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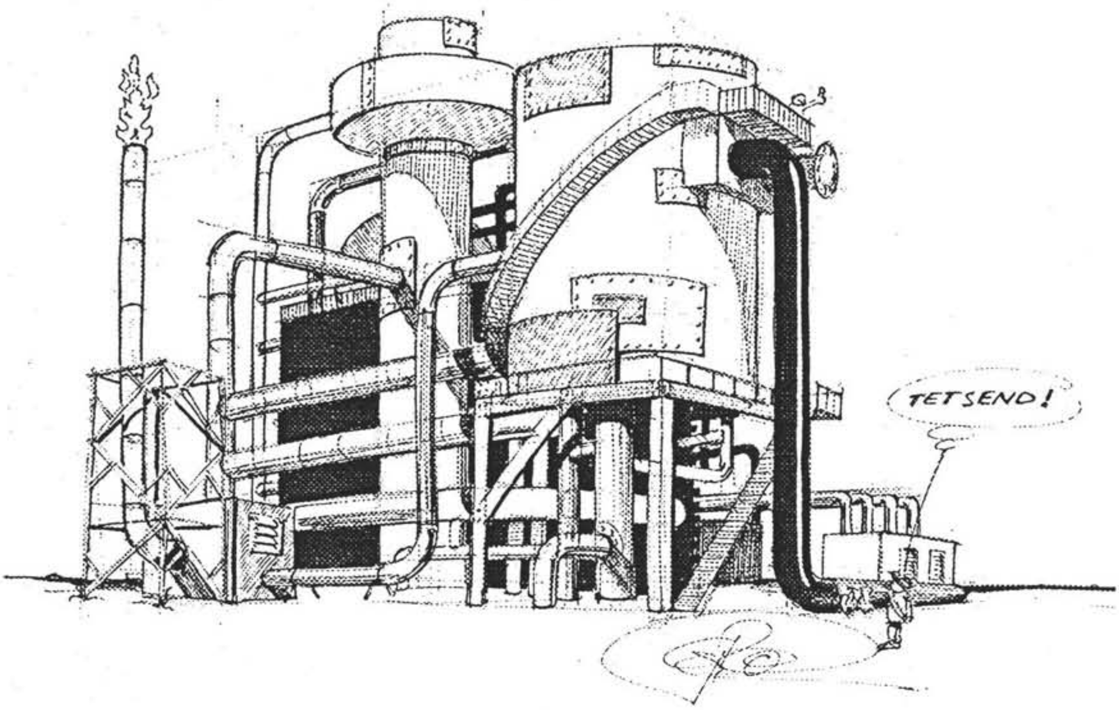

Conceptual Design for

The Manufacture of 200,000 t/a Fiber-grade MEG and 100,000 t/a Direct Sales Specification EO

CPD3244 team members
R.M. Boucke
F. Kleyn van Willigen
F.A. Sheldon
M.J. Sorgedrager

Date: January 1st 2000
Location: Stork Engineers and Contractors
Supervisors: ir. C.J. van Tiggelen
ir. J.A. Langerak

Preface

This design is made for the course Conceptual Process Design (ST493) at the subfaculty of chemical engineering, Technical University Delft, by 4 engineering students in the later stage of their studies. The findings of a 12 week team effort resulted in this design of an EO/MEG manufacturing plant, as presented in this document.

This design was made for, and under the guidance of, Stork Engineering and Contracting. Stork E&C gave us a much appreciated trial run on the work of an engineer in his own environment.
During this design we learned a lot about the way designing is done in the real world of engineers.
During the struggles of our team it became clear that design should be based both on technology and emotion, as well as basic engineering insight.

We wish to thank our supervisors Ir. C.J. van Tiggelen and Ir. J.A. Langerak for their effort in coaching during this design, as well as Stork Engineering and Contractors for the provided facilities. Thanks are also due to Ir. C.P. Luteijn for placing us with Stork.

For the artist's impression of a chemical factory on the cover page, we wish to thank P. Roeland.

Besides this extensive document, a CD-Rom containing all electronic information, such as Aspen simulations, Excel and Mathcad calculations, and Word documents is available.

Delft, December 1999

Summary

This conceptual design covers a design of a typical U.S. Gulf Coast plant for the production of 200,000 t/a fiber grade MEG and 100,000 t/a sales specification EO. The on stream factor of this plant is 91 % (8000 h per year). The global capacity for ethylene oxide (EO, or oxirane) currently stands at about 14.5 Mt/a, of which 2.5 Mt/a in Western Europe, and 5.2 Mt/a in North America. The plant design (equivalent with an EO capacity of 200,000 t/a) is middle class in size. The major 'players' on this market are Scientific Design, Shell, Nippon Shokubai and Dow. The EO and MEG demand is expected to grow to and past 2005.

A process is chosen in which ethylene is converted to ethylene oxide (EO), which is further used for the production of monoethylene glycol (MEG) by hydrolysis of the intermediate ethylene carbonate. An integrated scheme of a Shell based EO design and a Mitsubishi based MEG design is selected as the most economically viable design. A process based on carbonate route technology has been carried out only on a mini plant scale.

Most advantageous of using a carbonate route design (producing MEG via the intermediate ethylene carbonate) is a much higher selectivity (99%) to MEG compared with conventional hydrolysis routes (approximately 80 % selectivity) and the absence of evaporating large amounts of water (in a multi effect evaporator). if

The fixed capital calculated is 138 million USD. In comparison to other EO facilities as licensed by Scientific Design the fixed capital required by this design is lower. The annual raw material costs are 116 million USD. Annual utility costs are 12 million USD. Total annual production costs sum up to 175 million USD. Income from the sales of EO and MEG products are 246 million USD annually. This results in a payback time of 2.5 years, a sound economical plan Stability analysis shows that possible fluctuations in prices and production do not have lethal influence on the economic prospects of this design.

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15. APPENDICES

1. Introduction

This document presents the conceptual design of a plant for the production of 200,000 t/a fiber-grade monoethylene glycol (MEG) and 100,000 t/a direct sales specification ethylene oxide (EO). The principals are Stork Engineering and Contracting, and the subfaculty of Chemical Engineering, University of Technology, Delft.

The global capacity for ethylene oxide (EO, or oxirane) currently stands at about 14.5 Mt/a, with 2.5 Mt/a in Western Europe, and 5.2 Mt/a in North America[ref. 1]. The balance is accounted for largely by the Middle East and Asia. Globally, the demand for EO is growing strongly, due to increased demand for polyesters. Mono ethylene glycol (MEG, of 1,2-ethanediol) accounts for more than 70%[1] of the EO demand (split evenly between anti-freeze grade and fiber-grade), although in Europe the trend is towards higher value EO derivatives. EO is also used as fumigant and as a sterilizing agent for medical equipment, as well as feedstock for ethoxylate products. MEG is polymerized to polyesters and PET, and applied as freeze protection in cars, paints, and airport runways.

global? 1 ?

The total market for MEG has grown from 6.2 Mt/a in 1990 to 10.1 Mt/a in 1998. The growth in EO and MEG demand is expected to continue up to and past the year 2005[ref. 1,2].

The essentially single-train design described in this document has an equivalent capacity of 200,000 t/a EO, making it a middle-class plant as far as size is concerned. Table 1.1 reflects the current situation in North America for EO plants, compiled and updated from a number of sources. ? ref?

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The aim of this design is to produce 100,000 t/a sales specification EO and 200,000 t/a fiber-grade MEG, at the highest economically feasible selectivity's. Details of the product streams are given in chapter 3, basis of design. 70% of the operational costs of existing plants consist of ethylene feed. Therefore, a higher selectivity has a significant influence on total costs.

The production of MEG historically is accompanied by a substantial production of diethylene glycol (DEG), as well as higher glycols, and high energy consumption. It is the aim of this design to produce as little DEG and other by-products as possible with a low energy consumption and carbon dioxide production.

low

The design is based on a combination and integration of a Shell/CRI EO plant and a MEG process as patented by Mitsubishi. The EO plant is rather conventional, but the MEG process has not been applied in industry yet. Under the parameters of this project, this combination makes the most attractive design.

The EO production section consists of a multi-tube, water-cooled, plug-flow reactor, filled with a silver-based catalyst, over which approximately 10% conversion and 86% selectivity is achieved. This means 14% of the ethylene is combusted to CO_2 and H_2O. EO is scrubbed by a MEG/EC absorbent stream, which is split 50/50 for EO recovery (section 400) and MEG production (section 300). MEG production is a two-step process, the first a carbonation reaction, the second a hydrolysis reaction. Approximately

90% of the effluent of the carbonate reactor is recycled as absorbent. The remaining 10%, containing ethylene carbonate (EC) and MEG, is converted to 100% MEG in two hydrolysis reactors placed in series. Carbon dioxide that is liberated in this process is also recycled. Since little water is used in the process, the conventional multi-effect evaporators are unnecessary, and one dryer suffices. Fiber grade MEG is the product. Part of the unpurified MEG is recycled to the absorbent to preserve its ~1:1 MEG to EC ratio.

The design differs from conventional design in that the use of EC/MEG as absorbent instead of water substantially reduces production of undesired higher order ethylene glycols and the elimination of the multi-effect evaporators.

Table 1.1 North American EO producers, capacities, process types, and technology, for EO in 1992(ref 3, pg 942; 1).

Producer	Location	Capacity, 10^3 t/a	Process oxidant	Technology
BASF[1]	Geismar, La.	285	oxygen	Shell
Dow	Plaquemine, La.	270	oxygen	Dow
Dow	Alberta, Canada	285	oxygen	Dow
Eastman	Longview, Tex.	103	oxygen	Shell
Formosa Plastic Corp[2]	Point Comfort, Tex.	240	oxygen	unknown
Hoechst-Celanese	Clear Lake, Tex.	320	oxygen	Shell
Huntsman	Port Neches, Tex.	535[4]	unknown	unknown
Idesa[5]	Morelos, Mexico	160	unknown	unknown
Olin	Brandenburg, Ky.	50	oxygen	Shell
Oxy Petrochemicals	Bayport, Tex.	250	oxygen	Shell
PD Glycols	Beaumont, Tex.	320[4]	oxygen	Scientific Design
Pernex	Mexico	328	unknown	unknown
Quantum	Morris, Ill.	113	oxygen	Scientific Design
Shell	Geismar, La.	554[4] } 904	oxygen	Shell
Shell Canada[3]	Scotford, Canada	350	unknown	unkown
Sun Refining	Claymont, Del.	100[4]	oxygen	Shell
Texaco	Port Neches, Tex.	332	oxygen	Scientific Design
Tuntex[5]	Altamira, Mexico	400	unknown	unknown
Union Carbide	Seadrift, Tex.	349	air	Union Carbide
Union Carbide	Taft, La.	368	air, oxygen	Union Carbide
Union Carbide	Prentiss, Canada	480[4]	unknown	unknown
	Total	*6192*		

1. Being upgraded to ~420 t/a, ready in 2000
2. Being upgraded to ~270 t/a, ready in 2001
3. Joint venture with Mitsubishi Corp., due on stream mid-2000
4. multiple plants.
5. planned, post 2002.

2. Process Options and Selection

This chapter describes the selection of an ethylene oxide production process, the selection of a monoethylene glycol process, and the final combination of these two. Some historical background and chemical issues of each of the processes is also presented.

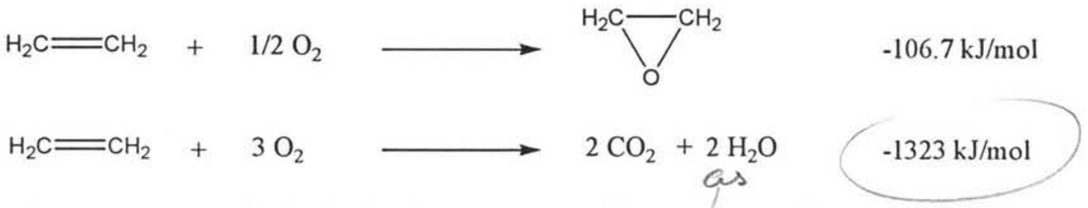
2.1. Ethylene Oxide Process

Ethylene oxide has been commercially produced by two basic routes: the ethylene chlorohydrin process and direct oxidation (DOEO). Other, commercially not viable, processes include: the arsenic-catalyzed liquid phase process, the thallium-catalyzed epoxidation process, the Lummus hypochlorite process, the liquid-phase epoxidation with hydroperoxides, the electrochemical process, the cyclic direct oxidation process, the fluid-bed direct oxidation process, and biological processes [ref 3, pg 939-970]. None of these processes have passed the pilot-plant phase.

The chlorohydrin process (first employed by BASF in Germany during World War I) is based on the production of ethylene oxide from ethylene chlorohydrin by dehydrochlorination using either sodium or calcium hydroxide. Calcium chloride, dichloroethane, bis-(2-chloroethyl)-ether, and acetaldehyde are also produced. The chlorohydrin process, although it appears simpler, is no longer commercially operated, largely due to higher capital costs and major environmental problems.

NaOH.

The direct oxidation (air-based first by Union Carbide Co., 1937, oxygen-based first by Shell Oil Co., 1958) of ethylene proceeds by the following reactions:



The first reaction results in the desired ethylene oxide, the second is the complete oxidation of ethylene, and is undesirable. The reactions are catalyzed by a silver on alpha-alumina catalyst (see appendix 5 for a comparison between various catalysts). The lifetime of such a catalyst is approximately three years with an average selectivity of 86%. Small amounts of reaction modifiers/promoters, such as methyl chloride, are also used – this enhances catalyst life and reduces the chance of hotspots. The large amount of heat produced, combined with a solid catalyst, would make a fluidized bed reactor appear to be the best choice, due to its capability for heat transfer. However, abrasion of the catalyst reaches unacceptable levels in fluidized beds. Therefore, only fixed-bed, multi-tubular reactors are employed in commercial processes.
Oxygen may be supplied either as air, or as a pure feed. Air based plants, eliminating the need for an air-purification plant, are presently economically feasible for a maximum production of ~25,000 t/a EO [ref 4]. Larger plants usually employ the oxygen based direct oxidation process, and air based plants are no longer built.

