80.0	175.0
Pressure	[bara]	:	1.1	18.0
Material		:	C.S.	C.S.

Remarks:
22.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22^{nd}, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E206				
NAME : C205 Condenser				
General Data				
Service	: - Condenser			
Type	: - Finned tubes			
Position	: - Horizontal			
Capacity	[kW]	: 46950		(Calc.)
Heat Exchanging Area	[m^2]	: 2390^2		(Calc.)
Overall Heat Transfer Coefficient	[W/m^2·°C]	: 600		(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	: 33		
Passes Tube Side		: 1		
Passes Shell Side		: n.a.		
Correction Factor LMTD (min 0.75)		: 1.00		
Corrected LMTD	[°C]	: 33		
Process Conditions				
			Shell Side	Tube Side
Medium		:	Air	IPA/water
Mass Stream	[kg/s]	:	1534.3	
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	1.02	
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	30.0	80.0
Temperature OUT	[°C]	:	60.0	80.0
Pressure	[bara]	:	1.0	0.8
Material		:	C.S.	C.S.

Remarks:
23. Fan duty: 383.67 kW
24. In practice 5 air coolers would be placed

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E207				
NAME : Recycle to C101				
General Data				
Service	:	- Cooler		
Type	:	- Fixed Tube Sheets		
Position	:	- Horizontal		
Capacity	[kW]	:	6809	(Calc.)
Heat Exchanging Area	[m^2]	:	473	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	21	
Passes Tube Side		:	4	
Passes Shell Side		:	2	
Correction Factor LMTD (min 0.75)		:	0.92	
Corrected LMTD	[°C]	:	19	

Process Conditions

			Shell Side	Tube Side
Medium		:	Water	Cooling water
Mass Stream	[kg/s]	:	8.7	108.6
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		4.18
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	95.0	25.0
Temperature OUT	[°C]	:	30.0	40.0
Pressure	[bara]	:	1.0	1.0
Material		:	C.S.	C.S.

Remarks:
25.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: :	**CPD3242** December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E208				
NAME : C203 Top Cooler				
General Data				
Service	: - Condenser			
Type	: - Finned tubes			
Position	: - Horizontal			
Capacity	[kW]	:	690	(Calc.)
Heat Exchanging Area	[m^2]	:	51	(Calc.)
Overall Heat Transfer Coefficient	[W/m^2·°C]	:	500	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	27	
Passes Tube Side		:	1	
Passes Shell Side		:	n.a.	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	27	

Process Conditions

			Shell Side	Tube Side
Medium		:	Air	Cyclohexane/water
Mass Stream	[kg/s]	:	22.6	
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	1.02	
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	30.0	74.7
Temperature OUT	[°C]	:	60.0	74.7
Pressure	[bara]	:	1.0	0.8
Material		:	C.S.	C.S.

Remarks:
26. Fan duty: 8.21 kW

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E209				
NAME : C204 Reboiler				
General Data				
Service	: - Reboiler			
Type	: - Thermosyphon			
Position	: - Vertical			
Capacity	[kW]	:	5100	(Calc.)
Heat Exchanging Area	[m^2]	:	380	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	800	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	21	
Passes Tube Side		:	4	
Passes Shell Side		:	2	
Correction Factor LMTD (min 0.75)		:	0.80	
Corrected LMTD	[°C]	:	17	

Process Conditions

			Shell Side	Tube Side
Medium		:	Water	LP Steam
Mass Stream	[kg/s]	:		2.1
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	100.0	196.8
Temperature OUT	[°C]	:	100.0	101.0
Pressure	[bara]	:	1.0	4.7
Material		:	C.S.	C.S.

Remarks:
27.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

Delft. University of Technology
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E210				
NAME : C204 Condenser				
General Data				
Service	: - Condenser			
Type	: - Finned tubes			
Position	: - Horizontal			
Capacity	[kW]	:	5000	(Calc.)
Heat Exchanging Area	[m^2]	:	320	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot {}^oC$]	:	500	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	31	
Passes Tube Side		:	1	
Passes Shell Side		:	n.a.	
Correction Factor LMTD (min 0.75)		:	1.00	
Corrected LMTD	[°C]	:	31	

Process Conditions

			Shell Side	Tube Side
Medium		:	Air	Cyclohexane
Mass Stream	[kg/s]	:	163.4	0.2
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	1.02	
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	30.0	78.6
Temperature OUT	[°C]	:	60.0	78.6
Pressure	[bara]	:	1.0	0.8
Material		:	C.S.	C.S.

Remarks:
28. Fan duty: 51.4 kW

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

Delft. University of Technology
Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E211				
NAME : C201 Top feed cooler				
General Data				
Service	: - Cooler			
Type	: - Fixed Tube Sheets			
Position	: - Horizontal			
Capacity	[kW]	:	12899	(Calc.)
Heat Exchanging Area	[m^2]	:	633	(Calc.)
Overall Heat Transfer Coefficient	[W/m^2·°C]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	28	
Passes Tube Side		:	4	
Passes Shell Side		:	2	
Correction Factor LMTD (min 0.75)		:	0.96	
Corrected LMTD	[°C]	:	27	

Process Conditions

			Shell Side	Tube Side
Medium		:	Water	Cooling water
Mass Stream	[kg/s]	:	51.3	617.2
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	4.20	4.18
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	95	25
Temperature OUT	[°C]	:	34	30
Pressure	[bara]	:	1	1
Material		:	C.S.	C.S.

Remarks:
29.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E212				
NAME : C204 Bottom cooler				
General Data				
Service	: - Cooler			
Type	: - Fixed Tube Sheets			
Position	: - Horizontal			
Capacity	[kW]	:	197	(Calc.)
Heat Exchanging Area	[m^2]	:	11	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	25	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	1	
Corrected LMTD	[°C]	:	24	
Process Conditions				
			Shell Side	Tube Side
Medium		:	Water	Cooling water
Mass Stream	[kg/s]	:	1.0	9.4
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:	2.83	4.18
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	100.0	25.0
Temperature OUT	[°C]	:	30.0	30.0
Pressure	[bara]	:	1.2	1.0
Material		:	C.S.	C.S.
Remarks: 30.				

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E213				
NAME : C203 Product cooler				
General Data				
Service	: - Cooler			
Type	: - Fixed Tube Sheets			
Position	: - Horizontal			
Capacity	[kW]	:	4765	(Calc.)
Heat Exchanging Area	[m^2]	:	350	(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot {}^oC$]	:	750	(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:	20	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min 0.75)		:	1	
Corrected LMTD	[°C]	:	18	
Process Conditions				
			Shell Side	Tube Side
Medium		:	IPA	Cooling water
Mass Stream	[kg/s]	:	5.8	228.0
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot {}^oC$]	:	15.74	4.18
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	82.3	25.0
Temperature OUT	[°C]	:	30.0	30.0
Pressure	[bara]	:	1.3	1.0
Material		:	C.S.	C.S.
Remarks: 31.				

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : E211				
NAME : Water feed to C201				
General Data				
Service	: - Cooler			
Type	: - Fixed Tube Sheets			
Position	: - Horizontal			
Capacity	[kW]	:		(Calc.)
Heat Exchanging Area	[m^2]	:		(Calc.)
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	:		(Approx.)
Log. Mean Temperature Diff. (LMTD)	[°C]	:		
Passes Tube Side		:		
Passes Shell Side		:		
Correction Factor LMTD (min 0.75)		:		
Corrected LMTD	[°C]	:		
Process Conditions				
			Shell Side	Tube Side
Medium		:	Water	Cooling water
Mass Stream	[kg/s]	:		
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[$kJ/kg \cdot °C$]	:		
Heat of Evap./Condensation	[kJ/kg]	:		
Temperature IN	[°C]	:	95	25
Temperature OUT	[°C]	:	34	40
Pressure	[bara]	:		
Material		:		C.S.

Remarks:
32.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

HEAT EXCHANGER - SPECIFICATION SHEET

EQUIPMENT NUMBER : F101			In Series : -	
NAME : R101 Feed Furnace			In Parallel : -	
General Data				
Service	: Furnace			
Type	: -Fixed Tube Sheets			
Position	: -Vertical			
Capacity	[kW]	: 14918		(Calc.)
Heat Exchanging Area	[m^2]	: 497.27		
Overall Heat Transfer Coefficient	[$W/m^2 \cdot °C$]	: n.a.		
Log. Mean Temperature Diff. (LMTD)	[°C]	: n.a.		
Passes Tube Side		: 1		
Passes Shell Side		: 1		
Correction Factor LMTD (min 0.75)		:		
Corrected LMTD	[°C]	:		
Process Conditions				
			Shell Side	Tube Side
Medium		:	Combustion gas	Process stream
Mass Stream	[kg/s]	:	0.47	13.73
Mass Stream to Evaporize	[kg/s]	:		
Mass Stream to Condense	[kg/s]	:		
Average Specific Heat	[kJ/kg·°C]	:		
Heat of Evap./Condensation	[kJ/kg]	:	n.a.	n.a.
Temperature IN	[°C]	:	900	300
Temperature OUT	[°C]	:	900	650.0
Pressure	[bara]	:	1	2[1]
Material		:	CS	CS

Remarks:
1. Based on recommended flow scheme

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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Process Systems Engineering

RECIPROCAL COMPRESSOR - SPECIFICATION SHEET

EQUIPMENT NUMBERS	:	K101	Operating	: 1
NAME		R101 vacuum compressor	Installed Spare	: 0

Service : vacuum compressor
Type : Reciprocating[1]
Number : 1

Operating Conditions & Physical Data

Compressed gas			:	Propane/Propylene/Light ends
Temperature	(T)	[°C]	:	180
Density	(ρ)	[kg/m^3]	:	0.056

Power

Capacity	(ϕ_V)	[m^3/s]	:	24.2
Suction Pressure		[bara]	:	0.2
Discharge Pressure		[bara]	:	0.67
Pressure Difference	(Δp)	[bara]	:	0.47
Theoretical Power		[kW]	:	1728
Pump Efficiency		[-]	:	72%
Power at Shaft		[kW]	:	2400

Construction Details

RPM	:	3000	Nominal diameter			
Drive	:	Electrical	Suction Nozzle	[...]	:	
Type electrical motor	:		Discharge Nozzle	[...]	:	
Tension [V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction	:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate	:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling	:	Yes	If Yes			
Pressure Gauge Suction	:	No	- Seal Liquid		:	Yes/No
			- Splash Rings		:	Yes/No
Pressure Gauge Discharge	:	Yes	- Packing Type		:	
			- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m [bar]	:		- N.P.S.H. {$=p_m.\rho.g$}	[m]	:	

Construction Materials

Pump House	:	CS[2]	Wear Rings	:
Pump Rotor	:	CS	Shaft Box	:
Shaft	:	CS		
Special Provisions	:			
Operating Pressure [bara]	:		Test Pressure [bara]	:

Remarks:
1. The reciprocal type is chosen because of the low pressure desired at the entrance.
2. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL COMPRESSOR - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	K102	Operating :	1
NAME	R102 HP compressor	Installed Spare :	0

Service : HP compressor
Type : Multi stage centrifugal
Number : 1

Operating Conditions & Physical Data

Compressed gas			: Propane/Propylene/Light ends
Temperature	(T)	[°C]	: 180
Density	(ρ)	[kg/m³]	: 3.25

Power

Capacity	(ϕ_V)	[m³/s]	: 0.42
Suction Pressure		[bara]	: 0.57
Discharge Pressure		[bara]	: 36
Pressure Difference	(Δp)	[bara]	: 35.43
Theoretical Power		[kW]	: 5771
Pump Efficiency		[-]	: 72%
Power at Shaft		[kW]	: 8015

Construction Details

RPM		: 3000	Nominal diameter		
Drive		: Electrical	Suction Nozzle	[...]	:
Type electrical motor		:	Discharge Nozzle	[...]	:
Tension	[V]	: 380	Cooled Bearings		: Yes/No
Rotational direction		: Clock	Cooled Stuffing Box		: Yes/No
Foundation Plate		: Combined	Smothering Gland		: Yes/No
Flexible Coupling		: Yes	If Yes		
Pressure Gauge Suction		: No	- Seal Liquid		: Yes/No
			- Splash Rings		: Yes/No
Pressure Gauge Discharge		: Yes	- Packing Type		:
			- Mechanical Seal		: Yes/No
Min. Overpressure above p_v/p_m	[bar]	:	- N.P.S.H. $\{=p_m \cdot \rho . g\}$	[m]	:

Construction Materials

Pump House		: CS[1]	Wear Rings		:
Pump Rotor		: CS	Shaft Box		:
Shaft		: CS			
Special Provisions		:			
Operating Pressure	[bara]	:	Test Pressure	[bara]	:

Remarks:
3. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL COMPRESSOR - SPECIFICATION SHEET

EQUIPMENT NUMBERS	:	K104	Operating	: 1
NAME		R102 recycle compressor	Installed Spare	: 0

Service : recycle compressor
Type : Axial flow[1]
Number : 1

Operating Conditions & Physical Data

Compressed gas			:	Propane/Propylene/Water/IPA
Temperature	(T)	[°C]	:	224[3]
Density	(ρ)	[kg/m^3]	:	69.90

Power

Capacity	(ϕ_V)	[m^3/s]	:	1.17
Suction Pressure		[bara]	:	29[3]
Discharge Pressure		[bara]	:	34
Pressure Difference	(Δp)	[bara]	:	5
Theoretical Power		[kW]	:	648
Pump Efficiency		[-]	:	72%
Power at Shaft		[kW]	:	900

Construction Details

RPM		:	3000	Nominal diameter			
Drive		:	MP Steam	Suction Nozzle	[...]	:	
Type electrical motor		:		Discharge Nozzle	[...]	:	
Tension	[V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate		:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling		:	Yes	If Yes			
Pressure Gauge Suction		:	No	- Seal Liquid		:	Yes/No
				- Splash Rings		:	Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:	
				- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. {$=p_m.\rho.g$}	[m]	:	

Construction Materials

Pump House		:	CS[2]	Wear Rings		:
Pump Rotor		:	CS	Shaft Box		:
Shaft		:	CS			
Special Provisions		:				
Operating Pressure	[bara]	:		Test Pressure	[bara]	:

Remarks:

4. The axial flow type was chosen because of the high volume flow
5. CS is carbon steel
6. Based on recommended flow scheme

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P101		Operating : 1	
NAME C101 Reflux pump		Installed Spare : 0	
Service : Reflux pump			
Type : Centrifugal			
Number : 1			
Operating Conditions & Physical Data			
Pumped liquid : Propane/Propylene			
Temperature (T) [°C] : -33			
Density (ρ) [kg/m³] : 391.0			
Power			
Capacity (ϕ_V) [m³/s] : 2.76E-02			
Suction Pressure [bara] : 34.8			
Discharge Pressure [bara] : 36.8			
Pressure Difference (Δp) [bara] : 2.0			
Theoretical Power [kW] : 5.52			
Pump Efficiency [-] : 77%			
Power at Shaft [kW] : 7.17			
Construction Details			
RPM :		Nominal diameter	
Drive : Electrical		Suction Nozzle [...] :	
Type electrical motor :		Discharge Nozzle [...] :	
Tension [V] : 380		Cooled Bearings :	Yes/No
Rotational direction : Clock		Cooled Stuffing Box :	Yes/No
Foundation Plate : Combined		Smothering Gland :	Yes/No
Flexible Coupling : Yes		If Yes	
Pressure Gauge Suction : No		- Seal Liquid :	Yes/No
		- Splash Rings :	Yes/No
Pressure Gauge Discharge : Yes		- Packing Type :	
		- Mechanical Seal :	Yes/No
Min. Overpressure above p_v/p_m [bar] :		- N.P.S.H. [m] : {$=p_m.\rho.g$}	
Construction Materials			
Pump House : CS[1]		Wear Rings :	
Pump Rotor : CS		Shaft Box :	
Shaft : CS			
Special Provisions :			
Operating Pressure [bara] :		Test Pressure [bara] :	

Remarks:

7. CS is carbon steel
8. The V101 accumulator has been designed to have a holdup time of 5 instead of 15 minutes; as a consequence, a tripping system must be installed on this pump.

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS	:	P102	Operating	:	1
NAME		R102 Water feed pump	Installed Spare	:	0

Service : Water feed pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid			:	Water
Temperature	(T)	[°C]	:	77.6
Density	(ρ)	[kg/m^3]	:	1000.0

Power

Capacity	(ϕ_V)	[m^3/s]	:	8.68E-03
Suction Pressure		[bara]	:	1.0
Discharge Pressure		[bara]	:	34.0
Pressure Difference	(Δp)	[bara]	:	33.0
Theoretical Power		[kW]	:	28.64
Pump Efficiency		[-]	:	70%
Power at Shaft		[kW]	:	40.91

Construction Details

RPM		:		Nominal diameter			
Drive		:	Electrical	Suction Nozzle	[...]	:	
Type electrical motor		:		Discharge Nozzle	[...]	:	
Tension	[V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate		:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling		:	Yes	If Yes			
Pressure Gauge Suction		:	No	- Seal Liquid		:	Yes/No
				- Splash Rings		:	Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:	
				- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. $\{=p_m.\rho.g\}$	[m]	:	

Construction Materials

Pump House		:	CS[1]	Wear Rings		:
Pump Rotor		:	CS	Shaft Box		:
Shaft		:	CS			
Special Provisions		:				
Operating Pressure	[bara]	:		Test Pressure	[bara]	:

Remarks:
9. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	P103	Operating :	1
NAME	C102 Absorbent feed pump	Installed Spare :	0

Service	:	Absorbent feed pump
Type	:	Centrifugal
Number	:	1

Operating Conditions & Physical Data

Pumped liquid			:	Water
Temperature	(T)	[°C]	:	30
Density	(ρ)	[kg/m^3]	:	1000.0

Power

Capacity	(ϕ_V)	[m^3/s]	:	2.62E-02
Suction Pressure		[bara]	:	1.0
Discharge Pressure		[bara]	:	29.0[2]
Pressure Difference	(Δp)	[bara]	:	28.0
Theoretical Power		[kW]	:	73.26
Pump Efficiency		[-]	:	76%
Power at Shaft		[kW]	:	96.40

Construction Details

RPM		:		Nominal diameter			
Drive		:	Electrical	Suction Nozzle	[...]	:	
Type electrical motor		:		Discharge Nozzle	[...]	:	
Tension	[V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate		:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling		:	Yes	If Yes			
Pressure Gauge Suction		:	No	- Seal Liquid		:	Yes/No
				- Splash Rings		:	Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:	
				- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. {$=p_m.\rho.g$}	[m]	:	

Construction Materials

Pump House		:	CS[1]	Wear Rings		:
Pump Rotor		:	CS	Shaft Box		:
Shaft		:	CS			
Special Provisions		:				
Operating Pressure	[bara]	:		Test Pressure	[bara]	:

Remarks:

10. CS is carbon steel
11. Based on recommended flow scheme

Designers:	ME Brons	JHM Jansen	Project ID-number :	**CPD3242**
	MP Hoff	EPE Seveke	Date :	December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P201		Operating : 1	
NAME : C201 Bottom pump		Installed Spare : 0	
Service : Bottom pump			
Type : Centrifugal			
Number : 1			
Operating Conditions & Physical Data			
Pumped liquid			: Water/IPA
Temperature	(T)	[°C]	: 86
Density	(ρ)	[kg/m³]	: 898.1
Power			
Capacity	(ϕ_V)	[m³/s]	: 1.16E-01
Suction Pressure		[bara]	: 1.4
Discharge Pressure		[bara]	: 3.4
Pressure Difference	(Δp)	[bara]	: 2.0
Theoretical Power		[kW]	: 23.25
Pump Efficiency		[-]	: 85%
Power at Shaft		[kW]	: 27.35

Construction Details

RPM :		Nominal diameter	
Drive :	Electrical	Suction Nozzle [...] :	
Type electrical motor :		Discharge Nozzle [...] :	
Tension [V] :	380	Cooled Bearings :	Yes/No
Rotational direction :	Clock	Cooled Stuffing Box :	Yes/No
Foundation Plate :	Combined	Smothering Gland :	Yes/No
Flexible Coupling :	Yes	If Yes	
Pressure Gauge Suction :	No	- Seal Liquid :	Yes/No
		- Splash Rings :	Yes/No
Pressure Gauge Discharge :	Yes	- Packing Type :	
		- Mechanical Seal :	Yes/No
Min. Overpressure above p_v/p_m [bar] :		- N.P.S.H. [m] : {$=p_m \cdot \rho . g$}	

Construction Materials

Pump House :	CS[1]	Wear Rings :	
Pump Rotor :	CS	Shaft Box :	
Shaft :	CS		
Special Provisions :			
Operating Pressure [bara] :		Test Pressure [bara] :	

Remarks:
12. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P202		Operating	: 1
NAME	V201 Inorganic phase pump	Installed Spare	: 0
Service : Inorganic phase pump			
Type : Centrifugal			
Number : 1			

Operating Conditions & Physical Data

Pumped liquid			: Water/IPA
Temperature	(T)	[°C]	: 34
Density	(ρ)	[kg/m^3]	: 296.0

Power

Capacity	(ϕ_V)	[m^3/s]	: 2.02E-02
Suction Pressure		[bara]	: 1.0
Discharge Pressure		[bara]	: 3.0
Pressure Difference	(Δp)	[bara]	: 2.0
Theoretical Power		[kW]	: 4.04
Pump Efficiency		[-]	: 75%
Power at Shaft		[kW]	: 5.39

Construction Details

RPM		:	Nominal diameter		
Drive		: Electrical	Suction Nozzle	[...]	:
Type electrical motor		:	Discharge Nozzle	[...]	:
Tension	[V]	: 380	Cooled Bearings		: Yes/No
Rotational direction		: Clock	Cooled Stuffing Box		: Yes/No
Foundation Plate		: Combined	Smothering Gland		: Yes/No
Flexible Coupling		: Yes	If Yes		
Pressure Gauge Suction		: No	- Seal Liquid		: Yes/No
			- Splash Rings		: Yes/No
Pressure Gauge Discharge		: Yes	- Packing Type		:
			- Mechanical Seal		: Yes/No
Min. Overpressure above p_v/p_m	[bar]	:	- N.P.S.H. $\{=p_m.\rho.g\}$	[m]	:

Construction Materials

Pump House		: CS[1]	Wear Rings		:
Pump Rotor		: CS	Shaft Box		:
Shaft		: CS			
Special Provisions		:			
Operating Pressure	[bara]	:	Test Pressure	[bara]	:

Remarks:
13. CS is carbon steel

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P203		Operating : 1	
NAME V201 Organic phase pump		Installed Spare : 0	
Service : Organic phase pump			
Type : Centrifugal			
Number : 1			
Operating Conditions & Physical Data			
Pumped liquid		:	NPA
Temperature (T)	[°C]	:	34
Density (ρ)	[kg/m³]	:	1156.0
Power			
Capacity (ϕ_V)	[m³/s]	:	4.61E-06
Suction Pressure	[bara]	:	1.0
Discharge Pressure	[bara]	:	3.0
Pressure Difference (Δp)	[bara]	:	2.0
Theoretical Power	[kW]	:	0.00
Pump Efficiency	[-]	:	40%
Power at Shaft	[kW]	:	2.3E-03

Construction Details

RPM :		Nominal diameter	
Drive :	Electrical	Suction Nozzle [...] :	
Type electrical motor :		Discharge Nozzle [...] :	
Tension [V] :	380	Cooled Bearings :	Yes/No
Rotational direction :	Clock	Cooled Stuffing Box :	Yes/No
Foundation Plate :	Combined	Smothering Gland :	Yes/No
Flexible Coupling :	Yes	If Yes	
Pressure Gauge Suction :	No	- Seal Liquid :	Yes/No
		- Splash Rings :	Yes/No
Pressure Gauge Discharge :	Yes	- Packing Type :	
		- Mechanical Seal :	Yes/No
Min. Overpressure [bar] above p_v/p_m :		- N.P.S.H. [m] {$=p_m \cdot \rho \cdot g$} :	

Construction Materials

Pump House :	CS[1]	Wear Rings :	
Pump Rotor :	CS	Shaft Box :	
Shaft :	CS		
Special Provisions :			
Operating Pressure [bara] :		Test Pressure [bara] :	

Remarks:
14. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS	:	P204	Operating	:	1
NAME		C202 Bottom pump	Installed Spare	:	0

Service	:	Bottom pump
Type	:	Centrifugal
Number	:	1

Operating Conditions & Physical Data

Pumped liquid			:	Water
Temperature	(T)	[°C]	:	95
Density	(ρ)	[kg/m³]	:	1000.0

Power

Capacity	(ϕ_V)	[m³/s]	:	1.04E-01
Suction Pressure		[bara]	:	1.5
Discharge Pressure		[bara]	:	3.5
Pressure Difference	(Δp)	[bara]	:	2.0
Theoretical Power		[kW]	:	20.89
Pump Efficiency		[-]	:	84%
Power at Shaft		[kW]	:	24.87