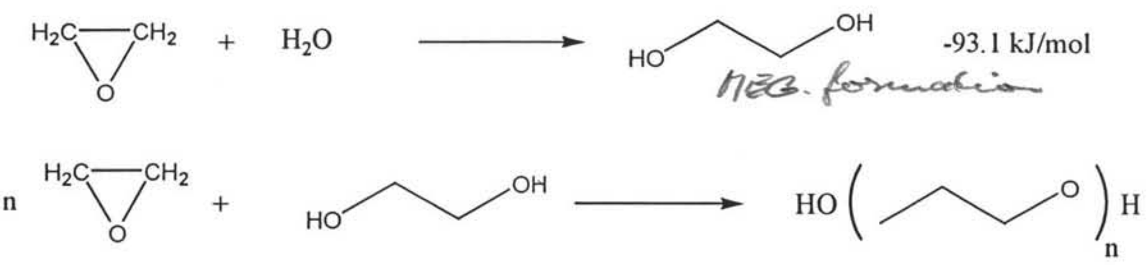
The conventional direct oxidation processes all employ water as absorbent for the recovery of ethylene oxide. There are, however, two drawbacks to using water. One problem is the by-production of (poly) ethylene glycols at the time of recovery. As much as 3-15% of all ethylene oxide can be lost. The ethylene glycol thus produced is difficult to purify to fiber-grade standard due to the high amounts of aldehydes and organic acids contained in this stream. This ethylene glycol, however, can be sold at anti-freeze grade. The second major drawback is the energy loss. A large amount of water is employed as absorbent. The absorption of ethylene oxide preferably proceeds at a relatively low temperature (5-40 °C), while the stripping step requires higher temperatures such as 85 to 130 °C. It is difficult to recover the large amount of energy from this water economically [ref. 5]. The use of alkyl carbonates as absorbent is an attractive alternative; however, the absorption of water, as well as higher percentages of ethylene scrubbed, makes its use in plants producing solely EO more difficult and uneconomic.

Major licensors for the EO oxygen process are Scientific Design, Shell, and Nippon Shokubai [ref. 3, pg 924]. Over 70% of present world capacity is based on their processes. Union Carbide Corp. and Dow Chemical, now merged, use their own processes [ref. 3, pg 939-940]. Dow does not license its technology. About 94% of U.S. EO capacity is located on the Gulf Coast, near secure and plentiful ethylene supplies.

These processes are all very similar (silver based catalyst, multi-tubular reactor, water-based absorption), with only minor variations in heat integration and reactor design. However, catalyst enhancement by means of doping is a very active art, with numerous new designs being patented (see appendix 5 for a comparison between different EO catalysts).
Based on such commercial factors as the number of plants already built/licensed, number of plants currently under construction/planned, catalyst quality (selectivity, activity, lifetime, costs), and reactor design, a Shell/CRI design has been selected as the EO process for our conceptual design. Shell/CRI are the market leaders in licensed EO plants. This choice also has a practical nature - many of the figures known to us apply to the Shell/CRI design [ref. 6]. Let it again be stressed that at the current time, based on the available literature, little variation between direct oxidation oxygen-based EO designs can be found.

2.2. Monoethylene Glycol Process

Globally, some 70% of EO is converted to ethylene glycols [ref. 1]. Currently the preferred method of ethylene glycol production is uncatalyzed thermal hydrolysis of ethylene oxide. Ethylene oxide absorbed in water is heated to ca. 200 ^{0}C, whereby the ethylene oxide is converted into ethylene glycol. Polyethylene glycols are also produced, but with lower yields. The yields of polyethylene glycols can be minimized if an excess of water is used; a 20 fold molar excess is usually employed. In practice almost 90% of the ethylene oxide can be converted to monoethylene glycol, the remaining 10% reacts to form higher monologues [ref. 7, pg 105).



-93.1 kJ/mol

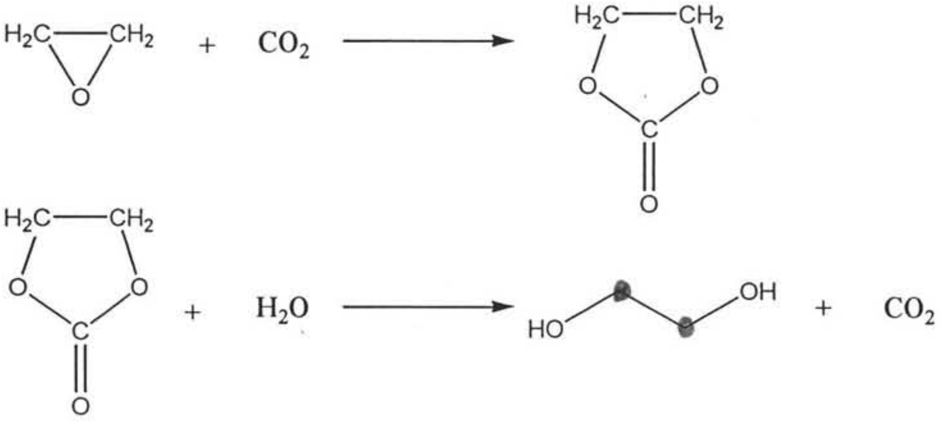
MEG. formation

n=1,2,3...
DEG formation: -1.89 kJ/mol

After leaving the reactor, the product mixture is purified by passing it through successive distillation columns with decreasing pressures. During this process, the large amounts of water added to enhance selectivity must be evaporated in multi-effect evaporators.
A literature study of the monoethylene glycol process resulted in two alternatives for the conventional process: a catalytic process and a carbonate process.

A lot of research is done on a catalytic hydrolysis process in order to optimize selectivity, since higher selectivity for ethylene oxide hydrolysis automatically reduces the excess of water required. The higher selectivity could however not be maintained for prolonged periods of time. Furthermore, the catalyst needs to be separated and fed back into the reactor in case of heterogeneous catalysis or replaced in case of homogeneous catalysis.

In the carbonate route process ethylene glycol is produced from ethylene oxide (eqn 2.5), with ethylene carbonate as an intermediate. The ethylene oxide is converted with carbon dioxide to ethylene carbonate. This is subjected to hydrolysis to produce ethylene glycol.[ref. 8]. Both reaction steps are catalytic. An anion-exchange resin in the halogen form can be used for this purpose. Disclosed as being especially suitable is a chloride form anion exchange resin such as DOWEX MSA-1, which is an anion-exchange resin containing benzyl-trimethyl ammonium groups as electropositive centers[ref. 8]. A typical lifetime of such a catalyst is approximately one year. Also homogeneous catalysts, quaternary ammonium salts, can be applied but because of separation problems heterogeneous catalysis has practical advantages.



-60.8 kJ/mol

-32.2 kJ/mol

The two most important selection criteria for the MEG part of the design are selectivity and the H_2O/EO ratio used for hydrolysis. The selectivity is important because the ethylene oxide efficiency and thus the ethylene efficiency (ethylene costs are a major part of the OPEX) is an important economic factor, as well as an environmental issue. The H_2O/EO ratio is important because of the large effect it has on the energy requirements of the separation section (separation of water and ethylene glycol with a multi effect evaporator).

of .

The carbonate route process reaches a conversion of nearly 100% and a selectivity of 99% or higher, only diethylene glycol (selectivity less than 1%) is produced. The H_2O/EO ratio is much lower than the other two considered processes, only a one to two-fold 1-2 molar excess is added, as compared to a 20-fold excess for the conventional thermal process and a 5 to 10–fold excess for the catalytic one step hydrolysis. Therefore much less energy is needed in the MEG purification this process. A~~n~~ comparison between the options is shown in table 2.1. The carbonate route process scores best on these two topics and should therefore be considered the process to be used for the further design. The process is patented by Mitsubishi ref. 8. One (major) drawback of this process, however, is the fact that it has not been implemented in the industry yet. The process has been carried out, with success, on a mini-plant scale ref. 8.

Table 2.1 MEG process comparison

Process	Conversion[2]	Selectivity	H_2O : EO molar ratio	Number of main components	Implementation in industry
Conventional Thermal Comment:	nearly 100%	88 %	20	5 EO, MEG, DEG TEG, H_2O	Widely employed
Conventional catalytic[1] Comment:	nearly 100%	98 %	10	4 EO, MEG,DEG H_2O	not employed selectivity decreases rapidly
Carbonate route process Comment:	nearly 100%	99 %	1.2	6 EO, MEG, DEG H_2O, CO_2, EC	not employed yet

Notes:

1. catalyst dependant. Best found values are given here.
2. Single pass conversion

Abbreviations:

EO: Ethylene oxide
EC: Ethylene carbonate
MEG: Monoethylene glycol
DEG: Diethylene glycol
TEG: Triethylene glycol

2.3. Process Concept Selection

Choosing a licensor's complete design - for example, an integrated Shell/CRI EO/MEG facility - may seem like the best route to follow. After all, companies such as Shell/CRI, Scientific Design, and Mitsubishi have each been involved in building tens of EO, MEG, or integrated EO/MEG plants, and therefore have a wealth of experience in designing, building, startup, and operation of such plants, and they can guarantee the workability of a plant. This makes the licensing of such a design very attractive to potential customers. The conceptual process design, however, is theoretically not hindered by the constraints of 'proven design'. This is not to say that the value of 'proven technology' is ignored, only that our design is not required to be a complete license from one licensor. Furthermore, the guidelines as laid out in the basis of design (chapter 3) must be followed. On the other hand, a design concept must be realistic and commercially viable. Therefore, a balance must be found between state of the art technology, novel technology, proven design/technology, and commercial interests.
Our selection process, based on different variations and combinations of a Shell/CRI EO reactor and a Mitsubishi MEG design, is illustrated in figure 2.1.

Scheme 1 reflects the most conservative design. A complete Shell/CRI ethylene oxide process, with water as absorbent, is followed by a complete Mitsubishi process. This involves splitting of the pure EO stream leaving the Shell/CRI plant in 1/2 for sales and 1/2 for MEG production. Naturally, extensive heat integration between the two processes is possible in this design.
However, the Mitsubishi process is based on the unrefined reactor effluent of an EO process. The absorbing and stripping of EO in the Shell process is thus followed by reabsorbtion in the MEG plant, albeit in a different absorbent. The absorption is therefore redundant, although this does have the advantage that catalyst pollution, if it is a factor, can be reduced. The unwanted production of low quality ethylene glycols due to the use of water as absorbent is not avoided in this scheme. The Halcon Company has patented such a design[ref. 9]. However, several more elegant schemes are possible.

Scheme 2 involves splitting the EO reactor effluent gas into approximately 1/2 for EO production and 1/2 for MEG production. The EO production train is 'standard' Shell/CRI design, and the MEG production is standard Mitsubishi design. However, this design contains 2 recycles to the EO reactor, since 2 absorbers, with 2 different absorbents, are now operating parallel. This makes control of the reactor feed, critical because of explosive and flammability dangers, very difficult.
Again, the use of water as an absorbent has the disadvantage that low quality ethylene glycols are produced, but a variation of this scheme involving propylene carbonate in the Shell/CRI EO absorption train[ref. 10] is a possibility. Propylene carbonate may be better suited for the requirements than an EC/MEG mixture, but it cannot be used as feed to the MEG plant – the separation of propylene carbonate and ethylene carbonate would be necessary. Besides, feeding propylene carbonate into an ethylene carbonate production unit, or introducing it solely as absorbent for an EO section, is not very logical. Absorption of EO in EC/MEG is described in a patent to the Dow Chemical Co[ref. 11].

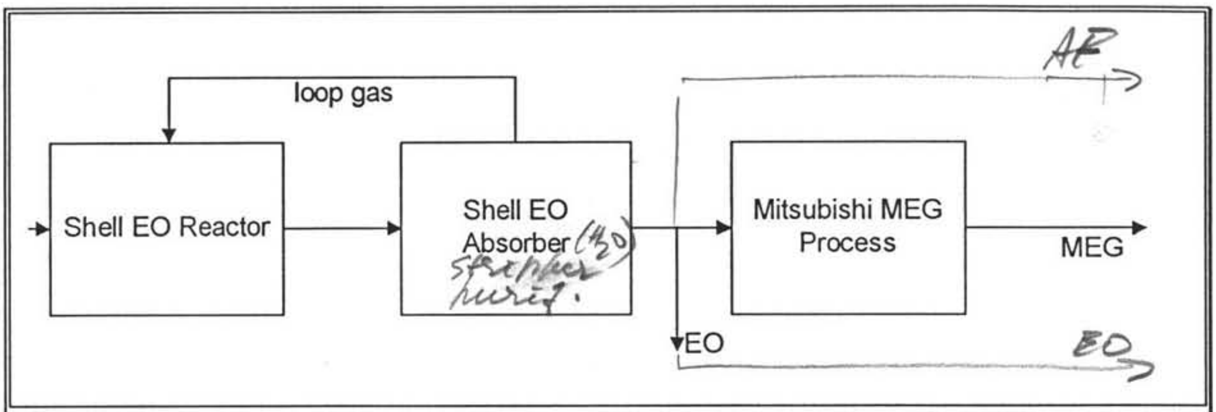
The third scheme is the most preferable. In this design, the whole effluent of the Shell/CRI EO reactor is fed directly to the Mitsubishi absorber, as intended in the

Mitsubishi design. The top stream of the absorber is recycled to the EO reactor, the bottom stream is split into a feed for the MEG process, and a stream which is stripped and processed to sales specification ethylene oxide. One absorber, with EC/MEG absorbent, is used as basis for both trains. The use of solely EC as absorbent is also described in patents[ref 10], but the major drawback of this is the high melting point of EC. By using a mixture of EC and MEG this problem is avoided. A 1/1 mass based mixture at standard pressure is still liquid[ref 8].
The heating and cooling of the large amounts of water as in the classical processes is not necessary in this design - much less absorbent, with a lower heat capacity (water: 4.18 J/g K, MEG/EC: 2.01 J/g K) so that less heat per unit is required, may be used. This design also eliminates a high-content ethylene oxide stream, the feed to the MEG plant, which is very advantageous to process safety. Furthermore, equipment costs are lower, since only one absorber is required. The production of anti-freeze grade MEG is eliminated, vs. a substantial loss in the other schemes.
Scheme 3 takes full advantage of the experience of Shell/CRI in the EO design and the advantages of the Mitsubishi MEG design, while minimizing costs and wastes, enhancing operational safety, and maximizing reliability.

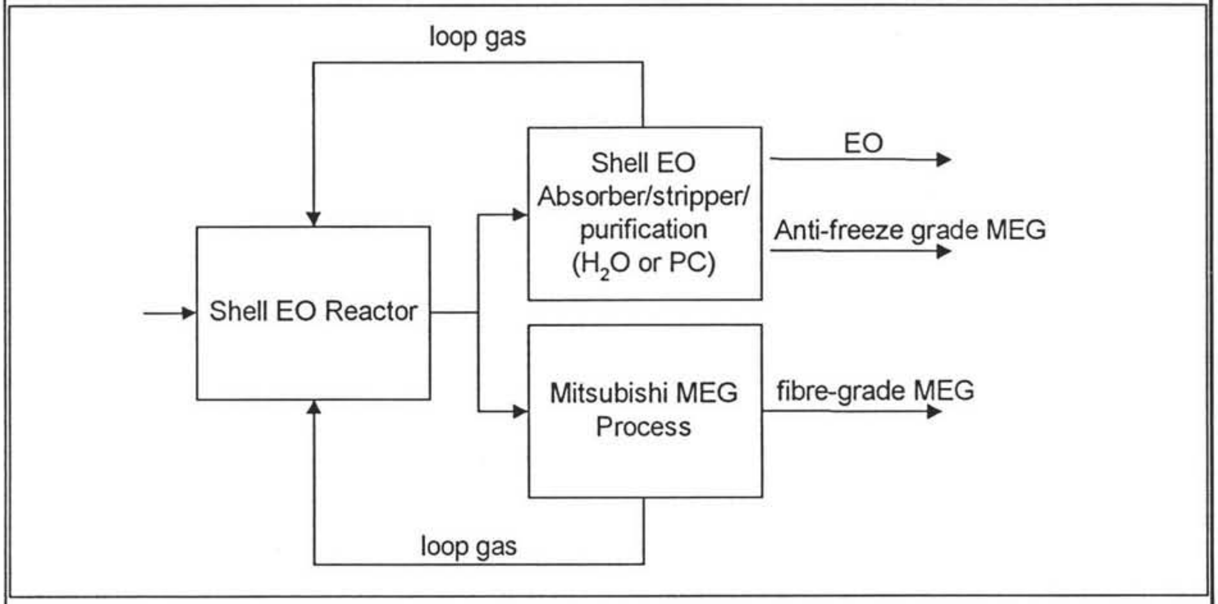
In table 2.2 a comparison of the different process schemes is presented. This takes into account that the EO part is based on Shell/CRI design and the MEG part is based on Mitsubishi design.

Table 2.2. Comparison of process options

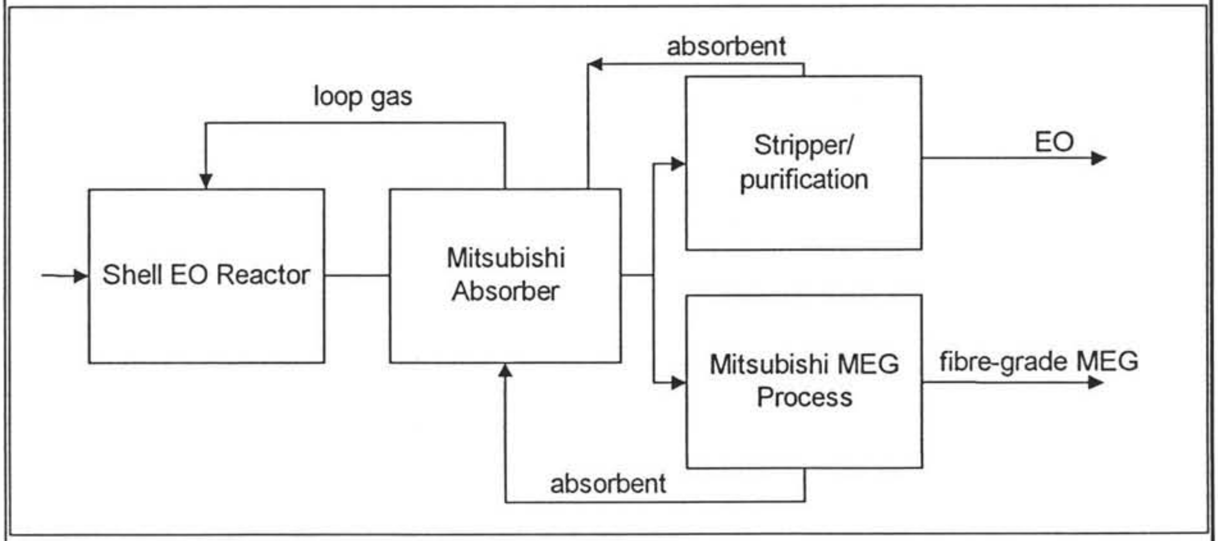
Process	Complete license benefits	High Energy efficiency	High Selectivity	Low Capital cost	Little anti-freeze MEG	Low absorbent amounts
Scheme 1	++	-	-	-	--	-
Scheme 2	+	-	+	+	-	+
Scheme 3	-	++	++	++	++	++



Scheme 1, Shell & Mitsubishi in series



Scheme 2, Shell & Mitsubishi parallel



Scheme 3, Shell & Mitsubishi integrated

Figure 2.1, Process selection

3. Basis of Design

The capacity of the plant is based on a production of 200,000 t/a of fiber-grade MEG and 100,000 t/a of sales specification EO. The plant is based on a typical U.S. Gulf Coast location. The duration of operation per year is set at 8000 hours/year. These design demands are set by Stork E&C. Econ. Plant Life?

The process is to be based on the best available technology. EO production must be based on direct oxidation with oxygen over a multi-tubular catalytic reactor, and integrated with a MEG plant based on the hydrolysis of EO. Deliverables include a utility summary, PFD's with heat and material balances, equipment lists, and datasheets of main equipment. An Aspen model and cost price calculations are also desired. A hazard- and fire and explosion analysis is also included. Product specifications are included in table 3.1

↳ required.

Appendix 1 shows a simplified scheme of the plant. This scheme defines what's inside the battery limit. All in- and outgoing streams pass the battery limit. The feed stock (ethylene, oxygen, methane, water) and utilities (HP steam, cooling and chilled water) are provided by a pipeline from OSBL. The entering water is demineralized, which assumes a demineralization unit outside the battery limit. A chilled water and cooling water facility is also assumed to be available OSBL. The treatment of purge streams is considered outside the battery limit as well as the treatment and storage of the products for further transportation.

→ The carbon dioxide removal is not considered in this design, and has been included as a black box in all PFD's, named C102.

Table 3.1 Product and feedstock specifications

Components	Temperature °C	Pressure bar	Quantity t/a	State (V/L/S)	Source	Prices US$/t
Feedstock						
Ethylene	Ambient	25	193,198	V	pipeline OSBL	520
Oxygen	Ambient	25	184,584	V	pipeline OSBL	50
Products						
EO	Ambient	2 ?	103,808	L	--	1,080
MEG	Ambient	1	205,328	L	--	650
By-products						
DEG	Ambient	--	1,332	L	--	551
CO_2	Ambient	--	75,248	V	--	--
Other Chemicals						
Methane	Ambient	25	11,592	V	pipeline OSBL	120
Water	Ambient	5	65,143	L	unit OSBL	2.5
Catalysts	Composition		Quantity	~~shape~~	Lifetime	Prices
EO-catalyst	silver on alumina		126 $M^3/3y$	~~cylinder~~	3 years	37,730
MEG catalyst Ethyl. Carb.	an alkyl ammonium salt		327 M^3/a	resin on spherical particles	1 year	11,764 US$/$m^3$

dimensions?

life time?

Further specifications of the feedstock, products, byproducts etc. are given in appendices 2, 3 and 4.

Table 3.2 Pure component properties

Pure component properties							
Component name		Technological data					Medical data
Design	Systematic	Formula	Mol. Weight g/mol	Boiling point[1] °C	Melting point[1] °C	Density of liquid[2] kg/m^3	MAC value mg/m^3
Acetaldehyde	Acetaldehyde	C_2H_4O	44.05	20.8	-121	783^{18}	100
Acetic acid	Ethanoic acid	$C_2H_4O_2$	60.05	117.97[7]	16.6	104^{20}	N.A.
Acetylene	Ethyne	C_2H_2	26.04	-84	81.8	620^{-82}	N.A.
Argon ✓	Argon	Ar	39.94	-185.7	-189.2	1400^{-186}	N.A.
Carbon dioxide ✓	Carbon dioxide	CO_2	44.01	-78.54[4]	-56.65[5]	1101^{-37}	9000
Carbon monoxide	Carbon monoxide	CO	28.01	-191.3	-207	793	29
Chlorine	Chlorine	Cl_2	70.9	-34.6	-100.98		3
Diethylene glycol	2,2'-oxybisethanol	$C_4H_{10}O_3$	106.14	245.8	-10.4	1119^{15}	N.A
Ethane	Ethane	C_2H_6	30.08	-88.6	-172	572^{-100}	N.A
Ethylene	Ethene	C_2H_4	28.05	-103.7	-1.69	567^{-104}	N.A.
Ethylene carbonate (EC)	Ethylene carbonate	$C_3H_4O_3$	88.06	248	39	1321	N.A.
Mono ethylene glycol (MEG)	1,2-ethanediol	$C_2H_6O_2$	62.07	197.3	-13	1108^{-20}	26
Ethylene oxide EO	1,2-ethane epoxide	C_2H_4O	44.05	10.6	-111.7	882^{10}	90
Formaldehyde	Methanal	CH_2O	30.3	-21	-92	815^{20}	1.5[8]
Hydrogen	Hydrogen	H_2	2.02	-252.8	-259.18		N.A.
Hydrogen sulfide	Hydrogen sulfide	H_2S	34.08	-60.4	-85.5	1539^{0}	15
Methane	Methane	CH_4	16.05	-161.5	-182.4	466^{-164}	N.A.
Methanol	Methanol	CH_4O	32.04	65.26[6]	-93.9	791^{20}	260
Nitrogen	Nitrogen	N_2	28.02	-195.8	-210	808^{-196}	N.A.
Oxygen	Oxygen	O_2	32	-183	-218.4	1149^{-183}	N.A.
Propylene carbonate	4,methyl-1,3-dioxolane-2-one	$C_4H_6O_3$	102.09	24.2	-48.8	1207^{20}	N.A.
Triethylene glycol	2,2'-(1,2-ethane diylbis(oxy)) bisethanol	$C_6H_{14}O_4$	150.2	285	-7	1127^{15}	N.A.
Water ✓	Water	H_2O	18.02	100	0	1000^{4}	N.A.

Notes:
1. At 1 bar
2. Superscript refers to the reference temperature (°C) at which density is measured.
3. Oral in g. for a male of 70 kg weight
4. Meltingpoint measured at 5.2 atm.
5. Sublimation point
6. Boiling point measured at 73 mm Hg
7. Boiling point measured at 10 mm Hg
8. MAC value in a mixture of 37 % water and 10% methanol

4. Thermodynamic Properties

Thermodynamic models that describe reality as accurately as possible are highly desirable, yet do not exist. If parameters for the models, based on experimental work, are available, model results may be good approximations of reality, but when experimental data is not available, or limited in its applicability, the choice becomes more difficult.
Initially, NRTL for the liquid activity coefficient calculations, combined with the equation of state model Redlich-Kwong for vapor phase calculations, was selected to model the required thermodynamic properties.
This model was expected to adequately model both the solubility of gases in absorbent, as well as the separation in the purification sections. This was found to be true for water as absorbent, but in the more polar MEG/EC absorbent mixture, the modeled solubility of gases, especially ethylene, was much to high in comparison with data from various sources (i.e. 7% was modeled versus 0-4% described in literature). The reason for this is because NRTL parameters are preferably retrieved from the experimental values as presented in Dechema databanks. In these databanks, most interaction parameters with water are available, but unfortunately no data on interaction with the EC/MEG absorbent mixture. Therefore, NRTL parameters are estimated with the UNIFAC group contribution method, resulting in inadequate modeling of the absorption of gases in the absorbent.
The Redlich-Kwong-Aspen model was eventually selected because it most adequately modeled the solubility of gases in MEG/EC absorbent.
A number of comparisons on which the model choice is based are listed below. The comparisons given first were valued the most, later comparisons are believed to be less applicable, valuable, or verifiable.

Data for the solubility of CH_4, CO_2, and N_2 at elevated temperatures and pressures in MEG are available from Zheng et al [ref.. 12], and was used to validate the RK-Aspen model. The model was found to very accurately describe data, as confirmed by Zheng. Zheng claims Soave-Redlich-Kwong discrepancies to be 2.10 % for CH_4/MEG, 2.04% for N_2/MEG, and 5.95% for CO_2/MEG. The simulations in Aspen agree to a large extend with these result.

The Mitsubishi patent [ref. 10] describes the solubility of EO in propylene carbonate and EC. Other gases (notably ethylene) are said not to dissolve in these absorbents. When modeling EO solubility in EC in Aspen, both the NRTL and the RK-Aspen models appear to approximate the experimental data. However, since MEG is not included in the experiments, this data is of limited value.

Tsang and Ainsworth, describe in US Patent 4221727 [ref. 13] to the Dow Chemical Co. the use of EC as absorbent for EO reactor gas. This patent confirms the absorption of approximately 4-5 mol% CO_2. In addition, the inventors claim "little or no ethylene, nitrogen, and oxygen were absorbed". In contradiction to this, patent EP 133763 [ref. 9], amongst other sources, doubts the ethylene absorption claimed by Tsang. Furthermore, the EO/MEG process described in this conceptual design makes use of an EC/MEG absorbent mixture instead of solely EC, hence this data is of limited value.

These comparisons justify the choice of the RK-Aspen model for this model, when employed it shows that less than 1 % of the ethylene is absorbed.

For reference, table 4.1 show the results of the primary EO absorber (C101), in which the EO reactor gas is contacted with an EC/MEG absorbent stream, as produced by the Aspen simulation. These results are reasonably consistent with the discussion above. However, the validity of this model in V-V and V-L, where L = H_2O, should be further researched for a more detailed design. The impact of possible discrepancies is expected to be minor.

design.

Table 4.1, primary EO absorber results

Component	Percentage absorbed (mol%) from gas	Concentration after absorption in rich absorbent (mol%) in liquid (max
EO	99.9	3.8 → determines flows
Ethylene	0.71	0.4
Methane	0.19	3.0
Oxygen	0.15	167 PPM
Argon	0.17	32 PPM
CO_2	3.78	0.5
Water	90.38	19.7

In conclusion, it is believed that the solubility of EO reactor gas in EC/MEG absorbent, the most important feature of our design, is reasonably modeled, and the thermodynamic model is adequate for the complete design simulations.

5. Process Structure and Description

Appendix 1 includes a graphical overview of the process, and process flow schemes for each of the sections, as defined below:

- Section 100 : Ethylene oxide (EO) production
- Section 200 : Monoethylene glycol (MEG) production
- Section 300 : MEG purification
- Section 400 : EO purification

The following chapter describes each of the units in the process, and the process as a whole.