Construction Details

RPM		:		Nominal diameter			
Drive		:	Electrical	Suction Nozzle	[...]	:	
Type electrical motor		:		Discharge Nozzle	[...]	:	
Tension	[V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate		:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling		:	Yes	If Yes			
Pressure Gauge Suction		:	No	- Seal Liquid		:	Yes/No
				- Splash Rings		:	Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:	
				- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. {$=p_m.\rho.g$}	[m]	:	

Construction Materials

Pump House		:	CS[1]	Wear Rings		:	
Pump Rotor		:	CS	Shaft Box		:	
Shaft		:	CS				
Special Provisions		:					
Operating Pressure	[bara]	:		Test Pressure	[bara]	:	

Remarks:
15. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS	:	P205	Operating	:	1
NAME		C202 Reflux pump	Installed Spare	:	0

Service	:	Reflux pump
Type	:	Centrifugal
Number	:	1

Operating Conditions & Physical Data

Pumped liquid			:	IPA/NPA/Water
Temperature	(T)	[°C]	:	80
Density	(ρ)	[kg/m^3]	:	753.7

Power

Capacity	(ϕ_V)	[m^3/s]	:	6.10E-02
Suction Pressure		[bara]	:	0.7
Discharge Pressure		[bara]	:	2.7
Pressure Difference	(Δp)	[bara]	:	2.0
Theoretical Power		[kW]	:	12.21
Pump Efficiency		[-]	:	80%
Power at Shaft		[kW]	:	15.26

Construction Details

RPM		:		Nominal diameter			
Drive		:	Electrical	Suction Nozzle	[...]	:	
Type electrical motor		:		Discharge Nozzle	[...]	:	
Tension	[V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate		:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling		:	Yes	If Yes			
Pressure Gauge Suction		:	No	- Seal Liquid		:	Yes/No
				- Splash Rings		:	Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:	
				- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. $\{=p_m.\rho.g\}$	[m]	:	

Construction Materials

Pump House		:	CS[1]	Wear Rings		:	
Pump Rotor		:	CS	Shaft Box		:	
Shaft		:	CS				
Special Provisions		:					
Operating Pressure	[bara]	:		Test Pressure	[bara]	:	

Remarks:

16. CS is carbon steel
17. The V202 accumulator has been designed to have a holdup time of 5 instead of 15 minutes; as a consequence, a tripping system must be installed on this pump.

Designers:	ME Brons	JHM Jansen	Project ID-number	:	**CPD3242**
	MP Hoff	EPE Seveke	Date	:	December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P206		Operating : 1	
NAME : C202 Top transfer pump		Installed Spare : 0	
Service : Top transfer pump			
Type : Centrifugal			
Number : 1			
Operating Conditions & Physical Data			
Pumped liquid		: IPA/NPA/Water	
Temperature (T) [°C]		: 80	
Density (ρ) [kg/m^3]		: 753.7	
Power			
Capacity (ϕ_V) [m^3/s]		: 1.53E-02	
Suction Pressure [bara]		: 0.7	
Discharge Pressure [bara]		: 2.7	
Pressure Difference (Δp) [bara]		: 2.0	
Theoretical Power [kW]		: 3.05	
Pump Efficiency [-]		: 75%	
Power at Shaft [kW]		: 4.07	
Construction Details			
RPM	:	Nominal diameter	
Drive	: Electrical	Suction Nozzle […]	:
Type electrical motor	:	Discharge Nozzle […]	:
Tension [V]	: 380	Cooled Bearings	: Yes/No
Rotational direction	: Clock	Cooled Stuffing Box	: Yes/No
Foundation Plate	: Combined	Smothering Gland	: Yes/No
Flexible Coupling	: Yes	If Yes	
Pressure Gauge Suction	: No	- Seal Liquid	: Yes/No
		- Splash Rings	: Yes/No
Pressure Gauge Discharge	: Yes	- Packing Type	:
		- Mechanical Seal	: Yes/No
Min. Overpressure above p_v/p_m [bar]	:	- N.P.S.H. [m] {$=p_m.\rho.g$}	:
Construction Materials			
Pump House	: CS[1]	Wear Rings	:
Pump Rotor	: CS	Shaft Box	:
Shaft	: CS		
Special Provisions	:		
Operating Pressure [bara]	:	Test Pressure [bara]	:

Remarks:
18. CS is carbon steel

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	P207	Operating :	1
NAME	C202 Product transfer pump	Installed Spare :	0

Service : Product transfer pump
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid			: IPA/Water
Temperature	(T)	[°C]	: 80
Density	(ρ)	[kg/m^3]	: 755.2

Power

Capacity	(ϕ_V)	[m^3/s]	: 9.00E-03
Suction Pressure		[bara]	: 0.7
Discharge Pressure		[bara]	: 2.7
Pressure Difference	(Δp)	[bara]	: 2.0
Theoretical Power		[kW]	: 1.80
Pump Efficiency		[-]	: 70%
Power at Shaft		[kW]	: 2.57

Construction Details

RPM		:	Nominal diameter		
Drive		: Electrical	Suction Nozzle	[...]	:
Type electrical motor		:	Discharge Nozzle	[...]	:
Tension	[V]	: 380	Cooled Bearings		: Yes/No
Rotational direction		: Clock	Cooled Stuffing Box		: Yes/No
Foundation Plate		: Combined	Smothering Gland		: Yes/No
Flexible Coupling		: Yes	If Yes		
Pressure Gauge Suction		: No	- Seal Liquid		: Yes/No
			- Splash Rings		: Yes/No
Pressure Gauge Discharge		: Yes	- Packing Type		:
			- Mechanical Seal		: Yes/No
Min. Overpressure above p_v/p_m	[bar]	:	- N.P.S.H. $\{=p_m.\rho.g\}$	[m]	:

Construction Materials

Pump House		: CS[1]	Wear Rings	:
Pump Rotor		: CS	Shaft Box	:
Shaft		: CS		
Special Provisions		:		
Operating Pressure	[bara]	:	Test Pressure [bara]	:

Remarks:
19. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS	:	P208	Operating	:	1
NAME		C203 Bottom pump	Installed Spare	:	0

Service	:	Bottom pump
Type	:	Centrifugal
Number	:	1

Operating Conditions & Physical Data

Pumped liquid			:	IPA
Temperature	(T)	[°C]	:	82.3
Density	(ρ)	[kg/m^3]	:	761.9

Power

Capacity	(ϕ_V)	[m^3/s]	:	7.61E-03
Suction Pressure		[bara]	:	1.3
Discharge Pressure		[bara]	:	15.0
Pressure Difference	(Δp)	[bara]	:	13.8
Theoretical Power		[kW]	:	10.47
Pump Efficiency		[-]	:	70%
Power at Shaft		[kW]	:	14.95

Construction Details

RPM		:		Nominal diameter			
Drive		:	Electrical	Suction Nozzle	[...]	:	
Type electrical motor		:		Discharge Nozzle	[...]	:	
Tension	[V]	:	380	Cooled Bearings		:	Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		:	Yes/No
Foundation Plate		:	Combined	Smothering Gland		:	Yes/No
Flexible Coupling		:	Yes	If Yes			
Pressure Gauge Suction		:	No	- Seal Liquid		:	Yes/No
				- Splash Rings		:	Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:	
				- Mechanical Seal		:	Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. {$=p_m.\rho.g$}	[m]	:	

Construction Materials

Pump House		:	CS[1]	Wear Rings		:
Pump Rotor		:	CS	Shaft Box		:
Shaft		:	CS			
Special Provisions		:				
Operating Pressure	[bara]	:		Test Pressure	[bara]	:

Remarks:
20. CS is carbon steel

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: :	**CPD3242** December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	P209	Operating	: 1
NAME	V203 Inorganic phase pump	Installed Spare	: 0
Service :	Inorganic phase pump		
Type :	Centrifugal		
Number :	1		

Operating Conditions & Physical Data

Pumped liquid			: Water/Cyclohexane
Temperature	(T)	[°C]	: 75
Density	(ρ)	[kg/m^3]	: 1000.0

Power

Capacity	(ϕ_V)	[m^3/s]	: 1.24E-03
Suction Pressure		[bara]	: 1.0
Discharge Pressure		[bara]	: 3.0
Pressure Difference	(Δp)	[bara]	: 2.0
Theoretical Power		[kW]	: 0.25
Pump Efficiency		[-]	: 50%
Power at Shaft		[kW]	: 0.49

Construction Details

RPM		:	Nominal diameter		
Drive		: Electrical	Suction Nozzle	[...]	:
Type electrical motor		:	Discharge Nozzle	[...]	:
Tension	[V]	: 380	Cooled Bearings		: Yes/No
Rotational direction		: Clock	Cooled Stuffing Box		: Yes/No
Foundation Plate		: Combined	Smothering Gland		: Yes/No
Flexible Coupling		: Yes	If Yes		
Pressure Gauge Suction		: No	- Seal Liquid		: Yes/No
			- Splash Rings		: Yes/No
Pressure Gauge Discharge		: Yes	- Packing Type		:
			- Mechanical Seal		: Yes/No
Min. Overpressure above p_v/p_m	[bar]	:	- N.P.S.H. {$=p_m.\rho.g$}	[m]	:

Construction Materials

Pump House		: CS[1]	Wear Rings		:
Pump Rotor		: CS	Shaft Box		:
Shaft		: CS			
Special Provisions		:			
Operating Pressure	[bara]	:	Test Pressure	[bara]	:

Remarks:
21. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P210		Operating : 1	
NAME : V203 Organic phase pump		Installed Spare : 0	
Service : Organic phase pump			
Type : Centrifugal			
Number : 1			
Operating Conditions & Physical Data			
Pumped liquid		:	Cyclhexane
Temperature (T)	[°C]	:	75
Density (ρ)	[kg/m³]	:	747.6
Power			
Capacity (ϕ_V)	[m³/s]	:	3.65E-02
Suction Pressure	[bara]	:	1.0
Discharge Pressure	[bara]	:	3.0
Pressure Difference (Δp)	[bara]	:	2.0
Theoretical Power	[kW]	:	7.30
Pump Efficiency	[-]	:	80%
Power at Shaft	[kW]	:	9.13

Construction Details

RPM :		Nominal diameter		
Drive :	Electrical	Suction Nozzle	[...] :	
Type electrical motor :		Discharge Nozzle	[...] :	
Tension [V] :	380	Cooled Bearings	:	Yes/No
Rotational direction :	Clock	Cooled Stuffing Box	:	Yes/No
Foundation Plate :	Combined	Smothering Gland	:	Yes/No
Flexible Coupling :	Yes	If Yes		
Pressure Gauge Suction :	No	- Seal Liquid	:	Yes/No
		- Splash Rings	:	Yes/No
Pressure Gauge Discharge :	Yes	- Packing Type	:	
		- Mechanical Seal	:	Yes/No
Min. Overpressure above p_v/p_m [bar] :		- N.P.S.H. {$=p_m.\rho.g$}	[m] :	

Construction Materials

Pump House :	CS[1]	Wear Rings :	
Pump Rotor :	CS	Shaft Box :	
Shaft :	CS		
Special Provisions :			
Operating Pressure [bara] :		Test Pressure [bara] :	

Remarks:
22. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number :	**CPD3242**
	MP Hoff	EPE Seveke	Date :	December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P211		Operating	: 1
NAME : C204 Bottom pump		Installed Spare	: 0
Service : Bottom pump			
Type : Centrifugal			
Number : 1			

Operating Conditions & Physical Data

Pumped liquid			:	Water/Cyclhexane
Temperature	(T)	[°C]	:	100
Density	(ρ)	[kg/m^3]	:	1000.0

Power

Capacity	(ϕ_V)	[m^3/s]	:	1.00E-03
Suction Pressure		[bara]	:	1.2
Discharge Pressure		[bara]	:	3.2
Pressure Difference	(Δp)	[bara]	:	2.0
Theoretical Power		[kW]	:	0.20
Pump Efficiency		[-]	:	45%
Power at Shaft		[kW]	:	0.44

Construction Details

RPM		:		Nominal diameter		
Drive		:	Electrical	Suction Nozzle	[...]	:
Type electrical motor		:		Discharge Nozzle	[...]	:
Tension	[V]	:	380	Cooled Bearings		: Yes/No
Rotational direction		:	Clock	Cooled Stuffing Box		: Yes/No
Foundation Plate		:	Combined	Smothering Gland		: Yes/No
Flexible Coupling		:	Yes	If Yes		
Pressure Gauge Suction		:	No	- Seal Liquid		: Yes/No
				- Splash Rings		: Yes/No
Pressure Gauge Discharge		:	Yes	- Packing Type		:
				- Mechanical Seal		: Yes/No
Min. Overpressure above p_v/p_m	[bar]	:		- N.P.S.H. {$=p_m.\rho.g$}	[m]	:

Construction Materials

Pump House		: CS[1]	Wear Rings		:
Pump Rotor		: CS	Shaft Box		:
Shaft		: CS			
Special Provisions		:			
Operating Pressure	[bara]	:	Test Pressure	[bara]	:

Remarks:
23. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	P212	Operating :	1
NAME	C204 Reflux pump	Installed Spare :	0
Service :	Bottom pump		
Type :	Centrifugal		
Number :	1		
Operating Conditions & Physical Data			
Pumped liquid	: Cyclohexane/Water		
Temperature (T) [°C]	: 78.6		
Density (ρ) [kg/m^3]	: 1000.0		
Power			
Capacity (ϕ_V) [m^3/s]	: 1.10E-03		
Suction Pressure [bara]	: 0.8		
Discharge Pressure [bara]	: 2.8		
Pressure Difference (Δp) [bara]	: 2.0		
Theoretical Power [kW]	: 0.22		
Pump Efficiency [-]	: 45%		
Power at Shaft [kW]	: 0.49		
Construction Details			
RPM :		Nominal diameter	
Drive :	Electrical	Suction Nozzle [...] :	
Type electrical motor :		Discharge Nozzle [...] :	
Tension [V] :	380	Cooled Bearings :	Yes/No
Rotational direction :	Clock	Cooled Stuffing Box :	Yes/No
Foundation Plate :	Combined	Smothering Gland :	Yes/No
Flexible Coupling :	Yes	If Yes	
Pressure Gauge Suction :	No	- Seal Liquid :	Yes/No
		- Splash Rings :	Yes/No
Pressure Gauge Discharge :	Yes	- Packing Type :	
		- Mechanical Seal :	Yes/No
Min. Overpressure above p_v/p_m [bar] :		- N.P.S.H. [m] : {$=p_m.\rho.g$}	
Construction Materials			
Pump House :	CS[1]	Wear Rings :	
Pump Rotor :	CS	Shaft Box :	
Shaft :	CS		
Special Provisions :			
Operating Pressure [bara] :		Test Pressure [bara] :	

Remarks:
24. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS : P213		Operating : 1	
NAME C204 Transfer pump		Installed Spare : 0	
Service : Transfer pump			
Type : Centrifugal			
Number : 1			
Operating Conditions & Physical Data			
Pumped liquid	: Cyclohexane/Water		
Temperature (T) [°C]	: 78.6		
Density (ρ) [kg/m^3]	: 1000.0		
Power			
Capacity (ϕ_V) [m^3/s]	: 2.40E-04		
Suction Pressure [bara]	: 0.8		
Discharge Pressure [bara]	: 2.8		
Pressure Difference (Δp) [bara]	: 2.0		
Theoretical Power [kW]	: 0.05		
Pump Efficiency [-]	: 40%		
Power at Shaft [kW]	: 0.12		
Construction Details			
RPM	:	Nominal diameter	
Drive	: Electrical	Suction Nozzle [...]	:
Type electrical motor	:	Discharge Nozzle [...]	:
Tension [V]	: 380	Cooled Bearings	: Yes/No
Rotational direction	: Clock	Cooled Stuffing Box	: Yes/No
Foundation Plate	: Combined	Smothering Gland	: Yes/No
Flexible Coupling	: Yes	If Yes	
Pressure Gauge Suction	: No	- Seal Liquid	: Yes/No
		- Splash Rings	: Yes/No
Pressure Gauge Discharge	: Yes	- Packing Type	:
		- Mechanical Seal	: Yes/No
Min. Overpressure above p_v/p_m [bar]	:	- N.P.S.H. [m] $\{=p_m.\rho.g\}$	:
Construction Materials			
Pump House	: CS[1]	Wear Rings	:
Pump Rotor	: CS	Shaft Box	:
Shaft	: CS		
Special Provisions	:		
Operating Pressure [bara]	:	Test Pressure [bara]	:

Remarks:
25. CS is carbon steel

Designers:	ME Brons	JHM Jansen	Project ID-number	: **CPD3242**
	MP Hoff	EPE Seveke	Date	: December 22nd, '99

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Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	P214	Operating	: 1
NAME	C205 Bottom pump	Installed Spare	: 0
Service : Bottom pump Type : Centrifugal Number : 1			
Operating Conditions & Physical Data			
Pumped liquid	: IPA/Water/NPA		
Temperature (T) [°C]	: 80		
Density (ρ) [kg/m³]	: 753.2		
Power			
Capacity (ϕ_V) [m³/s]	: 7.97E-03		
Suction Pressure [bara]	: 1.1		
Discharge Pressure [bara]	: 3.1		
Pressure Difference (Δp) [bara]	: 2.0		
Theoretical Power [kW]	: 1.59		
Pump Efficiency [-]	: 70%		
Power at Shaft [kW]	: 2.28		
Construction Details			
RPM	:	Nominal diameter	
Drive	: Electrical	Suction Nozzle [...]	:
Type electrical motor	:	Discharge Nozzle [...]	:
Tension [V]	: 380	Cooled Bearings	: Yes/No
Rotational direction	: Clock	Cooled Stuffing Box	: Yes/No
Foundation Plate	: Combined	Smothering Gland	: Yes/No
Flexible Coupling	: Yes	If Yes	
Pressure Gauge Suction	: No	- Seal Liquid	: Yes/No
		- Splash Rings	: Yes/No
Pressure Gauge Discharge	: Yes	- Packing Type	:
		- Mechanical Seal	: Yes/No
Min. Overpressure above p_v/p_m [bar]	:	- N.P.S.H. [m] $\{=p_m.\rho.g\}$	:
Construction Materials			
Pump House	: CS[1]	Wear Rings	:
Pump Rotor	: CS	Shaft Box	:
Shaft	: CS		
Special Provisions	:		
Operating Pressure [bara]	:	Test Pressure [bara]	:

Remarks:
26. CS is carbon steel

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

Delft, University of Technology
Process Systems Engineering

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBERS :	P215	Operating :	1
NAME	C205 Reflux pump	Installed Spare :	0
Service : Reflux pump Type : Centrifugal Number : 1			
Operating Conditions & Physical Data			
Pumped liquid	: IPA/Water		
Temperature (T) [°C]	: 80		
Density (ρ) [kg/m^3]	: 753.2		
Power			
Capacity (ϕ_V) [m^3/s]	: 1.46E-02		
Suction Pressure [bara]	: 0.8		
Discharge Pressure [bara]	: 2.8		
Pressure Difference (Δp) [bara]	: 2.0		
Theoretical Power [kW]	: 2.92		
Pump Efficiency [-]	: 75%		
Power at Shaft [kW]	: 3.89		
Construction Details			
RPM	:	Nominal diameter	
Drive	: Electrical	Suction Nozzle [...]	:
Type electrical motor	:	Discharge Nozzle [...]	:
Tension [V]	: 380	Cooled Bearings	: Yes/No
Rotational direction	: Clock	Cooled Stuffing Box	: Yes/No
Foundation Plate	: Combined	Smothering Gland	: Yes/No
Flexible Coupling	: Yes	If Yes	
Pressure Gauge Suction	: No	- Seal Liquid	: Yes/No
		- Splash Rings	: Yes/No
Pressure Gauge Discharge	: Yes	- Packing Type	:
		- Mechanical Seal	: Yes/No
Min. Overpressure [bar] above p_v/p_m	:	- N.P.S.H. [m] {$=p_m.\rho.g$}	:
Construction Materials			
Pump House	: CS[1]	Wear Rings	:
Pump Rotor	: CS	Shaft Box	:
Shaft	: CS		
Special Provisions	:		
Operating Pressure [bara]	:	Test Pressure [bara]	:

Remarks:

27. CS is carbon steel
28. The V205 accumulator has been designed to have a holdup time of 5 instead of 15 minutes; as a consequence, a tripping system must be installed on this pump.

Designers:	ME Brons MP Hoff	JHM Jansen EPE Seveke	Project ID-number Date	: **CPD3242** : December 22nd, '99

Hazard and Operability Study (HAZOP)

Table 7.1-1: HAZOP study for the dehydrogenation reactor R101

Vessel – **Reactor**

Intention – Dehydrogenation of propane, 0.33 - 0.66 bara, 650 °C

Guide Word	Deviation	Possible Causes	Consequences	Action Required
Not, No	NO PROPANE FEED FLOW	(1) No propane feed outside battery limit	Loss of feed to reaction section. Process on whole plant is halted.	(a) Ensure good communications with operator outside battery limit
		(2) Line blockage	As for (1)	(b) Institute regular inspection and maintenance of piping
		(3) Line rupture	As for (1), propane discharged into industrial area, possible explosion risk.	(c) As for (b); remain in close contact with fire brigade
	NO CATALYST REGENERATION	(4) No fuel gas feed outside battery limit	Too little heat for reaction; too little catalyst activity for reaction; process halted.	(d) As for (a)
		(5) Line blockage	As for (4)	(e) Institute regular inspection and maintenance of piping
		(6) Line rupture	As for (4), fuel gas discharged into industrial area	(f) As for (c)
More	MORE PROPANE FEED FLOW	(7) Increase of propane feed outside battery limit	Lower residence times in reactor, off spec production	(g) As for (a)
	MORE T	(8) Autoignition of material inside reactor due to leakage of air / oxygen	As for (3)	(h) Institute regular inspection and maintenance of vacuum seals
	MORE P	(9) Vacuum seals failure	Conversion drop, ultimately resulting in large overall recycles	(i) As for (h)
		(10) Compressor K102 failure	As for (8)	(j) Institute regular inspection and maintenance of compressor K102
Less	LESS PROPANE FEED FLOW	(11) Decrease of propane feed outside battery limit	Higher residence times in reactor, higher conversions	(k) As for (a)
	LESS T	(12) Too little fuel gas stream to pre-reactor furnace F101	Lower temperature in the reactor, lower conversions.	(l) As for (a) and (e)

Table 7.1-2: HAZOP study for the light ends column C101

Vessel – **Light ends column**

Intention – removal of light ends, 35 bara, -33/+77 °C

Guide Word	Deviation	Possible Causes	Consequences	Action Required
Not, No	NO FEED FLOW	(1) K101 or K102 failure	Column runs dry. Pumps run dry. Downstream process halted.	(a) Institute regular inspection and maintenance of compressors K101 and K102
		(2) Line blockage	As for (1)	(b) Institute regular inspection and maintenance of piping
		(3) Line rupture	As for (1), propane and propylene at high pressure discharged into industrial area	(c) As for (b), remain in close contact with fire brigade
	NO EFFLUENT FLOW	(4) Downstream line blockage	Liquid level rises, downstream process has to be stopped.	(d) As for (b)
	NO REFLUX FLOW	(5) Pump P101 failure	Rectification section runs dry; column operates at lower efficiency	(e) Have spare pump ready
		(6) Malfunctioning of K103 in condenser loop	As for (5), no condensation so all top product leaves V101 as gas	(f) Institute regular inspection and maintenance of compressor K103
More	MORE FEED FLOW	(7) Non-suppressed upstream fluctuations	Risk of weeping; column probably operates at lower efficiency	(g) Have adequate flow control system
	MORE T	(8) Insufficient cooling by E108	Probably loss of efficiency	(h) Have cooling overcapacity at E108
	MORE P	(9) K102 pressure control failure	Hardly any, within reasonable pressure differences	(i) None
Less	LESS FEED FLOW	(10) As for (7)	Risk of column running dry and thus loosing efficiency	(j) As for (g)
	LESS T	(11) Too much cooling by E108	Relative position of feed stage no longer optimal, loss of efficiency	(k) Have adequate control at E108
	LESS P	(12) Compressor K102 failure	As for (9)	(l) As for (f)
		(13) Pressure leakage through high-pressure seals failure	As for (9)	(m) Institute regular inspection and maintenance of seals
		(14) Line rupture in top section	Risk of BLEVE explosion due to ignition of expanding liquid propylene	(n) As for ©

Table 7.1-3: HAZOP study for the hydration reactor R102

Vessel – **Reactor** *Intention* – Hydration of propylene, 34 bara, 165 °C				
Guide Word	**Deviation**	**Possible Causes**	**Consequences**	**Action Required**
Not, No	NO PROPYLENE FEED FLOW	(1) Problem at upstream units	Operation must be stopped. Downstream process is halted.	(a) See possible problem analysis with solutions for upstream units (previous tables)
		(2) Line blockage	As for (1)	(b) Institute regular inspection and maintenance of piping
		(3) Line rupture	As for (1), propane and propylene discharged into industrial area, possible explosion risk.	(c) As for (b); remain in close contact with fire brigade
	NO CATALYST ACTIVITY	(4) Catalyst wash out due to too much water flow	No or almost no reaction	(d) Have adequate water flow control installed
More	MORE PROPYLENE FEED FLOW	(5) Non-suppressed upstream fluctuations	Lower residence times in reactor, off spec production	(e) Have adequate flow controls upstream
	MORE T	(6) E112 control failure	Risk of polymerisation of propylene, which is even more exothermic and can cause a runaway	(f) Have adequate control at E112; have sufficient H_2O supply and adequate control to quench starting runaway
	MORE P	(7) K102 pressure control failure	No problem, probably even better conversion	(g) None
Less	LESS PROPYLENE FEED FLOW	(7) As for (5)	Higher residence times in reactor, higher conversions	(g) None
	LESS T	(8) As for (6)	Lower temperature in the reactor, lower conversions.	(h) As for (f)
	LESS P	(9) As for (7)	Conversion drops, even larger recycle	(i) Institute regular inspection and maintenance of K102

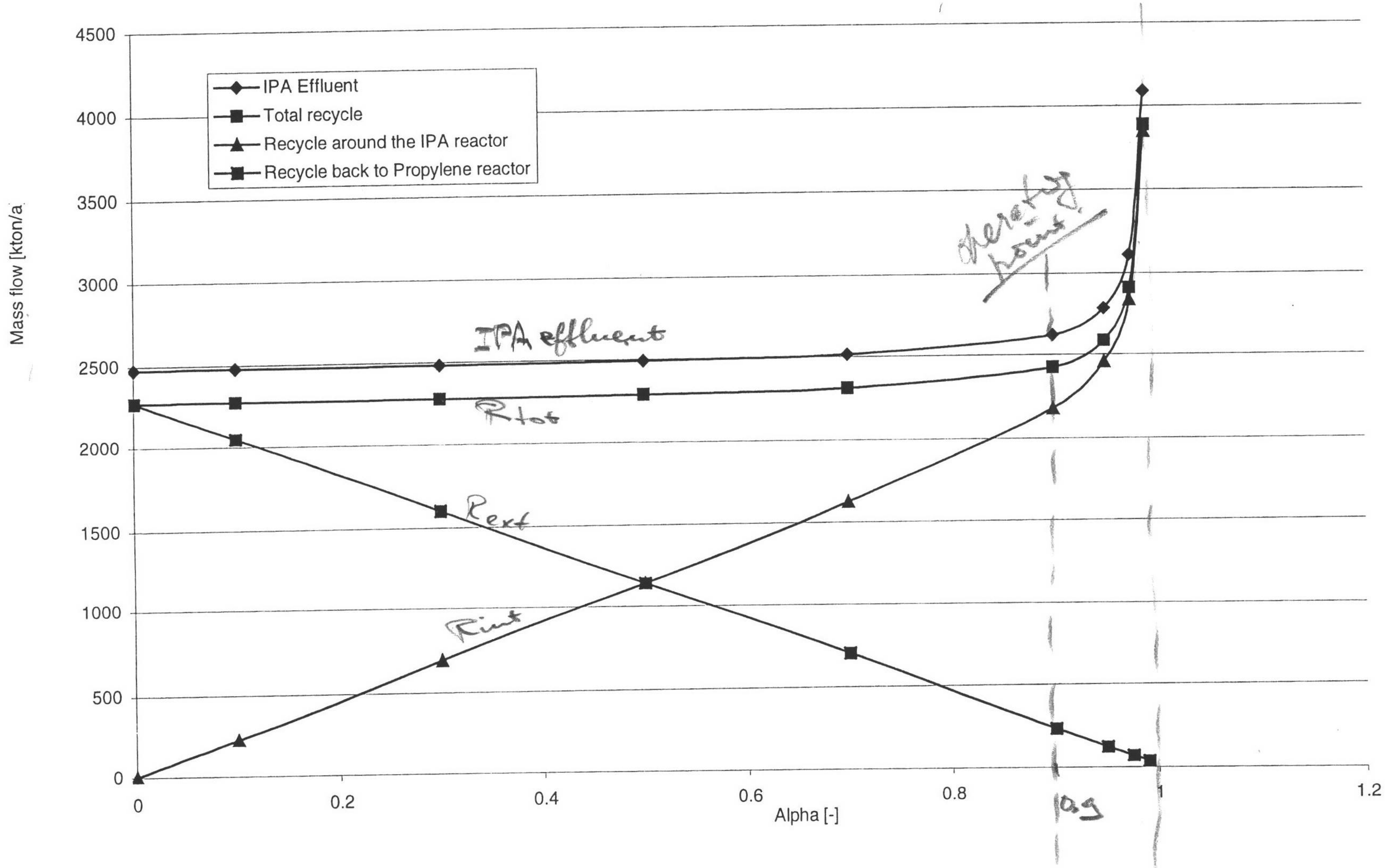
Fire and Explosion Index

Location	Date		
Malaysia	*December 20th, 1999*		
Plant	Process units	Evaluated by	Reviewed by
IPA from propylene	*propylene and IPA reactors, deethanizer*	*CPD3242*	

Materials and process

Materials in process unit *hydrogen, propylene, propane, light ends, impurities*	
State of operation [ ] Start-up [ ] Shut-down [x] Normal operation	Basic material(s) for material factor *hydrogen, propylene, propane*

1. General process hazards	Factorrange	Propylene Reactor Penalty used	IPA Reactor Penalty used	C101 Column Penalty used
Material factor		21	21	21
Base factor	1.00	1.00	1.00	1.00
A. Exothermic chemical reactions	0.30 to 1.25	-	*0.30*	-
B. Endothermic processes	0.20 to 0.40	*0.20*	-	-
C. Material handling & transfer	0.25 to 1.05	-	-	-
D. Enclosed or indoor process units	0.25 to 0.90	-	-	-
E. Access	0.20 to 0.35	-	-	-
F. Drainage and spill control	0.25 to 0.50	-	-	-
General process hazards factor (F1)		*1.20*	*1.30*	1.00
2. Special process hazards				
Base factor		1.00	1.00	1.00
A. Toxic material(s)	0.20 to 0.80	*0.20*	*0.20*	*0.20*
B. Sub-atmospheric pressure (<0.67 bar)	0.50	*0.50*	*0.87*	-
C. Operation in or near flammable range [ ] Inerted [x] Not inerted		-	-	-
1. Tanks farms storage flammable liquids	0.50	-	-	-
2. Process upset or purge failure	0.30	-	*0.30*	*0.30*
3. Always in flammable range	0.80	-	-	-
D. Dust explosion	0.25 to 2.00	-	-	-
E. Pressure Operating pressure	0.15 to 1.10	-	*0.87*	*0.88*
F. Low temperature	0.20 to 0.30	-	-	*0.20*
G. Quantity of flammable material: W: lbs, Hc: 2500 BTU/lb				
1. Liquids, gases and reactive materials in process	0.15 to 3.00	*0.10*	*0.10*	*0.10*
2. Liquids or gases in storage	0.10 to 1.60	-	-	-
3. Combustible solids in storage, dust in process	0.10 to 4.00	-	-	-
H. Corrosion and erosion	0.10 to 1.50	*0.10*	*0.10*	*0.10*
I. Leakage-Joints and packing	0.10 to 1.50	*0.10*	*0.30*	*0.30*
J. Use of fired heaters	0.10 to 1.00	*0.25*	*0.10*	*0.14*
K. Hot oil heat exchange system	0.15 to 1.15	-	-	-
L. Rotating equipment	0.50	*0.50*	*0.50*	*0.50*
Special process hazards factor (F2)		*2.75*	*3.47*	*3.72*
Unit hazard factor (F1*F2=F3)	1.00 to 8.00	*3.30*	*4.51*	*3.72*
Fire and explosion index (F3*MF=F&EI)	0 to 200	***69.30***	***94.76***	***78.20***


IPA Effluent
Total recycle
Recycle around the IPA reactor
Recycle back to Propylene reactor
Mass flow [kton/a
Alpha [-]
4500
4000
3500
3000
2500
2000
1500
1000
500
0
0.2
0.4
0.6
0.8
1
1.2
IPA effluent
Rtot
Rext
Rint
0.9

Economic evaluation

Assumptions:

- Due to company policy no installed spares for equipment will be used.
- The make up of cyclohexane, is considered to be miscellaneous material. ≈ Raw Mat!

Currency

Dfl/$	ROE	2.00
Dfl/pond		3.50
Dfl/DM		1.14
$/Pound		1.75 £
$/Dfl		0.50
$/DM		0.57
pound/DM		0.33

Inflation

%

Inflation correction from 1970	1.03	2.36
Inflation correction from 1992	1.03	1.23
Inflation correction from 1997	1.03	1.06

$x^{29} = 2.36$

$29 \ln x = \ln 2.36$

$\ln x = \frac{\ln 2.36}{29}$

$x = e^{\frac{\ln 2.36}{29}}$

Note that the inflation correction is based on 3% a year.

Table 9.1. Reactors

		Number [-]	Bare vessel [kPound]	Material factor [-]	Pressure factor [-]	Cost of the catalyst [DM/kg]	Reactor load [kg]
R101	Propylene reactor	3	30	1	1	20	31793
R102	IPA reactor	1	50	1	1.6	12	74183
Total							

Table 9. 2. Columns

	Bare vessel [kPound]	Material factor [-]	Pressure factor [-]	Cost of a plate [Pound]	Number of plates [-]	Total [kPound]
C101 Deethanizer	50	1	1.6	650	51	113
C201 Light ends column	80	1	1	700	45	112
C202 IPA CBM column	90	1	1	675	58	129
C203 IPA drying column	25	1	1	650	50	58
C204 Cyclohexane recovery column	14	1	1	250	35	23
C205 NPA recovery column	35	1	1	700	28	55
Total Columns						489

Table 9. 3. Absorber = Column!

		Bare vessel [kPound]	Material factor [-]	Pressure factor [-]	Packing cost [kPound/m3]	Volume packing [m3]	Total [kPound]
C102	Washing column	18	1	1.4	1.0	45.661	76
Total							76

Table 9. 4. Vessels

		Bare vessel [kPound]	Material factor [-]	Pressure factor [-]	Vessel costs [kPound]
V101	C101 accumulator	11	1	1.6	18
V102	HP Separator	10	1	1.6	16
V201	Light ends column phase sep.	22	1	1	22
V202	C202 accumulator	25	1	1	25
V203	Drying column phase sep.	25	1	1	25
V204	C204 accumulator	4	1	1	4
V205	C205 accumulator	10	1	1	10
Total	Vessels				120

Table 9. 5. Dryer

	Number [-]	Bare vessel [kPound]	Pressure factor [-]	Cost of the dryer [DM/kg]	Dryer load [kg]	Total [kPound]
D101 A/B/C Recycle dryer	2	25	1.2	2	50000	182
Total						182

Table 9. 6. Heat exchangers

Equip. number	Equipment name	Type	Area [m2]	Bare cost [kPound]	Type factor [-]	Pressure factor [-]	Total [kPound]
E101	R101 Feed-Effluent A	FH	92	12	1	1	12
E102	R101 Feed-Effluent B	FH	204	22	1	1	22
E103	R101 Feed-Effluent C	FH	217	22	1	1	22
E104	R101 Feed-Effluent D	FH	189	21	1	1	21
E105	R101 Effluent - Steam	FTS	62	10	0.8	1.1	9
E106	K101 Effluent	AIR	*				13
E107	C203 Reboiler	TS	49	9	1.3	1.3	15
E108	C101 Feed	FH	522	40	1	1.3	52
E109	C101 Condenser	FH	1065	82	1	1.3	107
E110	C101 Condenser Heat pump	FH	394	37	1	1.3	48
E111	C101 Reboiler	TS	139	17	1.3	1.3	29
E112	R102 Feed heater	FH	475	40	1	1.3	52
E113	C201 Reboiler	TS	945	80	1.3	1.3	135
E114	R102 Effluent cooler	AIR	*				223
E115	C102 Feed cooler A	FH	1706	135	1	1.3	176
E116	C102 Feed cooler B	FH	645	60	1	1.25	75
E201	V201 Feed cooler	FTS	185	20	0.8	1	16
E203	C202 Reboiler	TS	742	60	1.3	1.1	86
E204	C202 Condenser	AIR	*				439
E205	C205 reboiler	TS	496	40	1.3	1.1	57
E206	C205 Condenser	AIR	2390				406
E207	Recycle to C101	FTS	473	39	0.8	1	31
E208	C203 Condenser	AIR	*			1	6
E209	C204 Reboiler	TS	380	36	1.3	1	47
E210	C204 Condenser	AIR	*			1	43
E211	C201top feed cooler	FTS	634	60	0.8	1	48
E212	C204 Bottom cooler	FTS	11	3	0.8	1	2
E213	C203 Product cooler	FTS	350	34	0.8	1	27
Total	HE's						2218