5.1. Ethylene Oxide Production (Section 100)

Ethylene from OSBL is mixed into the loop gas, and methane is added as ballast in order to replenish the loss to the purge. An amount of oxygen based on the relative selectivity and activity of the EO catalyst is injected in the mixing station. Oxygen is added as the last component, so as to keep oxygen concentration as low as possible. A small amount of methylene chloride (not shown) must also be added as promoter.

The reactor feed gas is heated by exchange with the hot reactor effluent, and passed to the EO reactor. In the reactor the gasses are heated to reaction temperature with generated steam of the reaction section of the reactor. The reaction heat is removed by generating steam. After leaving the reaction section the gasses should be cooled directly to minimize EO decomposition. The EO reactor is described in detail in chapter 8.

After being cooled in heat exchangers E101 and E104 which are integrated with the reactor R101, the gas is counter-currently contacted with absorbent in the primary EO absorber. The EO is completely absorbed in the absorbent. Most of the scrubbed loop gas is cycled back to the reactor, approximately 7% is passed to the carbon dioxide removal unit. A small part is purged to prevent the build-up of impurities, notably argon.

Carbon dioxide removal is carried out in black box C102, where water and carbon dioxide are removed with 95% efficiency. The CO_2 removal system possibly removes CO_2 by contacting the lean cycle gas with a hot potassium carbonate solution, in which the carbonate is converted to potassium bicarbonate by reaction with CO_2. The CO_2-rich solution is then regenerated using stripping steam. Although mono-ethanol amine (MEA) is a more commonly used and better absorbent for CO_2 removal systems it cannot be used in the EO/MEG design. EO reacts easily with MEA to form polymers, thus MEA is unsuitable as an absorbent because of the always present trace amounts of EO in the loop gas with which it would react.

The carbon dioxide unit also scrubs loop gas and carbon dioxide stripped from the absorbent in later units. Specifically, loop gas is recovered from the secondary EO absorber, and CO_2 is recovered after each of the three reactors in the MEG production section. The carbon dioxide recovered is fed to the MEG reaction section (section 200). The CO_2-lean gas is returned to the bulk of the cycle gas.

Before the ethylene, oxygen, and methane are added to the loop gas, it is brought back to pressure in the loop gas compressor.

recovered

Absorbent is ~~gathered~~ from the two EO strippers from the EO purification (section 400) and from the MEG production section (section 300). A small amount of unpurified MEG is added from section 300, in order to preserve the composition of the absorbent. Let it again be clear that an approximately 1/1 mixture EC/MEG is necessary to prevent solidifying of absorbent plugging pipes, pumps, etc. The absorbent is first heat exchanged with the MEG reaction feed (E201), then cooled and chilled by the absorbent coolers (E105a and E105b), and fed to the top of the packed absorber.
The rich absorbent is split into a feed stream for the MEG reaction section and the EO purification section.

5.2. MEG Reaction Section (Section 200)

The rich absorbent feed from section 100 is brought to higher pressure so as to enable more CO_2 to dissolve in the stream. Before being fed to the EC reactor, the stream is heated to reaction temperature by exchange with the absorbent stream <126>. Carbon dioxide recovered in the CO_2 unit is fed in shots to the reactor, which is explained in more detail in chapter 8. A 1.2 molar excess of CO_2 to EO is used. The excess is removed in a flash vessel, together with any other loop gas components absorbed in the primary absorber (C101). The vapor stream is returned to section 100; the liquid stream is split (1:10) into an absorbent stream and a smaller stream for the conversion to MEG.

A small part of the EC/MEG absorbent stream <206>, which contains little or no EO or other gases, is sent to section 400 for use in the secondary EO absorber. The major part of the stream is returned to the primary EO absorber in section 100.

EO vessel. excess

Water is added to the remaining EC/MEG stream <208> and heated to reaction temperature. In the hydrolysis reactors, two of which are placed in series, CO_2 is relieved ~~liberated~~, and removed in the top of the fixed bed reactors with the use of chimney trays. The liquid residence time is approximately 60 minutes. Each of the hydrolysis reactors is followed by a condensation vessel (S202 and S301) for the vapor stream, from which MEG and, for in the first drum, MEG and EC are recovered. The liquid stream from these vessels is returned to the major MEG flow. Vapor streams are returned to the carbon dioxide unit in section 100.

5.3. MEG Purification Section (Section 300)

The crude MEG stream, polluted by water and DEG, is purified in section 300 in two vacuum columns. The water removal column (C301) separates water, aldehydes, and any carbon dioxide from the MEG/DEG stream, at a pressure of 0.1 bar. At this point a part of the MEG/DEG stream is split off and added to the bulk of the absorbent stream.

The MEG purification column (C302) operates at a pressure of 0.09 bar, splitting a purified MEG stream over the top and a DEG stream over the bottom. The MEG is condensed and sold as fiber-grade MEG. The DEG stream contains substantial amounts of EC and MEG, and is not further purified.

5.4. EO Purification Section (Section 400)

C402

The rich absorbent stream from C101 is thermally stripped, and compressed in two stages to 10 bar. It is subsequently reabsorbed in the secondary EO absorber, which draws lean absorbent from section 200. Gas leaving the top of this packed absorber is recycled to the carbon dioxide removal unit in section 100. The rich absorbent is again thermally stripped, this time at reduced pressure, with the recovered absorbent being added to the bulk absorbent stream.

The EO- and CO_2-rich vapor stream, after being pressurized in a multi-stage compressor, is partially condensed with cooling water, and fed to the light-end separation column (C405). A liquid EO stream containing trace amounts of CO_2 is withdrawn over the bottom, and a CO_2/EO stream is recycled to the secondary EO absorber. The carbon dioxide fraction in this recycle loop approaches 10%, the EO fraction is approximately 87%. The recycle ratio is approximately 3.7. Reducing this ratio would require more refluxing over the light-end separator (C405) and is not economically preferable.

The raw EO is subsequently passed through the heavy ends separator (C404). Heavy ends are withdrawn over the bottom, and sales specification EO is, after condensation, produced.

MEG purif. Col!

5.5. Heat Integration

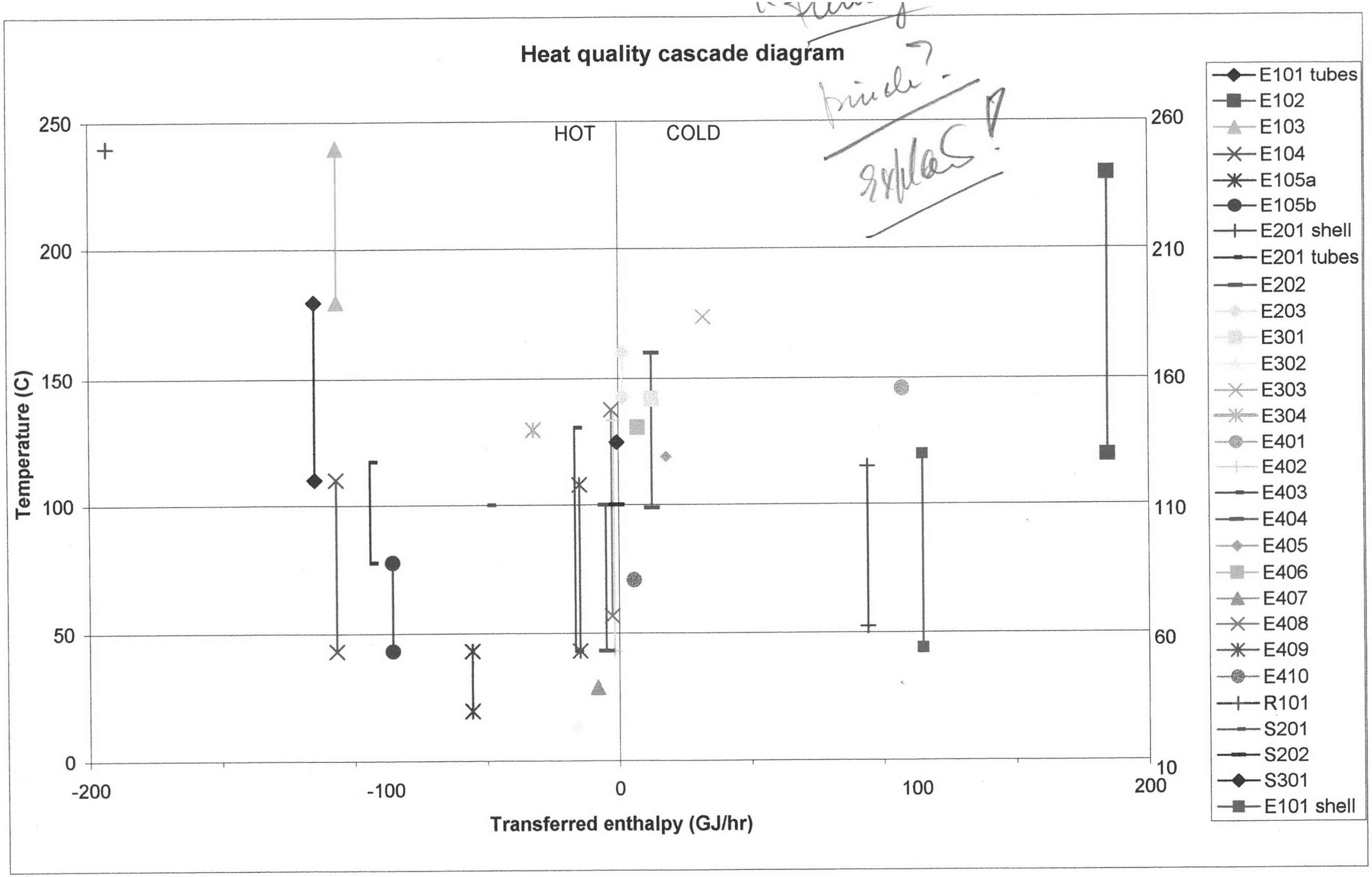
Heat integration of the EO/MEG process is done by means of exchanging some streams against each other and by steam generation and distribution. Most of the heat produced, is produced in the EO reactor. The heat removed in this reactor is of the highest quality in this design. The removal of this heat has to be carried out with the generation of steam. From a safety point of view it would be highly undesirable to lead another process stream through the reactor to remove the heat from the reaction stream. The generated steam will be used somewhere else in the process. Direct heat integration between streams is therefore limited to some places in the process, described below.

?

5.5.1. Heat integration between streams

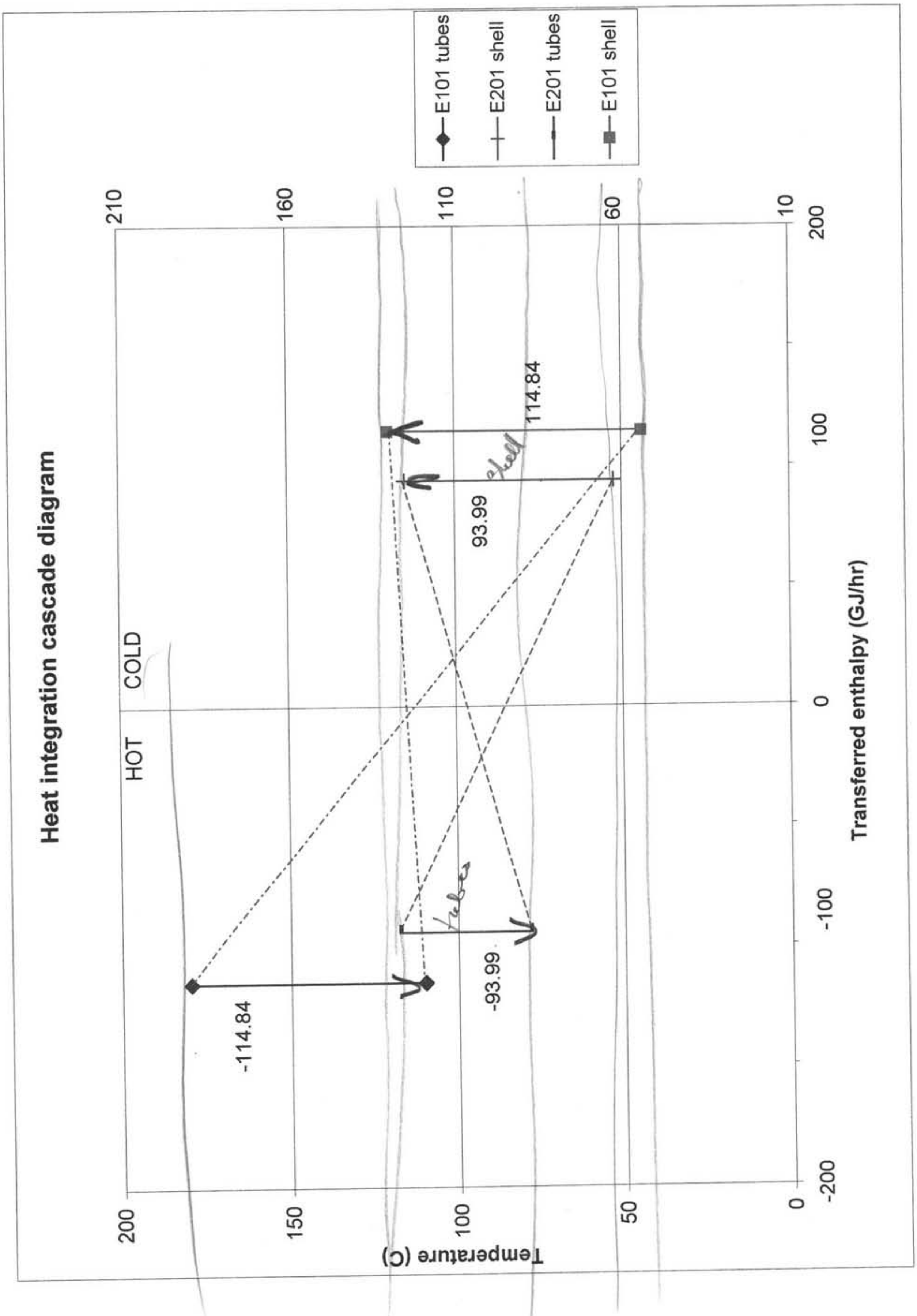
Feed/Prod.