* Calculated according to a rule of thumb:
0.001Dfl/MJ

Table 9. 7. Furnace

		driver power [kW]	Constant [kPound]	Index [-]	Total [kPound]
F101	R101 Feed furnace	14918	290	0.77	474
Total					474

Table 9. 8. Boiler

	Steam flow [kg/hr]	Constant [kPounds]	Index [-]	Total [kPounds]
Not mentioned in Flow Scheme	17700	30	0.8	75
Total				75

Table 9. 9. Pumps

		Actual power [kW]	Total cost [kPound]
P101	C101 Reflux pump	7.17	24310
P102	R102 Water feed pump	40.91	93185
P103	C102 Absorbent feed pump	96.40	101195
P201	C201 Bottom pump	27.35	91200
P202	V201 Inorganic phase pump	5.39	24310
P203	V201 Organic phase pump	0.00	18130
P204	C202 Bottom pump	24.87	62700
P205	C202 Reflux pump	15.26	46300
P206	C202 Top transfer pump	4.07	24310
P207	C202 Product transfer pump	2.57	21540
P208	C203 Bottom pump	2.18	21540
P209	V202 Inorganic phase pump	0.49	18130
P210	V202 Organic phase pump	9.13	32800
P211	C204 Bottom pump	0.44	18130
P212	C204 Reflux pump	0.49	18130
P213	C204 Transfer pump	0.12	18130
P214	C205 Bottom pump	2.28	21540
P215	C205 Reflux pump	3.89	24310
P216	C205 Transfer pump	2.09	21540
Total			701430

Table 9. 10. Compressors

		Number	Driver power [kW]	Constant [Pound]	Index [-]	Total [kPound]
K101	R101 vacuum compressor	1	2400	700	0.8	354
K102	R102 HP compressor	1	8042	500	0.8	666
K103	C101 heat pump compressor	1	4.488	500	0.8	417
K104	R102 recycle compressor	1	900	500	0.8	115
Total						1553

Table 9. 11. Turbine

		Number	driver power [kW]	Constant [Pound]	Index [-]	Total [kPound]
T101	Steam turbine	1	900	3200	0.5	96
Total						96

Table 9. 12. Storage tanks

		Number	Volume [m3]	Constant [kPound]	Index [-]	Total [kPound]
T201 A/B/C/D	Run down tank	4	224	1500	0.55	118
T202	Off spec tank	1	4902	1500	0.55	161
T203	On spec tank	1	60080	1500	0.55	637
T204	Sales tank	1	28008	1500	0.55	419
Total						1335

Table 9. 13. Physical plant cost

Equipment	Price [million $]
Reactors	1.96
Columns	1.18
Vessels	0.26
Dryer	0.32
Heat exchangers	4.77
Boiler	0.16
Furnaces	1.02
Pumps	0.37
Compressors (incl. K103)	3.34
Turbine	0.21
Storage Tanks	2.87
PPC [million $]	16.47
Lang factor direct costs	2.85
Lang factor indirect costs	1.45
Total Lang factor	4.13

Table 9. 14. Summary of fixed, working capital and the total investment

		Results [M$]
Fixed capital [million $]	C_F	68.07 ✓
Working capital	$0.15^* C_F$	10.21 ✓
Total investment	$1.15^* C_F$	78.28

BFW?

Table 9. 15. Raw materials

	Flow ton/a	Prices [$/ton]	Total [k$/a]
Propane <101>	146748	120.00	17610
Process Water <109>	89880	1.25	112
Total cost			17722 ✓

Table 9. 16. Utility costs

Utility	[unit]	Amount [unit/a]	Costs [k$/unit]	Total [k$/a]
Steam LP	kton	-3	7	-22
Steam MP	kton	206	9	1851
Steam HP	kton	1512	12	18144
Electricity	MWh	150822	0.04	6033
Fuel gas	kton	21	120	2532
Cooling Water	kton	61867	0.05	3093
BFW	kton	144	1.25	180
Total				31811

Table 9. 17. Waste

	Flow [ton/a]*	Treatment costs [$/ton]*	Total [k$/a]
Water <230>	176.4	300	53
Total			53

** ton is measured in weight of organic substances*

Table 9. 18. Summary of production costs

	Costs [M$/a]	Remarks
Variable costs		
Raw materials	17.72	
Miscellaneous materials	0.68	10% of maintenance
Utilities	31.81	
Waste	0.05	
Subtotal A	50.27	
Fixed costs		
Maintenance	6.81	10% of fixed capital
Operating labour	2.00	
Laboratory costs	0.50	25% of operating labour
Supervision	0.40	20% of operating labour
Plant overheads	2.00	100% of operating labour
Capital charges	10.21	15% of fixed capital
Insurance	0.68	1% of fixed capital
Local taxes	1.36	2% of fixed capital
Royalties	0.68	1% of fixed capital
Subtotal B	24.64	
Direct prod costs (A+B)	74.91	
Sales expense	18.73	25 % of direct prod. costs
General overheads		
Research and development		
Indirect costs (C)	18.73	
Annual prod costs (A+B+C)	93.63	
Production cost [$/ton]	535.65	

Table 9. 19. Products

		Flow [kton/a]	Prices [$/ton]	Total [k$/a]
IPA	<222>	174.80	550	96142
Self made fuel gas	<116>	15.96	120	1915
NPA	<207>	0.17	120	20
Total	<206>? LE	3.80		98057 → 456

Table 9. 20. Net cash flow

	Results [M$/a]
Gross income	98.06
Production costs	93.63
Net Cash Flow	4.42

Table 9. 21. DCFRR determination

Year	Net Annual Cash flow [million $]	DCFRR [-]	Net Cash flow [million $]	
0				
1	-39.1	1.00	-39.1	2 years
2	-39.1	0.99	-38.7	
3	4.4	0.98	4.3	
4	4.4	0.97	4.3	
5	4.4	0.96	4.2	
6	4.4	0.94	4.2	
7	4.4	0.93	4.1	
8	4.4	0.92	4.1	
9	4.4	0.91	4.0	
10	4.4	0.90	4.0	
11	4.4	0.89	4.0	
12	4.4	0.88	3.9	20 years.
13	4.4	0.87	3.9	
14	4.4	0.86	3.8	
15	4.4	0.85	3.8	
16	4.4	0.84	3.7	
17	4.4	0.83	3.7	
18	4.4	0.82	3.6	
19	4.4	0.82	3.6	
20	4.4	0.81	3.6	
21	4.4	0.80	3.5	
22	4.4	0.79	3.5	
Total		1.14%	0.0	

Table 9. 22. Results

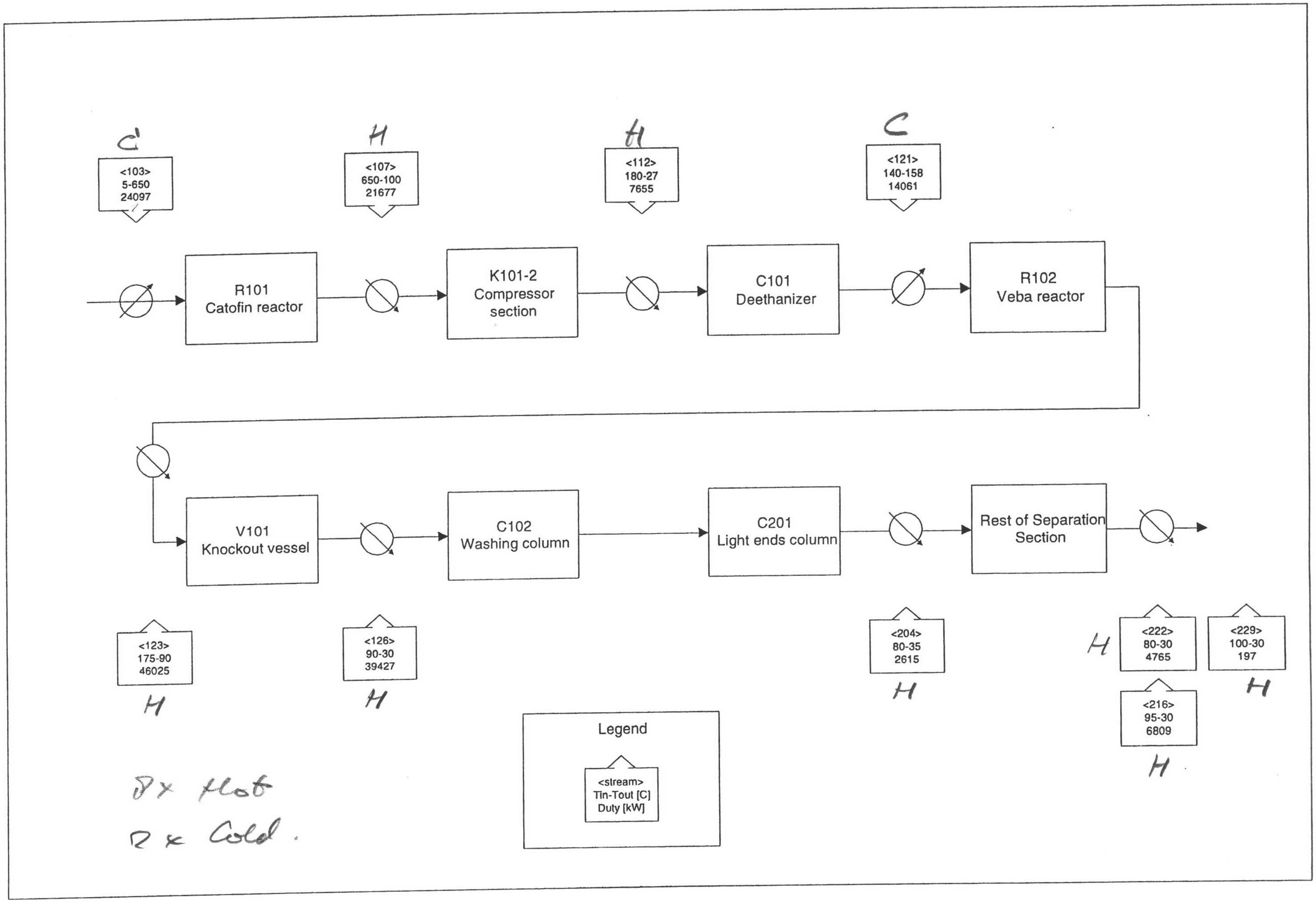
	Results
ROR	0.05
POT	19.46
DCFRR	1.14%

Table 9. 23. Margin

	Total [million $/a]
Products	98
Raw materials	18
Margin [million $/a]	80

Table 9. 24. Maximum allowable investment.

Year	Net Annual Cash flow [million $]	NPW (10%)	NPW [million $]	NPW [million $]
0				
1	0.0	1.00	0.0	-39.1
2	0.0	0.91	0.0	-39.1
3	80.3	0.83	66.4	27.3
4	80.3	0.75	60.4	87.6
5	80.3	0.68	54.9	142.5
6	80.3	0.62	49.9	192.4
7	80.3	0.56	45.3	237.7
8	80.3	0.51	41.2	278.9
9	80.3	0.47	37.5	316.4
10	80.3	0.42	34.1	350.5
11	80.3	0.39	31.0	381.5
12	80.3	0.35	28.2	409.6
13	80.3	0.32	25.6	435.2
14	80.3	0.29	23.3	458.5
15	80.3	0.26	21.2	479.6
16	80.3	0.24	19.2	498.9
17	80.3	0.22	17.5	516.3
18	80.3	0.20	15.9	532.2
19	80.3	0.18	14.4	546.7
20	80.3	0.16	13.1	559.8
21	80.3	0.15	11.9	571.8
22	80.3	0.14	10.9	582.6
Total				582.6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Deethanizer
C
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Knockout vessel
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C201
Light ends column
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6809
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100-30
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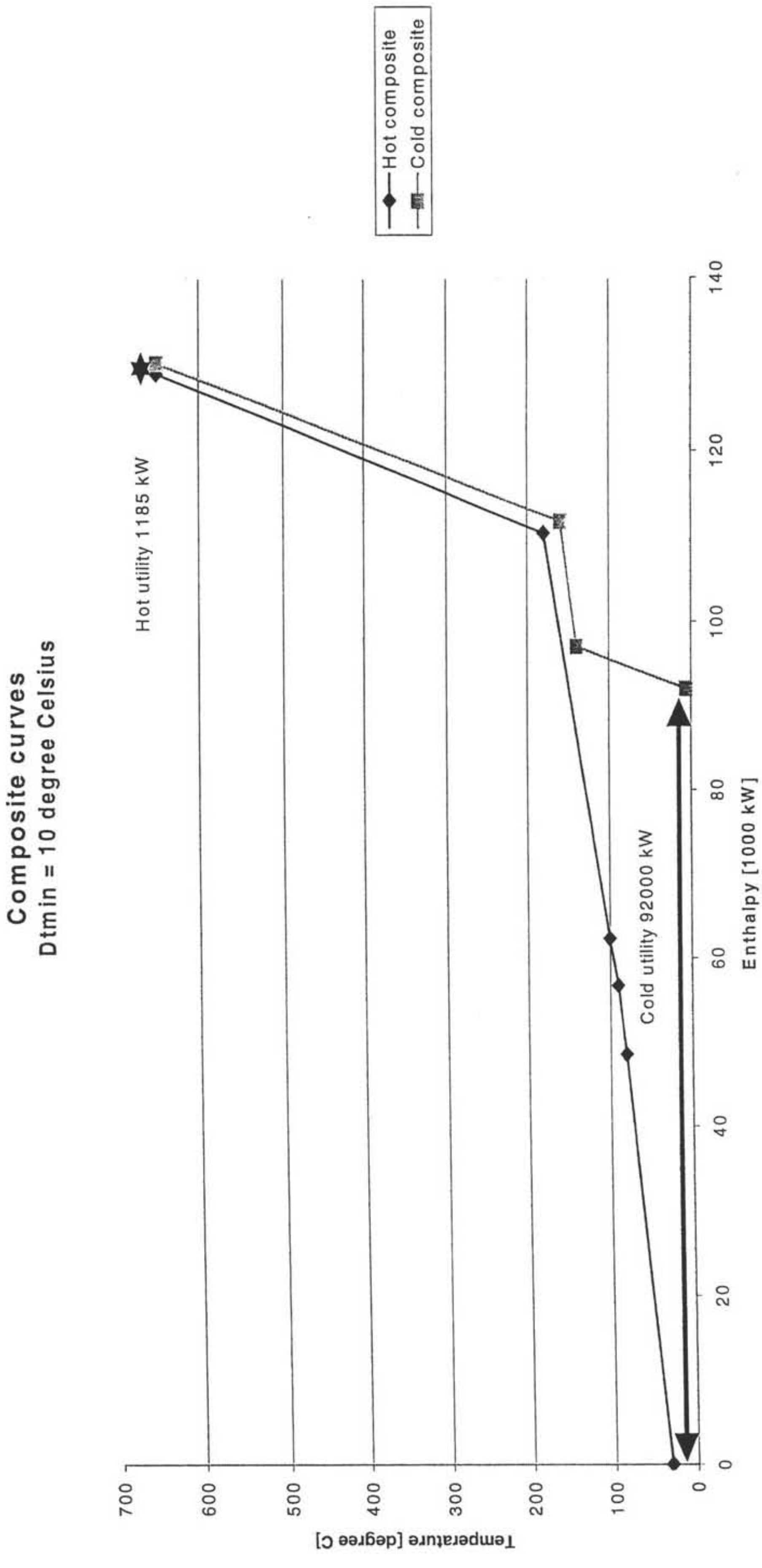
2x Cold.


Composite curves
Dtmin = 10 degree Celsius
Hot composite
Cold composite
Hot utility 1185 kW
Cold utility 92000 kW
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Enthalpy [1000 kW]
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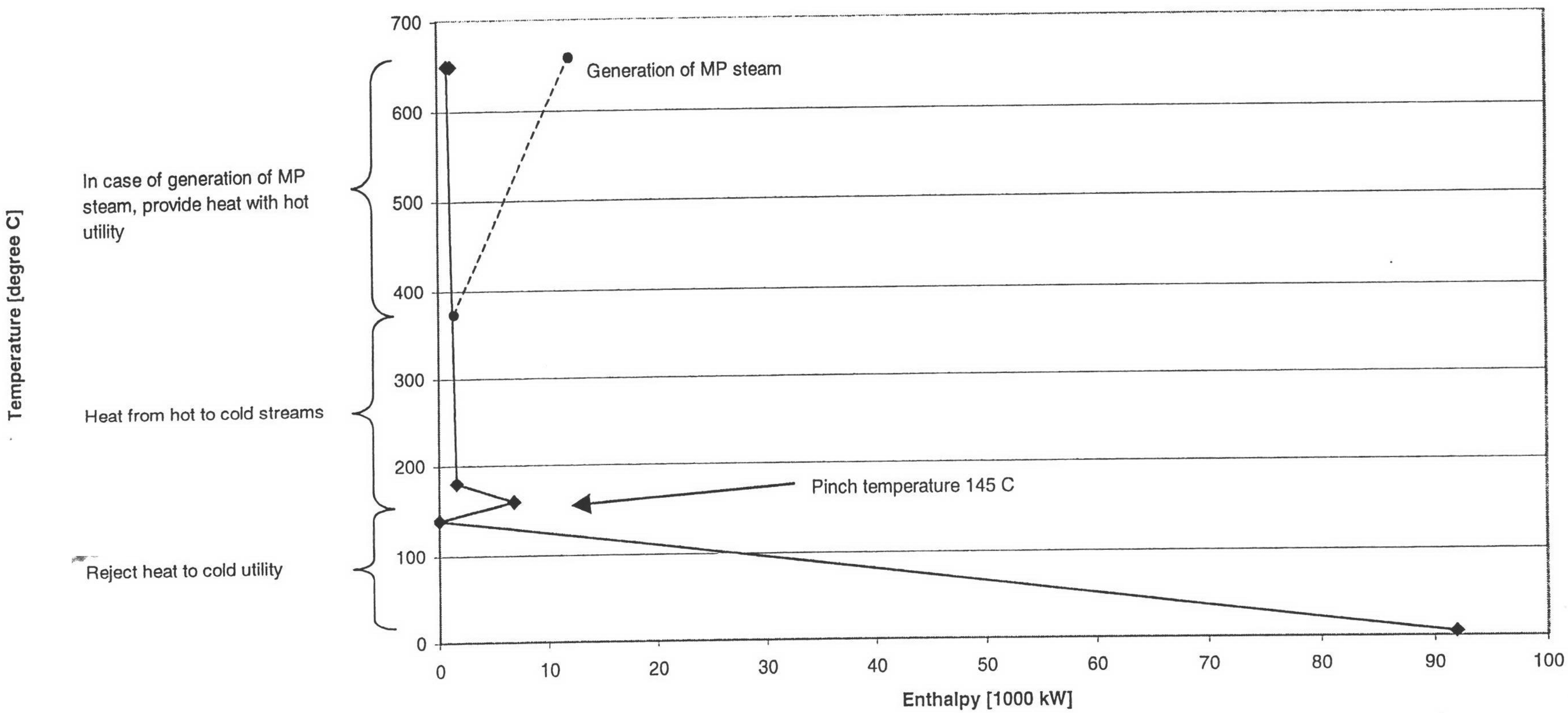
20
40
60
80
100
120
140

Grand composite curve



Available Utilities

a. Steam

Conditions	Steam Class		
	High pressure	Medium Pressure	Low Pressure
P [Bara]	18	12	4
T (superheated) [degree C]	225	200	150
T (condensation) [degree C]	207	188	144

b. Electricity

Electricity is assumed to be available without restrictions [electricity cost includes investment costs]

c. Water

	T (in) [degree C]	T (out) [degree C]	P [Bara]	cp [kJ/kg K]
Cooling water	25	40	3	4.18
Remarks: (*) Pressure in heat exchangers assumed to be 1 Bara				

d. Air

	T (in) [degree C]	T (max) [degree C]	P [Bara]
Air	30	40	1

e. Utility costs

Utility costs

	Units		LHV	Unit Costs [$/unit]	
Utility	Quant.	Energy	Energy Per Quant	Quant.	Energy
Steam LP	ton			7	
Steam MP	ton			9	
Steam HP	ton			12	
Electricity		kWh			0,04
Fuel gas	kg	MJ	40	0,12	0,003
	ton	GJ	40	120	3
Cooling Water	m3			0,05	
BFW/Process w.	m3			1,25	

Recommended PFS after the IPA reactor.

at

In the current design the knockout drum V102 operates as a flash vessel and pressure is reduced from 34 bara to 1 bara. The absorber column C102 also operates ~~on~~ atmospheric pressure. This was done because of modelling problems with Aspen 10, where the knockout drum V102 was finally modelled as a flash vessel. However, according to information provided by Shell[i], the pressure in these two vessels should be kept at 34 bara for an optimal separation. Knockout drum V102 then behaves as a high-pressure separator and the separation of IPA/water from propylene/propane is much better, in fact it becomes almost ideal. In the case of the high pressure all propylene and propane leaving the IPA reactor ~~come~~ over the top of the absorber column. And less water and ~~no~~ IPA leave the absorber column over the top. escape

or hardly any

or will be

This has consequences:

- The dryer in the overall recycle ~~becomes~~ smaller and loss of water is reduced.
- Compressor K104 becomes much smaller because the IPA recycle enters the compressor at 29 bara ~~and~~ exits at 34 bara. (instead of 1 to 34 bara).
- There is no propylene or propane loss.
- There is a strong driving force to transport the overall recycle back to the propylene reactor.

These consequences have a large impact on heat integration, utility consumption and the economic analysis.
No adjustments were made to the process stream summary in appendix 2. The data in the process stream summary has been taken from the Aspen models and have not been adjusted. However the use of utilities, design of units operations and the heat integration are based on other pressures and temperatures than are given by the Aspen models in order to give a more realistic design. The adjusted values for pressure and temperature are stated in the next table:

Table 12.1-1. Adjusted PFS values used for economics and design.

Stream number	Stream name	T in PSS	P in PSS	Adjusted T	Adjusted P
<XX>		[^{0}C]	[Bara]	[^{0}C]	[Bara]
<123>	TO E113	175	33	175	32.9
<124>	FROM E113	141	33	141.3	32.9
<125>	TO V102	90	33	90	32
<126>	FROM V102	50	1	90	30
<127>	TO E116	40	1	40	29.7
<128>	TO C102	30	1	30	29
<130>	TOP C102	35	1	35	29
<131>	TO K104	35	1	35	29
<132>	IPA RECYCLE	224	1	224	34
<133>	TO DRYER	35	1	35	29
<135>	BOTTOM HP SEPARATOR	50	1	90	30.2
<136>	BOTTOM ABSORBER	31	1	31	29.1