Heat integration is not directly implemented in the Aspen flow sheet design except for exchanger E101. In Aspen, heat loss or gain of streams is modeled with basic heat exchangers using cold and hot utility to dissipate heat to and gain heat from. In order to integrate heat afterwards a cascade diagram is presented in figure 5.1. Direct heat integration between streams is determined by the heat quality of the streams. The temperature of a stream determines its heat quality; higher temperatures for hot streams and lower temperatures for cold streams result in higher heat qualities. As can be seen, many low temperature hot streams are available. Extensive heat integration is therefore not possible, simply because there is a lot of energy to lose from those low temperature hot streams and nowhere to lose it but in cold utility. Large amounts of cooling and chilled water are thus a necessity increasing the operating costs.
Also in figure 5.2 is demonstrated that heat exchange between reactor feed <106> and reactor effluent <110> in exchanger E101 is possible as well as between the main absorbent stream <126> and the carbonate reactor feed <201> in exchanger E201.



pinch?

explain!


Heat integration cascade diagram
E101 tubes
E201 shell
E201 tubes
E101 shell
HOT
COLD
114.84
93.99
-93.99
-114.84
shell
tubes
Temperature (C)
200
150
100
50
0
210
160
110
60
10
Transferred enthalpy (GJ/hr)
-200
-100
0
100
200

Because cooling of <110> and <126> is not sufficient, additional coolers are installed (E104 (cooling water), and E105a (cooling water) and E105b (chilled water) respectively). Implementing the exchangers E101 and E201 in the design reduces the necessary amounts of hot and cold utility substantially. Although some integration between streams in different units of the process is still possible this is not done because the amount of heat transferable is not of substantial influence on the amount of cold and hot utility necessary.

5.5.2. Steam generation and distribution

Three units in the process generate heat of high enough quality to generate steam (see table 5.1).
The EO reactor, in which highly exothermic reactions occurs, generates an amount of 193.05 GJ/hr. The outlet temperature of the reaction section is 240 °C. Therefore HPS is generated with the heat generated in this section.
Heat exchanger (E103) is part of the EO reactor (R101) and is meant to cool the effluent of the reaction section in order to prevent isomerization to acetaldehyde. The reactor effluent is cooled from 240 °C to 180 °C, while medium pressure steam is being generated.
In E304, the condenser of distillation column C302, operating at a temperature of 130 °C, LPS is generated.

Table 5.1 Steam generating units

Name	Unit	Heat Duty GJ/hr	Steam	Steam Temperature °C	Mass flow kg/hr	Steam Pressure bar
EO Reactor	R101	-193.05	HPS	230	106557	27.6
Heat exchanger	E103	-106.38	MPS	160	51117	7.8
Condenser C302	E304	-32.20	LPS	120	14621	1.9

In table 5.2 the units that require HPS for heating are listed.
E303, the reboiler of column C302, is operated at a temperature of 164 °C, thus HPS is needed to heat the reboiler. ΔT ?
E102, the feed preheater, which is part of the EO reactor, heats the reaction section inlet with the HPS generated in the reaction section. The reboiler of C401 (E401) is the third unit that requires HPS for heating. Unfortunately, the EO reactor (R101) doesn't generate enough steam to operate all HPS consuming equipment. Therefore additional HPS is acquired from outside the battery limit.

Table 5.2 HPS requirement

Name	Unit	Heat Duty GJ/hr	Steam mass flow kg/hr
Reboiler C302	E303 ✓	32.00	17663
Feed preheater	E102 ✓	184.64	101918
Reboiler C401	E401 ✓	107.3	59226
Total			178807

The majority of the heat requiring units are operating in a temperature range of 100 °C to 150 ° C, for these units MPS is used to provide their heat requirement. Table 5.3 provides a list of these units. The total amount of MPS generated in E103 is more than sufficient and therefore an amount of 34840 kg/hr MPS is leftover. This MPS is elsewhere upgraded to HPS.

Table 5.3 MPS requirement

Name	Unit	Heat Duty GJ/hr	Steam mass flow kg/hr
Reboiler C301	E301	12.6	6054
Reboiler C404	E406	7.3	3512
Exchanger 202	E202	12.5	6008
Exchanger 203	E203	1.5	703
Total			16277

The LPS generated in E403, the condenser of distillation column C302, is used in E405, the reboiler of the secondary EO stripper C403, and in E410, the reboiler of light end separator C405. The LPS requirements are shown in table 5.4

Table 5.4 LPS requirement

Name	Unit	Heat Duty GJ/hr	Steam mass flow kg/hr
Reboiler C403	E405	18.00	8173
Reboiler C405	E410	5.7	2588
Total			10761

Again, 3860 kg/hr of LPS is leftover. This is elsewhere upgraded to HPS and used in the process.

An overall view of the steam distribution of the EO/MEG process is given in figure 5.3.

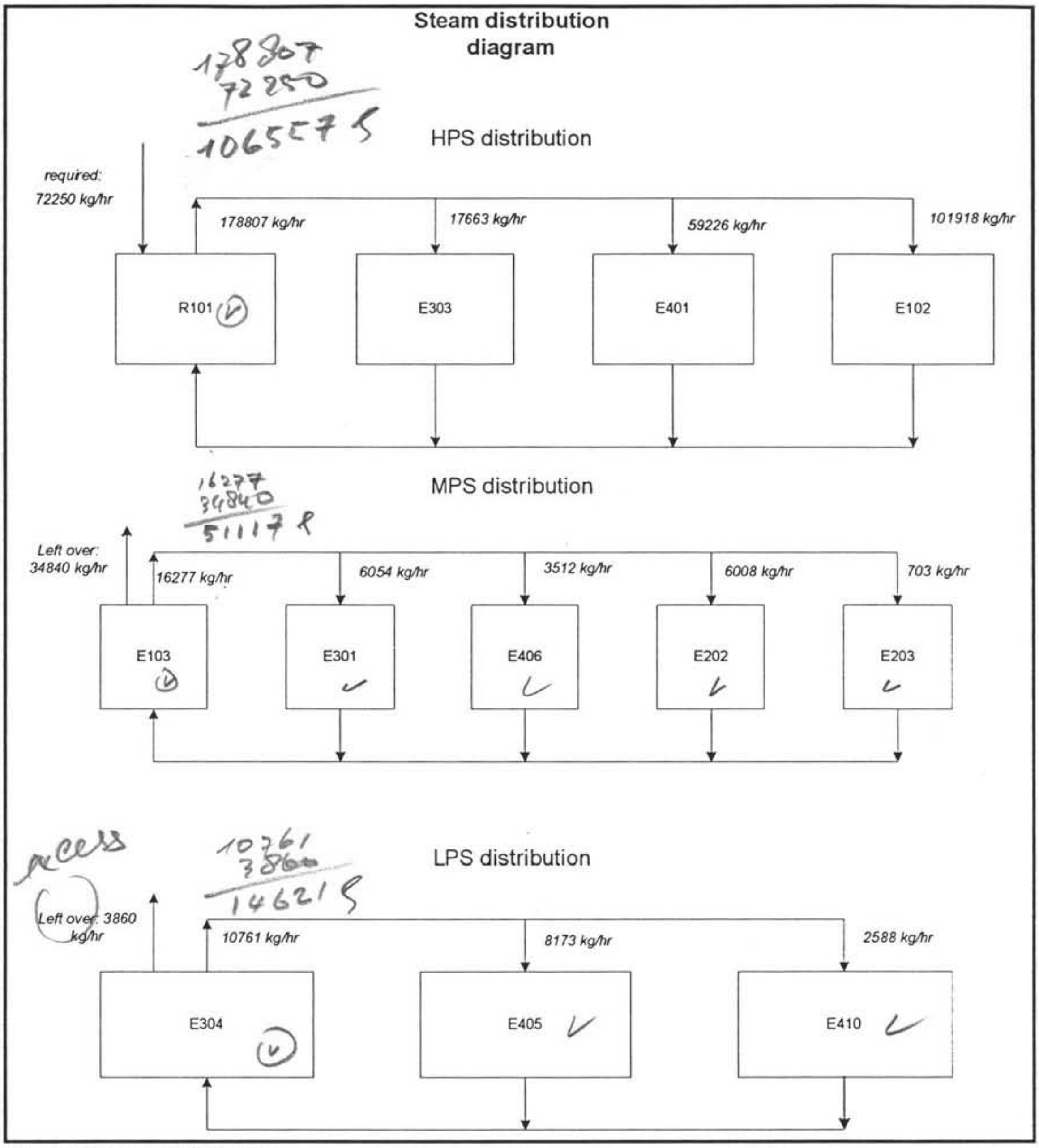


Fig. 5.3 Steam distribution

5.5.3. Cooling requirements

Most of the units that require cooling are cooled to a temperature of 43 °C. This is the minimal temperature that can be accomplished with the use of cooling water. Some streams require cooling to lower temperatures. For this purpose chilled water is required. Cooling water is provided at a temperature of 33 °C. Chilled water is provided at a temperature of 4 °C. Both cooling and chilled water are obtained from outside the battery limit. In table 5.5 the units that require cooling, by means of cooling or chilled water are shown.

Table 5.5 Cooling units

Unit	Duty GJ/hr	Process stream		Cooling medium			
		T in	T out	Tin	Tout	Type	Mass flow t/hr
S202	-0.80	133	100	33	43	CW	19
S301	-0.54	132/125	125	33	43	CW	13
E104	-106.79	110	43	33	43	CW	2,555
E105a	-40.82	43	20	4	10	CHW	2,210
E105b	-194.305	117	43	33	43	CW	2,049
E302	-15.89	14	14	4	10	CHW	630
E402	-1.95	134	43	33	43	CW	47
E403	-16.69	131	43	33	43	CW	399
E404	-4.89	100	43	33	43	CW	117
E407	-8.28	29	29	4	10	CHW	328
E408	-2.68	138	57	33	43	CW	64
E409	-14.89	108	43	33	43	CW	356

CW = Cooling water, CHW = Chilled water

6. Process Control

In this chapter the concepts of process control in the EO/MEG plant design are outlined. For every unit operation some basic process parameters will be controlled to provide safe and normal operation (in the steady state). Since the performance of the EO reactor and the feed mixing is of the utmost importance for a safe process operation, some additional process control is implemented for this purpose. Some aspects of the process control will be described in more detail below. The process control loops are shown in the PFD (appendix 1).

6.1. EO Reactor and Cooling

The cooling of the reactor is controlled by measuring the temperature in the corresponding reactor section and controlling the steam pressure. In other words: if the temperature of the reactor raises the pressure of the steam is lowered, allowing heat to be removed faster. The inflow of boiler feed water is controlled by a liquid level controller, which measures the liquid level in the steam drum. This cooling mechanism is a very important part of the reactor operation, because of the highly exothermic reaction. If the heat of reaction cannot be removed surely a thermal runaway will definitely occur, with disastrous consequences.

For safety reasons 'emergency control' is necessary for the temperature of the reaction section. When the temperature of the reaction stream exceeds an upper limit, the emergency valve will open. The feed and the contents of the reactor (by pressure relieve) will be led to the flare. The ethylene, methane, and oxygen feeds will be closed and nitrogen gas will be fed to the system at the end of the reactor to empty the system (reactor and recycle loop) of the dangerous reaction medium. This because a stagnant EO containing gas presents a serious hazard (decomposition of EO resulting in hot spots and a local runaway). In case this emergency system interferes with other controllers, override control is used with a high selection switch. This high selection switch is switching from normal control action to 'emergency control action' and attempt to prevent a process variable from exceeding an allowable upper limit.

6.2. Carbonate Reactor

The EO feed and the CO_2 feed to the carbonate reactor are controlled with ratio control so the two feeds are added in stoichiometric amounts. The liquid level is controlled by the bottom stream. To prevent the creation of large vapor phase a pressure controller in the top of the reactor also controls the CO_2 feed. Under normal operation the ratio controller controls the CO_2 feed of which most will dissolve in the liquid phase. Only when insufficient dissolving takes place the pressure controller prevents the formation of a large vapor phase.

6.3. MEG Reactors

The liquid levels on the chimney trays are kept constant, in order to maintain sufficient residence time for gasses to escape the liquid phase, by a controller that controls the liquid effluent of the reactors. The pressure is controlled by the overhead vapor flow.

6.4. Absorbers

For the absorbers to work properly the ratio between the absorbent stream and the feed, with the components to be absorbed, has to be kept constant. For this purpose ratio control is applied between both streams. The liquid level in the absorber is controlled by the flow of the bottom stream, to prevent a liquid build-up in the absorber. The pressure is controlled by the flow of the overhead stream.

6.5. Strippers

The strippers used are all thermal. In order to operate them properly the temperature in the column is controlled by the steam flow through the boiler. The liquid level is controlled by the flow of the bottom stream withdrawn. The pressure in the stripper is controlled by the flow of the overhead stream.

6.6. Distillation Columns

The liquid level inside the column is controlled by the flow of the bottom stream, and the pressure in the distillation column is controlled by the stream of coolant through the condenser. The temperature in the column is controlled by the steam flow through the boiler. The reflux ratio is controlled with a flow controller in the reflux stream. The liquid level in the knock out vessel is controlled by the condensed top stream of the distillation column. In case of non-total condensation a pressure controller coupled with the vapor flow or vacuum pump controls the pressure.

6.7. Flash-Vessels and V/L Separators

In these vessels the liquid levels are controlled by the liquid bottom streams and the pressure is controlled with the overhead vapor stream. In the case that heat-coils are present (for instance S201 and S202), the temperature of the vessel is controlled with the flow of the coolant medium.

6.8. Mixing Points and Splits

Mixing points and split points of streams are mostly controlled with ratio control. The mixing points for the fresh feeds of the EO reactor have an additional override control for safety reasons. An analysis controller controls the flow of fresh oxygen, because the oxygen concentration is important for the safety of the operation. If reactor influent exceeds its limits the reaction mixture will become more dangerous. Furthermore this analysis controller will also measure the concentration EO. Also it is most important for safety reasons that the build up of EO in the reactor feed is as low as possible to prevent the effluent to exceed the explosion limits (see also chapter 10).

7. Mass and Enthalpy Balances

Mass and enthalpy balances are calculated on basis of flow sheet simulations conducted with Aspen Plus. No major discrepancies were found; the causes of minor discrepancies are noted in the relevant sections.

Table 7.1 reflects the component mole and mass balances. Mass balances over each of the sections of the process are subsequently presented, including an overall mass balance. Enthalpy balances are presented in a similar fashion in chapter 7.2.