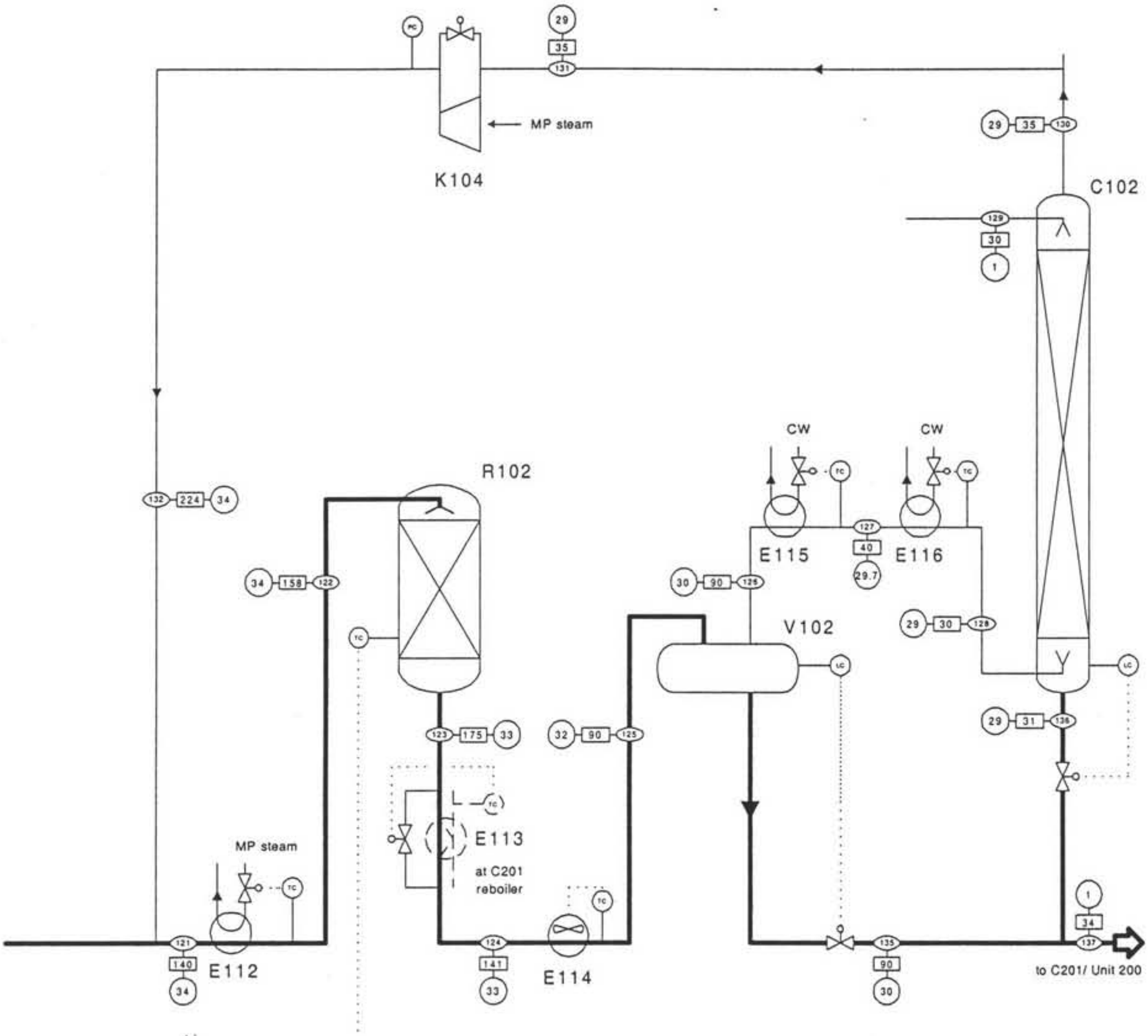
These adjusted values have consequences for the duties of several unit operations:

Table 12.1-2. Adjusted duties for equipment used for economics and design

Unit operations	Aspen duty	Adjusted duty
Code	[kW]	[kW]
E113	-20201	-20201
E114	-25824	-25824
E115	-13589	-36459
E116	0	-2968
K104	27890	900

The adjusted temperatures, pressures and duties were used for equipment design, heat integration and the economical analysis. A corrected part of the total Process Flow Scheme can be found in appendix 12.2 'Recommended Process Flow Scheme'.

[i] Park L. Morse, Synthetic ethanol and isopropanol, report No. 53, Process Economics program, Stanford Research Institute, Menlo Park, California, october 1969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C102
R102
CW
CW
E115
E116
V102
E113
at C201 reboiler
MP steam
E112
E114
to C201/ Unit 200

Catofin: Patent & literature searches

Author, Article / patent	Title – Description
Houdry, US XXX, 1943	Original Catofin patent [?]
Houdry, US 2419997, 1947	Original Catofin patent [?]
Hydrocarbon Proc., Sep. 1976	Catadiene process summary
Craig (Air Prod.), CEP, Feb. 1997, p. 62	The Catadiene process Subtitle: Improved economics, selectivity, and conversion are extending the life of this mature process → General process description.
Gussow (Air Prod.), Oil & Gas J., Dec. 8 1980, p. 96	Dehydrogenation links LPG to more octanes → Catofin process description, including yields and economics
Bhatt (Air Prod.), US 4581339, 1985	Catalytic dehydrogenation reactor cycle → Tackle problem of thermal gradient by successive oxidation and reduction with air and hydrogen in regeneration step
Spence (Air Prod.) , US 4560824, 1985	Heat balance control in catalytic dehydrogenation of isobutane → Mix n-butanes in feed to invoke more coke deposition, more exothermic energy available in regeneration
Feldman (ABB Lum.), US 5315056, 1992	Catalyst regeneration in a dehydrogenation process → Reduce needed regeneration air; pass regenerating air through two off-stream reactors in series, replace 1 reactor at a time; employ air for preheating dehydrogenation feed and/or steam generation
Gartside (ABB Lum.), US 5510557, 1994	Endothermic catalytic dehydrogenation process → Minimise activity differences over reactor axis by a) applying regeneration gas countercurrent to feed, b) varying catalyst activity in the reactor

	Oleflex	Catofin	STAR	FBD-4
	C3 C5	C4	C4	C4 !
Feed spec				
Pretreatment / heating	Desulphurised Demoisterised Deoxygenated Mixed with hydrogen recycle Heat exchanged	Vaporised with steam Furnace Heat exchanged	Feed vaporised, steam added Heat exchanged Furnace	Feed vaporised Heat exchanged
Reactor section				
Number of reactors	4 side by side	3 parallel	8 parallel	1
Kind of reactor	Radial flow	Fixed bed [?]	Parallel tubes with catalyst	Fluidised bed
Heat operation	Adiabatic	Adiabatic	Isothermal	Isothermal
Heating	Interstage heating [furnace]	Preheated reactor	Cocurrently heated by flue gas	Heat capacity from regen. cat
Continuous/Swing	Continuous	Swing	Swing	Continuous
Pressure	Superatmospheric	Vacuum [0.33-0.66 atm.]	1-20 atm	1.2-1.5 atm
Temperature	400-500 °C [Robin]	525 – 635 °C	510-620°C	525-590 °C
Conversion	40-50%	85-95%	45-55%	-
Selectivity	~90%	75-90%	85-95%	-
Catalyst				
Active compound	Platinum	Chromium oxide	Platinum	Chromium oxide
Activity	Uniform profile	Decreases with T		Uniform profile [?]
Catalyst life	> 1 yr	2 yr	1-2 yr	-
Regeneration unit				
Kind of equipment	CCR unit as for naphta cracker	Reactor on preheat		Regenerator f.b.
Utility	Oxygen needed [?]	Burnt with air	Burnt with air (1 hour)	Burnt with air
Separation from reactor	Wth hoppers	Valves		Filter
Cycle time	Continuous: 2-7 days	Every 7-15 min.	Every 7 hours	Continuous
Product treatment	Dried in molecular sieve	Dried by flashing water off	Dried by flashing water off (molecular sieve)	Dried in molecular sieve
Product separation				
By-product	Hydrogen recovery	Hydrogen recovery in PSA [?]	Same as Catofin (2 extra flash drums)	Same as Catofin
Columns	Light ends column PP splitter	Fuel gas / propene splitter		
Special equipment	Compressor after reactor section Hoppers	Product compressor Cold box [?] Air cooler	Same as Catofin	Same as Catofin Filters High eff. cyclone system Water scrubber Pneumatic conveying
Special utilities		Steam for feed	Steam for feed	
Wastes	Nitrogen consumption minimised			
Remarks	Avoid C4 accumulation in recycle			Cat. is very toxic!

idea?

Const. heat!

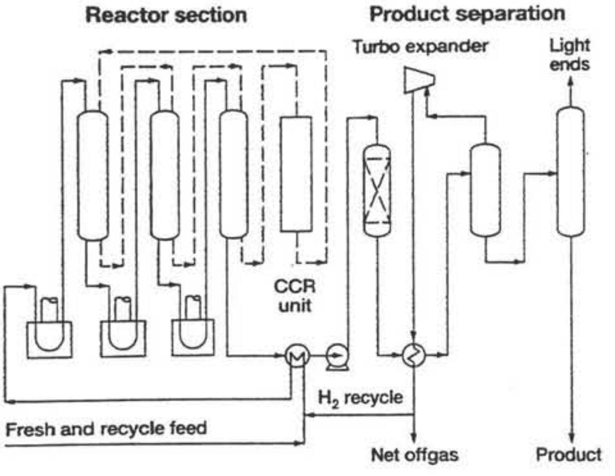

Fig. Dehydrogenation process for producing etherification feedstock from isobutane. proprietary Oleflex

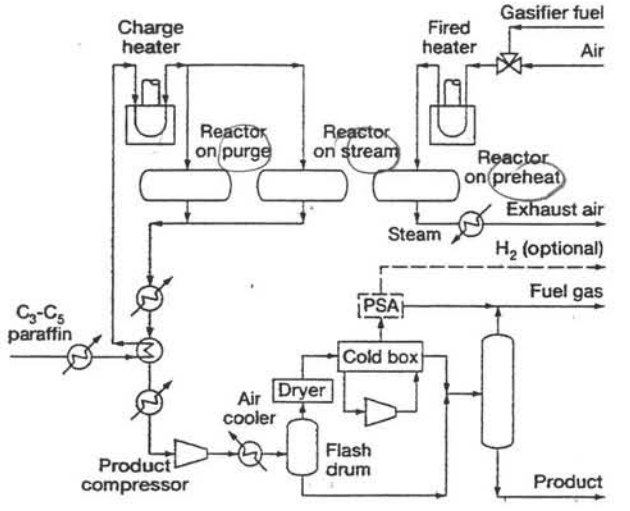

Fig. Proprietary Catofin dehydrogenation process for producing etherification feedstock.

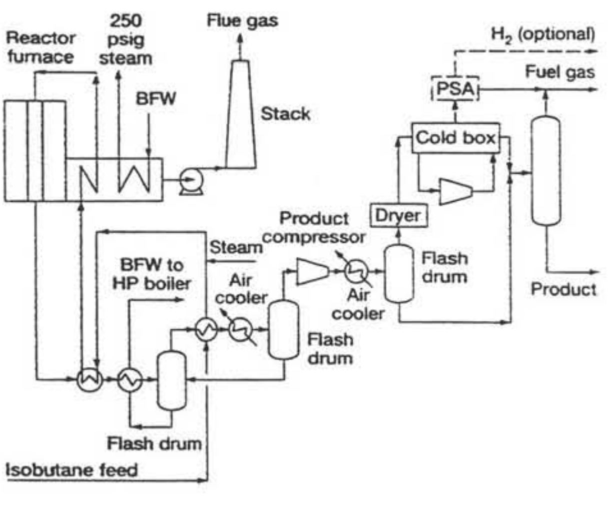

Fig. Proprietary STAR process.

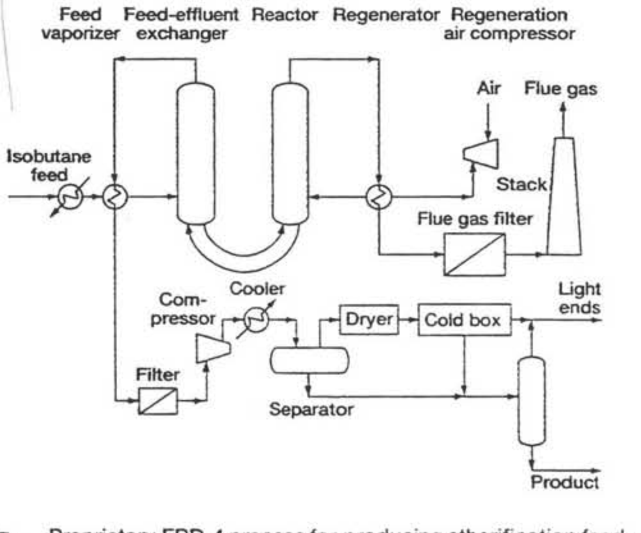

Fig. Proprietary FBD-4 process for producing etherification feedstock—initially developed in Russia.

Overview of IPA processes

	Veba	Texaco
Reactor section		
Number of reactors	3 Parallel	4 Parallel
Type of reactor	Fixed bed	Trickle flow
Phases	G/S	G/L/S
Cooling	Non	Internal quench
Continuous/Swing	Continuous	Continuous
Pressure [bar]	40,7	82,7
Temperature [°C]	218	138,9
Conversion per pass [%]	5,6	75,3
Selectivity [%]	97,5	96
Minimal propylene/propane ratio	0,8499	0,8
Molar ratio of water to propylene	0,94	13,8
Reactor space-time-yield [g IPA/(hr X Litre)]	103	108
Utilities per kg of IPA		
Steam HP/LP [kg]	4,58	6,1
Cooling water [m3]	0,41	0,23
Electricity [kWh]	0,0286	0,026
Natural gas [J]	5	0
Nitrogen [m3]	0,2	-

FIGURE

ISOPROPANOL BY DIRECT HYDRATION OF PROPYLENE-VAPOR PHASE