7.1. Mass Balances

Table 7.1 Component mole and mass balances

Conventional Components (kmol/hr)	In	Out	Production	Relative Diff.
EO	0	295	295	-1.03E-05
MEG	0	414	414	-1.13E-04
ETHYLENE	860.836	30	-831	6.22E-08
WATER	452	294	-179	-4.71E-02
METHANE	90.3211	90	0	2.00E-03
OXYGEN	719.287	7	-712	-2.99E-04
ARGON	1.41865	1	0	1.58E-03
EC	0	0	0	-3.87E-02
CO_2	455	692	237	-9.81E-07
DEG	0	2	2	-1.92E-05
ACETALDEHYDES	0	0	0	-2.66E-05
Conventional Components (kg/hr)	In	Out	Production	Relative Diff.
EO	0	12975	12975	-1.32E-01
MEG	0	25726	25723	-2.92E+00
ETHYLENE	24146.4498	829	-23318	-2.52E-02
WATER	8145.04	5304	-3225	-3.84E+02
METHANE	1449.653655	1447	0	2.90E+00
OXYGEN	23017.184	233	-22791	-6.87E+00
ARGON	56.660881	57	0	8.95E-02
EC	0	10	9	-3.73E-01
CO_2	20024.55	30462	10437	-4.40E-02
DEG	0	190	190	-4.25E-03
ACETALDEHYDES	0	0	0	-2.64E-06

7.1.1. Section 100 - EO Reaction Section

Mass stream in	Stream number	Flow (kg/hr)	Mass stream out	Stream number	Flow (kg/hr)
Ethylene feed	101	24150	CO2 purge	131	9406
Methane feed	102	1449	CO2 to MEG reaction section	202	20024
Oxygen feed	103	23073	Water purge	132	851
Loop gas from EO purification	421	3337	Argon purge	118	3032
CO2 from MEG purification	312	9732	EO to MEG reaction section	201	675184
CO2 from MEG reaction	220	9541	EO to EO purification	401	473088
CO2 from MEG reaction	207	1916			1181586
Absorbent from MEG reaction	128	583963			
Absorbent from MEG purification	310	28032			
Absorbent from primary EO stripper	414	456295	Discrepancy (kg/hr):		4
Absorbent from secondary EO stripper	420	40094			
		1181582			

7.1.2. Section 200 - MEG Reaction Section

Mass stream in	Stream number	Flow (kg/hr)	Mass stream out	Stream number	Flow (kg/hr)
Reaction water	223	8143	CO2 to EO reaction section	207	1916
CO2 from EO reaction section	202	20025	CO2 to EO reaction section	220	9541
EO feed from reaction section	201	675184	V-effluent to MEG purification	222	9777
		703351	L-effluent to MEG purification	221	58154
			Absorbent to EO reaction section	128	583963
			Absorbent to EO purification section	422	40000
					703351
			Discrepancy (kg/hr):		0

7.1.3. Section 300 - MEG Purification Section

Mass stream in	Stream number	Flow (kg/hr)	Mass stream out	Stream number	Flow (kg/hr)
V-effluent from MEG purification	222	9777	CO2 to EO reaction section	312	9732
L-effluent from MEG purification	221	58154	Water / aldehydes (V)	308	1152
		67931	Water / aldehydes (L)	313	3139
			Purified MEG	304	25666
			Heavy components	309	210
			Absorbent to EO reaction section	310	28032
					67931
			Discrepancy (kg/hr):		0

7.1.4. Section 400 - EO Purification Section

Mass stream in	Stream number	Flow (kg/hr)	Mass stream out	Stream number	Flow (kg/hr)
EO from EO reaction section	401	473088	Loop gas to EO reaction section	421	3337
Absorbent from MEG reaction section	422	40000	Purified EO	419	12976
		513088	Heavy components	418	387
			Absorbent to EO reaction section	420	40094
			Absorbent to EO reaction section	414	456295
					513088
			Discrepancy (kg/hr):		0

7.1.5. Overall Mass Balance

Mass stream in	Stream number	Flow (kg/hr)	Mass stream out	Stream number	Flow (kg/hr)	
Ethylene feed	101	24150	Argon purge	118	3032	3033
Methane feed	102	1449	Water purge	132	851	851
Oxygen feed	103	23073	Water / aldehydes (L)	313	3139	3139
Water	223	8143	Water / aldehydes (V)	308	1152	1152
		56815	MEG (purified)	304	25666	25666
			Heavy components	309	210	210
			EO (purified)	419	12976	12976
			Heavy components	418	387	387
			CO2 purge	131	9406	9406
					56819	
			Discrepancy (kg/hr):		4	

The 3.9 kg/hour discrepancy the balance over section 100 and in the overall mass balance is due to the tolerance set during Aspen simulation. This tolerance was set at 0.0001.

7.2. *Enthalpy Balances*

7.2.1. Section 100 - EO Reaction Section

Stream in	Stream number	Flow (GJ/hr)
Ethylene feed	101	44.3
Methane feed	102	-6.8
Oxygen feed	103	-0.1
Loop gas from EO purification	421	-18.2
CO2 from MEG purification	312	-87.4
CO2 from MEG reaction	220	-85.9
CO2 from MEG reaction	207	-9.6
Absorbent from MEG reaction	128	-4212.6
Absorbent from MEG purification	310	-199.2
Absorbent from primary EO stripper	414	-3268.8
Absorbent from secondary EO stripper	420	-290.8
		-8135.1

Unit	Duty: (kW)	(GJ/hr)
R101		-193.1
K101	3058.6	11.0
K102	745.4	2.8
K103	286.9	1.0
P101	331.6	1.2
P102	13.8	0.0
E102		184.6
E103		-106.4
E104		-106.8
E105a & E105b		-141.3
E201		-93.9
C101	0.0	0.0
C102	0.0	0.0
	Sum	-440.6

Stream out	Stream number	Flow (GJ/hr)
CO2 purge	131 (*)	
CO2 to MEG reaction section	202	-176.7
Water purge	132 (*)	-97.7
Argon purge	118	-9.4
EO to MEG reaction section	201	-4877.3
EO to EO purification	401	-3417.4
		-8578.6

Stream difference	443.5 GJ/hr
Units	-440.6 GJ/hr
Discrepancy	2.8 GJ/hr
Correction R101	2.9
Corrected discrepancy	-0.1

(*) <131> and <132> combined in Aspen

Duty over the EO reactor (R101) is inconsistent with stream enthalpies due to heat of reaction specification, resulting in the major part (2.8 GJ/hr) of the above discrepancy. When the heats of reaction are specified in Aspen, these will be used to calculate the heat duty of the reactor, but stream enthalpies are always based on the elemental enthalpy reference state.

7.2.2. Section 200 - MEG Reaction Section

Stream in	Stream number	Flow (GJ/hr)	Stream out	Stream number	Flow (GJ/hr)
Reaction water	223	-129.1	CO2 to EO reaction section	207	-9.6
CO2 from EO reaction section	202	-176.7	CO2 to EO reaction section	220	-85.9
EO feed from reaction section	201	-4877.3	V-effluent to MEG purification	222	-87.6
		-5183.1	L-effluent to MEG purification	221	-442.5
			Absorbent to EO reaction section	128	-4212.6
			Absorbent to EO purification section	422	-288.6
					-5126.8
Unit	Duty : (kW)	(GJ/hr)			
P201	257.1	0.9			
P202	0.1	0.0	Stream difference		-56.3 GJ/hr
P203	2.1	0.0	Units		58.8 GJ/hr
E201		94.0	Discrepancy		2.5 GJ/hr
E202		12.5	→ Correction R201		2.4 GJ/hr
E203		1.5	Corrected discrepancy		0.1 GJ/hr
S201		-49.3			
S202		-0.8			
	Sum	58.8			

Duty over the carbonate reactor (R201) is inconsistent with stream enthalpies due to heat of reaction specification, resulting in the major part (2.44 GJ/hr) of the above discrepancy.

7.2.3. Section 300 - MEG Purification Section

Stream in	Stream number	Flow (GJ/hr)	Stream out	Stream number	Flow (GJ/hr)
V-effluent from MEG purification	222	-87.6	CO2 to EO reaction section	312	-87.4
L-effluent from MEG purification	221	-442.5	Water / aldehydes (V)	308	-14.2
		-530.2	Water / aldehydes (L)	313	-48.8
			Purified MEG	304	-182.8
			Heavy components	309	-1.2
			Absorbent to EO reaction section	310	-199.2
					-533.6
Unit	Duty : (kW)	(GJ/hr)			
S301		-0.5			
C301 reboiler		12.6	Stream difference		3.5 GJ/hr
C301 condenser		-15.9	Units		-4.0 GJ/hr
C302 reboiler		32.0	Balance		-0.5 GJ/hr
C302 condenser		-32.2			
P303	2.4	0.0			
	Sum	-4.0			

7.2.4. Section 400 - EO Purification Section

Stream in	Stream number	Flow (GJ/hr)
EO from EO reaction section	401	-3417.4
Absorbent from MEG reaction section	422	-288.6
		-3706.0

Stream out	Stream number	Flow (GJ/hr)
Loop gas to EO reaction section	421	-18.2
Purified EO	419	-22.7
Heavy components	418	-6.0
Absorbent to EO reaction section	417	-290.9
Absorbent to EO reaction section	414	-3268.9
		-3606.6

Unit	Duty : (kW)	(GJ/hr)
C401 reboiler		107.3
C403 reboiler		18.0
C404 reboiler		7.3
C404 condenser		-8.3
C405 reboiler		5.7
K401	448.5	1.6
K402	644.0	2.3
K403	205.6	0.7
K404	1067.6	3.8
K405	451.1	1.6
P401	165.2	0.6
P404	20.0	0.1
E402		-2.0
E403		-16.7
E404		-4.9
E408		-2.7
E409		-14.9
	Sum	99.7

Stream difference	-99.4	GJ/hr
Units	99.7	GJ/hr
Balance	0.3	GJ/hr

7.2.5. Overall Enthalpy Balance

Unit	Duty : (kW)	(GJ/hr)	Stream in	Stream number	Flow (GJ/hr)
C301 condenser		-15.9	Ethylene feed	101	44.3
C301 reboiler		12.6	Methane feed	102	-6.8
C302 condenser		-32.2	Oxygen feed	103	-0.1
C302 reboiler		32.0	Reaction water	223	-129.1
C401 reboiler		107.3			-91.7
C403 reboiler		18.0			
C404 condenser		-8.3			
C404 reboiler		7.3	Stream out	Stream number	Flow (GJ/hr)
C405 reboiler		5.7	Total purge	131+132 (*)	-97.7
E102		184.6	Argon purge	118	-9.4
E103		-106.4	Water / aldehydes (V)	308	-14.2
E104		-106.8	Water / aldehydes (L)	313	-48.8
E105		-141.3	Purified MEG	304	-182.8
E202		12.5	Heavy components	309	-1.2
E203		1.5	Purified EO	419	-22.7
E402		-2.0	Heavy components	418	-6.0
E403		-16.7			-382.8
E404		-4.9			
E408		-2.7			
E409		-14.9	(*) 131 and 132 combined in Aspen		
K101	3058.6	11.0			
K102	745.4	2.7			
K103	286.9	1.0			
K401	448.5	1.6			
K402	644	2.3			
K403	205.6	0.7			
K404	1067.6	3.8			
K405	451.1	1.6	Stream difference	291.1 GJ/hr	
P101	331.6	1.2	Units	-286.3 GJ/hr	
P102	13.78	0.0	Balance	4.8 GJ/hr	
P201	257.146	0.9	Correction R101 & R201	5.4 GJ/hr	
P202	0.2	0.0	Corrected discrepancy	-0.5 GJ/hr	
P203	2.1	0.0			
P303	2.4	0.0			
P401	165.227	0.6			
P404	19.962	0.1			
R101		-193.1			
S201		-49.3			
S202		-0.8			
S301		-0.5			
		-286.3	Sum		

8. Process and Equipment Design

8.1. Integration by Process Simulation

A steady state flow sheet modeling program is used to model the design. For this purpose Aspentech's Aspen Plus 10.0-1, provided by Stork E&C, was used.

8.2. Equipment Selection and Design

8.2.1. Ethylene Oxide Reactor (R101)

The ethylene oxide reactor is a multi-tubular reactor, in which the tubes are filled as a packed bed of catalyst particles. A multi-tubular reactor with a cooling effluent in the outer shell is chosen because of the necessity of substantial heat removal, together with the fact that a packed bed of catalyst is necessary.

What Catalyst ??.

An industrial reactor designed for EO Production by Shell/CRI or Halcon ref. 14, 15 is shown in figure 8.1. This reactor consists of three sections. After heat exchange between the influent and effluent gasses of the reactor, the influent (110 °C) is preheated to reaction temperature (220 °C) with steam produced in the reaction section. This preheat zone is symbolized by E102 in the PFD. In the reaction section the heat of reaction is removed and in the cooling zone (E103) the effluent gasses from the reaction section (240 °C) are cooled as quickly as possible to approximately 180 °C to minimize aldehyde production.

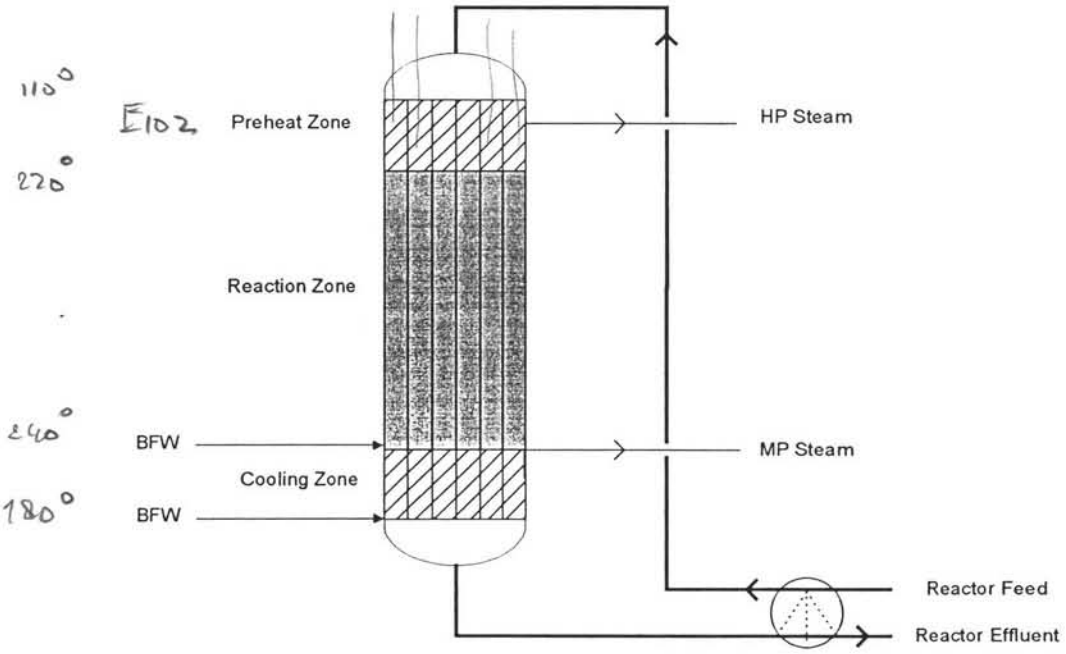


Fig 8.1 EO reactor

In the design two parallel reactors are used, consisting of 6265 tubes each and a volume of 90.8 m^3. Each tube has a diameter of 4 cm and a length of 11.5 meter, divided over the sections as follows: 2.2 m. for the preheat zone, 8 m. for the reaction zone and 1.3 m. for the quench. The reaction section is packed with spherical catalyst particles of 0.8 cm ref. 15. The catalyst is a silver-doped, on α-alumina based Shell catalyst. This catalyst is chosen for its high selectivity (86 %) for EO production. The productivity of this catalyst is 250 kg EO/m^3 reactor/h ref. 16. The single pass ethylene conversion is about 10% with a residence time of 6 seconds.

With the data of the catalyst productivity and EO mass flow, the volume of the catalyst bed is calculated. Since the geometry of the tubes is set ref. 14, 15, the number of tubes follows directly from the reaction volume, as well as the area available for heat transfer (see appendix 13).

The pressure drop is calculated with the Ergun equation. (For the bed porosity fig 5-56 on page 5-54 of Perry's Chemical Engineers' Handbook ref. 17 is used)

$$\Delta P := \frac{-v_{ms}\cdot\left(1-\varepsilon_{bed}\right)}{\rho\cdot d_p\cdot\varepsilon_{bed}^{\,3}}\cdot\left[\frac{150\left(1-\varepsilon_{bed}\right)\cdot\mu}{d_p}+1.75\,v_{ms}\right]\cdot L_{tube}$$

Where:
P: Pressure [Pa]
v_{ms}: Superficial mass velocity [kg/m^2s]
d_p: Particle diameter [m]
ε_{bed}: Bed porosity [-]
ρ: Density [kg/m^3]
μ: Viscosity [Ns/m^2]
L_{tube}: Tube length [m]

With this equation the pressure drop over the reactor is found to be 2.5 bar.

The amount of reaction heat to be removed is: 53625 kJ/s. The design of the EO reactor also depends on the fact that this heat has to be removed (this requires a minimum heat transfer area). Calculations based upon catalyst activity resulted in a reactor with sufficient heat transfer area. The heat of reaction is removed by producing steam from boiler feed water at a rate of 107 ton/h. The design of the preheat section and cooling section are described in heat exchanger design. Complete calculations and data sheets concerning the EO reactor can be found in the appendices.

Although in reality a reactor design as described above is used, the modeling is done by separating the reaction section, the preheating- and cooling section (Feed heater, Quench) and calculating them as stand alone operations.

The tubes of the reactor are constructed of stainless steel, because ethylene oxide is present. Possibly formed rust, if carbon steel would be used, together with ethylene oxide presents an explosion and fire risk. The outer shell is constructed of carbon steel, because no aggressive chemicals or EO are present.

8.2.2. Carbonate Reactor (R201)

The liquid feed of the carbonate reactor contains amounts of CO_2. The rest of the necessary CO_2 must be added gradually to replenish the liquid phase and to keep the

vapor volume in the reactor as low as possible, while the concentration CO_2 in the liquid phase will be kept high. For this purpose a trickle flow fixed bed reactor with intermediate feed positions for the gas phase is used[ref. 21, pg 567].
The liquid flow direction is from top to bottom (counter-current flow to the rising bubbles CO_2) , to accomplish a better CO_2 uptake than in the case of co-current flow. In the case of a normally fed bubble column a large gas recycle would be required and the vapor fraction in the reactor would also be higher resulting in much larger equipment[ref. 22].

The chosen counter-current flow reactor is shown in figure 8.2.

The reactor is constructed of stainless steel, because ethylene oxide is present. Possible formed rust, if carbon steel would be used, together with ethylene oxide is explosion dangerous.

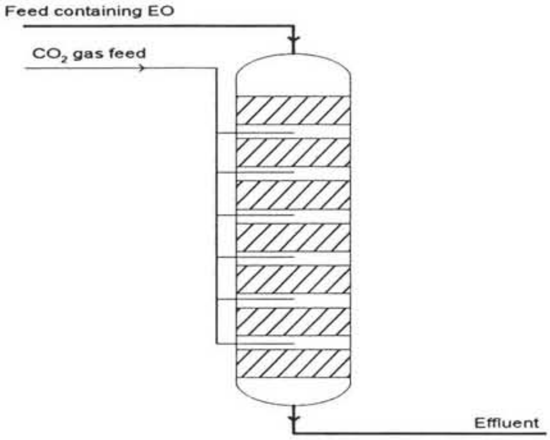


Fig 8.2 Carbonate reactor

reactor volume ?

The reactor has a volume of 202 m^3 filled with a spherical catalyst, comparable with DOW MSA-1. The productivity of the catalyst 300 kg conversion of EO/m^3/h[ref. 11].
With the data of the catalyst productivity and EO mass flow, the volume of the catalyst bed is calculated. This volume is corrected for a gas phase volume of approximately 10% (20 m^3) resulting in the volume stated above (see appendix 13).

The pressure drop is calculated with Ergun's equation and found to be 10 bar. The complete calculations and data sheets can be found in the appendices.
Because the heat of reaction is low, intermediate cooling is not required. The reaction stream will rise approximately 25 °C in temperature. The conversion to ethylene carbonate is nearly 100% with a selectivity of 99% or higher to the eventually produced MEG[ref. 8]. A residence time of the liquids of 19 minutes is required to accomplish to conversion described above.

8.2.3. Ethylene glycol reactors (R202, R203)

The reactors used to convert ethylene carbonate to ethylene glycol are shown in fig 8.3. The reaction takes place in two sequentially placed packed bed reactors, in order to keep the produced vapor flow within practical limits. These limits are determined by possible

↳ limitering!

equipment size, unwanted entrainment of liquids and solids in the vapor flow, substantial gas/liquid separation on top of the reactor. For the same reasons the reactor should have a substantial cross sectional area. A height over diameter (H/D) ratio of approximately 1.5 is used.

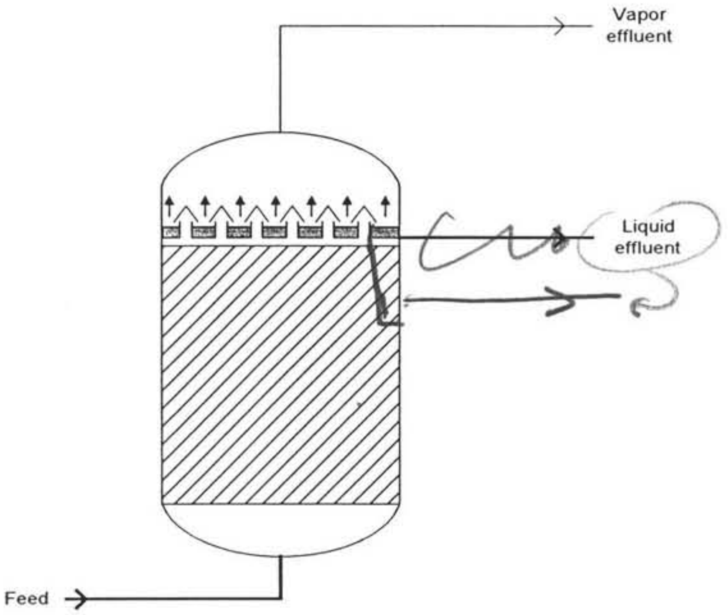


Fig 8.3 **MEG reactors**

A packed bed reactor is used without intermediate cooling because the heat of reaction is low. During the reaction with water CO_2 is released and a substantial vapor flow is produced. This physical process consumes heat, which nearly balances the heat produced by reaction. The reactor is filled with catalyst particles comparable with DOW MSA-1. The productivity of the catalyst for this reaction is 750 kg conversion of EC/m^3/h[ref. 11]. With the data of the catalyst productivity and EO mass flow, the volume of the catalyst bed is calculated. This volume is corrected for a gas phase volume of approximately 10% resulting in the volume stated above (see appendix 13).

In the top section of the reactor the gas liquid separation takes place. For this purpose chimney trays are installed. The vapor/liquid stream runs through the pipes of the chimney tray where at the top the liquid is forced down and the gas stream up. The liquid phase has a residence time of 3 minutes on top of the chimney tray for the gasses to escape.

The residence time in the reactors of 83 m^3 are approximately 1 hour to result in a conversion of nearly 100%. The selectivity over the carbonation step and the hydrolysis step is 99% or more towards MEG. The pressure drop over these reactors is low because the H/D ratio is low. For the calculation of the pressure drop the Ergun equation is used. A more complete calculation of the reactors can be found in the appendices together with the datasheets.

Both reactors are constructed of carbon steel, because no aggressive materials or EO are present.

8.2.4. Absorbers, strippers, and distillation columns

Separation calculations

The column sizing and design is done by computer simulation with Aspen as much as possible. The numbers of theoretical stages were preferably calculated with Aspen by the DSTWU shortcut distillation model, employing $R = 3 \times R_{min}$ as design criteria, but more preferably data from different patent literature[ref. 8] was used. The DSTWU model employs the Winn, Underwood, and Gilliland methods to estimate the minimum number of stages, the minimum reflux ratio, and the relation between the number of stages and the reflux ratio when one is specified. The Winn equation is a variation of the Fenske equation, wherein the relative volatility is not taken as constant, and is used to estimate the minimum number of stages. The Underwood equation is used for the calculation of the minimum reflux ratio.

Winn: $$\frac{d_i}{b_i} = \alpha_i^{N_m} \left[\frac{d_r}{b_r} \right]$$

Underwood: $$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = R_M + 1$$

Where:	d_i	flow-rate of component i in the tops
	b_i	flow-rate of component i in the bottoms
	d_r	flow-rate of the reference component in the tops
	b_r	flow-rate of the reference component in the bottoms
	N_m	minimum number of stages at total reflux, including the reboiler
	α_I	average relativity
	R_m	minimum reflux ratio
	$x_{i,d}$	concentration of component i in the tops at minimum reflux

and θ is the root of the equation:

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q$$

Where: $x_{i,f}$ the concentration of component i in the feed, and q depends on the condition of the feed, as used in McCabe-Thiele diagrams.

DSTWU also estimates the reboiler and condenser duty's, as well as the optimum feed stage location. Murphree efficiencies are rather arbitrarily taken from comparable separations[ref. 21, pg. 444]. Column internals were selected on the basis of capacity, pressure drop, contacting area, column dimensions, efficiency, and costs.

Sizing calculations

Mellapak structured packing, supplied by Sulzer Chemtech, are employed in the absorbers and strippers. This structured packing has been employed in industry for over

20 years, in columns up to 12 m in diameter. The availability of data [ref 34] and a computer-modeling program for structured packings by Sulzer (SulPack) allows reasonably detailed design. Sulzer Mellapak 250.Y is the most commonly available, and has been employed in the secondary absorber and both strippers (402, C401, C403). The higher temperature in the two strippers requires PVDF (max 150 °C), while PP can be used in the absorber (max 110 °C). The lower height equivalent to a theoretical plate motivated the choice for Mellapak 250.X for the primary absorber (PP packing) (C101) and the MEG/DEG separation column (CS packing, low pressure drop) (C302).

The columns where pressure drop and/or contacting area are either smaller or less of an issue all employ conventional sieve trays. A tray-spacing of 610 mm was adopted as initial estimate [ref. 20].

Data for Sulzer structured packing ,with the exception of HETP [HETP source: ref. 34], is available within Aspen, as well as experimental data for sieve trays, and thus Aspen can provide reasonable detailed column designs. The calculations are based on vendor-recommended procedures whenever these are available, as they are for Mellapak (Sulzer) and sieve trays (Fair). These calculations can be done with the RADFRAC model, but are not available in the DSTWU model. The relevant equations may be found in the Sulzer literature [ref 34].
Thus, this requires taking a preliminary DSTWU model for the number of theoretical stages, and subsequently simulating the same separation with the rigorous RADFRAC model, because the DSTWU model does not support packing or tray design. This process is fairly simple when the results of the shortcut distillation are used as inputs for the RADFRAC model.

As a simple check of the Aspen calculations, the Lowenstein equation (based on the Souders and Brown equation) was used to calculate column diameters for the sieve columns.

Lowenstein:

$$\hat{u}_v = (-0.171 \cdot l_t^2 + 0.27 \cdot l_t - 0.047) \cdot \left[\frac{(\rho_L - \rho_V)}{\rho_V}\right]^{1/2}$$

$$D_c = \sqrt{\frac{4\hat{V}_w}{\pi \rho_V \hat{u}_V}}$$

Where: $\hat{u}_v$ maximum allowable vapor velocity, based on the gross column cross-sectional area, [m/s]
l_t plate spacing, [m]
D_c column diameter [m]
V_w maximum vapor rate, [kg/s]

The results are show in the table below, and compared with the computer calculations.

Table 8.1 Comparison of column diameters

Column	Dimensions: diam [mm]	height [mm]	Lowenstein diam [mm]
C301	2770	6100	2589
C404	1220	11582	1281
C405	910	6100	972

The relatively large diameter of the water removal column (C301) is understandable when one considers the operation of this column : the evaporation of relatively large amounts of water, while withdrawing MEG over the bottom. Perhaps an (multi-effect) evaporator might have been a better design.

Ethylene oxide vapor is prone to polymerization and disproportionation, especially when rust with high surface area is available as catalyst. Therefore, stainless steel is employed in a number of columns, especially when the surface to volume area is high.

Equipment datasheets for the columns are included in appendix 11, the Lowenstein calculations are included in appendix 16.

8.2.5. Heat exchangers

Shell and tube heat exchangers are the most commonly used type of heat-transfer equipment in the chemical industries.
This type of heat exchanger is chosen because of:

- The large surface area it has in a small volume
- The fabrication techniques are well established
- This type of heat exchanger can be constructed from a wide range of materials
- The design procedures are well established

Most of the heat exchangers used in the process are fixed tube heat exchangers. The disadvantage of this type of heat exchanger is the fact that it can only be operated at a relatively low shell pressure (8 bar).

Design of the heat exchangers is carried out as described in Coulson & Richardson's Chemical Engineering [ref. 20].
The log mean temperature difference (LMTD) was calculated with the formula:

$$\Delta T_{\ln} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \dfrac{(T_1 - t_2)}{(T_2 - t_1)}}$$

Where: T_1 inlet shell side fluid temperature
T_2 outlet shell side fluid temperature
t_1 inlet tube side temperature
t_2 outlet tube side temperature

The overall heat transfer coefficients used based upon experimental data are presented in appendix 9. The overall heat transfer coefficients where obtained from literature [ref. 20, pg 580-581]. These coefficients together with the LMTD and the required heat duty are used to calculate the heat transfer area needed in an exchanger.
The heat exchange area is then calculated in the following way:

$$Q = U \cdot A \cdot \Delta T_{\ln}$$

Where: Q heat duty (GJ/hr)
U overall heat transfer coefficient (W/m^2K)
A heat exchange area (m^2)

In the tables below the heat exchangers of the entire process are listed per section of the process.

Table 8.2 Heat exchangers in section 100

Name	Duty (GJ/hr)	Number of exchangers	Area per exchanger (m^2)	Tube length per exchangers (m)	Tube diameter (m)	Number of tubes per exchanger
R101	-193.05	2	3464	8.00	0.044	6265
E101	114.84	4	993	4.88	0.025	2589
E102	184.64	2	1928	2.22	0.044	6265
E103	-106.38	2	2276	1.31	0.044	6265
E104	-106.79	4	825	4.88	0.025	2150
E105a	-55.68	1	477	2.44	0.025	2484
E105b	-85.63	1	825	2.44	0.025	4301

Table 8.3 Heat exchangers in section 200

Name	Duty (GJ/hr)	Number of exchangers	Area per exchanger (m^2)	Tube length per exchangers (m)	Tube diameter (m)	Number of tubes per exchanger
E201	-93.99	3	983	4.88	0.025	2562
E202	12.5	1	111	2.44	0.025	581
E203	1.46	1	24	2.44	0.025	123

Table 8.4 Heat exchangers in section 300

Name	Duty (GJ/hr)	Number of exchangers	Area (m^2)	Tube length (m)	Tube diameter (m)	Number of tubes
E301	12.60	1	100	2.44	0.025	521
E302	-15.89	1	337	2.44	0.025	1756
E303	32.00	1	148	2.44	0.025	772
E304	-32.20	1	894	2.44	0.025	4662

Table 8.5 Heat exchangers in section 400

Name	Duty (GJ/hr)	Number of exchangers	Area (m^2)	Tube length (m)	Tube diameter (m)	Number of tubes
E401	107.3	1	497	2.44	0.025	2589
E402	-1.95	1	49	2.44	0.025	257
E403	-16.69	1	431	2.44	0.025	2249
E404	-4.89	1	67	2.44	0.025	350
E405	18.00	1	143	2.44	0.025	745
E406	7.31	1	51	2.44	0.025	265
E407	-8.28	1	52	2.44	0.025	272
E408	-2.68	1	48	2.44	0.025	251
E409	-14.89	1	188	2.44	0.025	979
E410	5.70	1	26	2.44	0.025	138

When the area of the heat exchanger is known (calculated in the way described above) a tube length and diameter can be calculated. A diameter of 1 inch (2.54 cm) is usually chosen, because of its wide availability. The tube length is then chosen as a multiple of 2.44 m. (2.44 m or 4.88 m) because this is an industrial standard. Tube length, diameter and required area then determine the number of tubes.

The area per tube, the number of tubes and the shell diameter are calculated with the formulas below

$$A_{tube} = \pi \cdot d_{tube} \cdot L$$

$$N_t = \frac{A}{A_{tube}}$$

$$D_b = d_0 \left(\frac{N_t}{K_1} \right)^{\frac{1}{n_1}} + 0.5 \cdot d_o$$

Where:		
	A_{tube}	area per tube (m^2)
	D_{tube}	tube diameter (m)
	L	tube length (m)
	N_t	Number of tubes
	Db	Tube bundle diameter (m)

For sizing the reaction section of the reactor two aspects play a role: the reaction volume necessary and the heat transfer area necessary. The heat transfer area resulting from reaction volume calculations are sufficient for the reaction heat transfer.

The preheating and cooling zone of the reactor (E102 and E103) have some constrains determined by the reactor design. The number of tubes as well as the diameter and the tube pitch are the same as for the reaction section, because these three operations (preheating, reaction and cooling) are integrated in the same equipment. For the preheat section another design constraint is set by the available steam from the reaction section. This leads to design calculations by which only the tube length determines the area of heat exchange. The heat transfer coefficients are shown in tables 8.1-8.5. These together with the required heat duty, determine the tube length of the preheating and cooling sections.

8.2.6. Vessels

In the design of all knock out vessels, two main factors are of influence. First, the velocity of the vapor escaping the liquids and the liquid volume in the drum, and secondly, the velocity of the gases escaping the liquids must be lower than the gas velocity, as calculated with the following equation [ref 20, pg. 410].

$$u_v = 0.035 * \sqrt{\frac{\rho_L}{\rho_v}}$$

Where:

u_v = maximal gas velocity [m/s]
ρ_L = density of the liquid phase [kg/m^3]
ρ_v = density of the gaseous phase [kg/m^3]

The reason for this is that small liquid droplets that are always present should be able to settle down into the liquid phase. When the gas velocity is too high, small droplets are entrained with the out flowing gases.

This minimum velocity is then used to calculate a minimum cross sectional area for the vessel. Also, a disengagement height should be provided above the liquid level. This height is taken to be as high as the vessel diameter. The liquid level will depend on the hold-up time required for smooth operation and control, typically five minutes would be allowed and is thus used in this design. All calculations are presented in appendix 13

The streams flowing through all vessels is non-corrosive nor contains EO, hence, carbon steel suffices as construction material for all vessels. The characteristics of all vessels are presented in table 8.6.

Table 8.6 Vessel characteristics

Vessel	Orientation	Volume m^3	Height m	Diameter m
V101 Steam drum	Vertical	35.6	4.5	3.2
V102 Steam drum	Vertical	27.7	3.7	3.1
V301 V/L sep.	Vertical	1.2	1.3	1
V302 V/L sep.	Vertical	11.1	2.8	2.3
V401 V/L sep.	Vertical	1.1	0.7	1.4
S201 CO2 Flash drum	Vertical	131.9	8.8	4.4
S202 Flash drum	Vertical	0.9	1.1	1.0
S301 CO2 flash drum	Vertical	0.8	1.0	1.0

8.2.7. Pumps and compressors

The process-fluids to be transported by pumps have such properties and conditions that normal centrifugal pumps can be applied. Viscosities are not to high and only if the flows are very small (as for pumps P202, P304, P402) positive displacement pumps are applied. In all other cases centrifugal pumps are used.

The power of pumps is calculated with the formulae shown below. Some of the calculations are done by Aspen, which also calculates efficiency. For transport pumps not calculated by Aspen the equations below are used with an efficiency of 0.72.

$$P_{theoretical} = \phi_v \cdot \Delta p$$
$$P = \eta \cdot P_{theoratical}$$

Where: P: Power [w]
ϕ_v: Volume flow [m^3/s]
ΔP: Pressure change [Pa.]
η: Efficiency

The compressors are mostly rotary compressors, chosen with regard to the moderate pressure changes and the moderate to large flows. The two compressors used for the vacuum operation of the MEG purification columns (C301, C302) are liquid ring pumps [ref 21, pg. 149], because normal rotary compressors cannot provide low pressures on the suction side, but are designed to provide higher pressures on the discharge side.

The power of the compressors is calculated with the use of the equations below. For the efficiency of compressors 0.72 is used. Aspen, using the thermodynamic chosen model described in chapter 4, calculates all powers.

$$P_{isentropic} = \int_{P_1}^{P_2} \frac{1}{\rho} \partial p$$
$$P = \eta \cdot P_{isentropic}$$

Where: P: Power [w]
η: Efficiency
ρ: Density [kg/m^3]
p: pressure [Pa.]

For protection of the compressors knock-out drums are placed upstream of the machinery. Both pumps used for transportation of the liquid effluents of the MEG reactors (P205, P305) also have a vessel upstream to prevent possible vapor bubbles entering the pumps. Although these vessels are not concluded in this conceptual design they are mentioned here.
Since EO attacks most organic compounds, care must be taken in the choice of seals and gaskets.

9. Wastes

In this chapter the wastes produced by the process will be described. 'Waste' covers all compounds/utilities/streams produced, which do not belong to the products. These are summarized in table 9.1.

Table 9.1 Overview of the direct wastes

Wastes	Argon	CO_2	H_2O	H_2O old V	Stream number H_2O old L	Glyc HE	EO HE's		
kg/h State	118 V	131 V	132 L	308 V	313 L	309 L	418 L	total (kg/h)	total (t/a)
EO	0.001						0.1	0.13	1.05
MEG	0.045				53.5	25.7	2.9	82.13	657.06
Ethylene	828.899			0.1				828.96	6631.66
Water	1.817		850.8	586.4	3085.4	trace	383.9	4908.37	39266.95
Methane	1446.098			0.0				1446.10	11568.79
Oxygen	232.753			trace				232.75	1862.02
Argon	56.583			trace				56.58	452.66
EC	0.006					5.1	0.4	5.49	43.90
CO_2	465.668	9406.0		565.5				10437.12	83496.94 ✓
DEG	trace					179.0	0.0	179.01	1432.04
Aldehydes	trace						trace	0.1	0.9
Total -								—	— 1

The components that form the major part of the wastes are: contaminated water, methane, ethylene, oxygen and CO_2, in the liquid or vapor phase. The vapor wastes consist mostly of ethylene, methane, carbon dioxide, and oxygen. The liquid wastes, consist mostly of water contaminated with aldehydes, ethylene carbonate, MEG and DEG.

The vapor wastes are flared. This will result in large amounts of carbon dioxide discharged to the atmosphere (CO_2 from the process and CO_2 from combustion of the ethylene and methane wastes in the flare). Also some carbon monoxide resulting from combustion of ethylene and methane is discharged into the atmosphere.
The carbon dioxide resulting from flaring the ethylene and methane will result in approximately 50 kt/a extra CO_2. Hence, the total CO_2 discharge will be approximately 134 kt/a. Reduction of the CO_2 produced by the process is impossible with current technology. Only the development of EO catalysts with higher selectivity would aid in this, but this cannot be expected from the current generation of catalysts.

The liquid wastes are purged to the surface water. For components such as MEG and DEG this is not a problem, because they biodegrade easily. Furthermore, both components can effectively be treated in conventional wastewater treatment facilities. Aldehydes have a more substantial influence on the environment, but the amounts of aldehydes purged are very low; hence further treatment is not necessary before discharging.

Little EO is discharged as waste. EO is a hazardous chemical for environment and human health (MAC: 90 mg/m^3, LC$_{50}$^{fish}: 57-274 mg/l $^{ref. 31}$). Trace amounts will be discharged via the vapor phase. These small amounts will combust in the flare and do not form a problem to health or the environment. The largest part of the discharged EO is in the liquid phase (dissolved in water). The concentration of EO in the purge stream <418> is about 26.7 mg/l. This value is far under the LC_{50} value for fish. Furthermore when discharging this stream to the surface water this concentration will rapidly decrease (diluted by the liquid purge streams <132>, <313>,<309> and by the surface water itself).

10. Process Safety

10.1. Hazards of the Chemicals

The EO/MEG process is inherently unsafe because of the use of dangerous chemicals. Ethylene, methane, EO and MEG are all flammable and even explosive substrates, which, especially in the presence of oxygen, present a profound hazard. Data concerning fire and explosive limits of EO, ethylene and methane are presented in the tables below.

Table 10.1 Fire and explosion data for EO ref. 32

Property	Value
lower flammable limit[1], mol %	3.0
upper flammable limit[1], mol %	100
deflagration K_G index, J	18.8 - 24.6
deflagration P_{max} index, kPa	968 - 975
detonable range, mol %	5 - 30
theoretical flame temperature, °C	2129
auto-ignition temperature, °C	445
minimum ignition energy[2], J	$2.7 \cdot 10^{-2}$
minimum ignition energy[3] (10.4 mol %), J	$6 \cdot 10^{-5}$
1. In air 2. Vapor 3. Vapor-air mixture	

Table 10.2 Fire and explosion data for methane, ethylene and MEG ref. 33

Property	Methane	Ethylene	MEG
lower flammable limit[1], mol %	4.4	2.7	3.2
upper flammable limit[1], mol %	16.0	34	21.6
auto-ignition temperature °C	537	425	400
minimum ignition energy[2], J	$0.28 \cdot 10^{-3}$	$0.07 \cdot 10^{-3}$	
1. In air 2. Vapor			

As can be seen in these tables, it is of utmost importance to keep the concentration of EO and ethylene below 2.7 mol% or 4 mass%. To achieve this low concentration in the reaction gases a carrier gas is needed. For its superior heat capacity and stability methane is chosen as a carrier gas over other possible carrier gases such as nitrogen or other inert gases. The use of methane, however, requires a minimum concentration of this gas, 16.0 mol%, because of its flammability limits. During normal operation the methane concentration is always sufficient; however, this concentration should be monitored during startup and shutdown procedures.

The polymerization and disproportionation, as well as decomposition, of EO, especially when catalyzed by iron oxide in stagnant liquids and at raised temperatures, is a major hazard. The use of proper materials and the purging equipment with inert gases such as nitrogen after cleaning can minimize the formation of iron oxide.

10.2. The Process Safety

10.2.1. The EO reactor

Prevention of the presence of an explosive EO mixture is of the utmost importance in the EO reactor. This can and should be monitored by analyzing reactor inlet concentrations as well as reactor outlet concentrations of all substances. Some possible causes for a too high concentration of EO are listed below:

- The methane ~~inlet~~ is too ~~small~~. Low. Feed
- The ethylene and oxygen ~~inlets~~ are too large. high .
- The gas purge is too high decreasing the amount of loop gas, raising reactant concentrations.
- If EO is present in the reactor influent. Most probable reasons for EO at the reactor inlet are:
    - EO absorber (C101) failure – incomplete EO absorption
    - MEG reactor (R201) failure – incomplete conversion of EO
    - EO absorber (C402) failure – incomplete EO absorption

The failure of one of the absorbers may occur when problems with the absorbent feed arise, for instance, not enough absorbent throughput. Placing a hold-up vessel before each absorber, or perhaps using a shared vessel for the two absorbers, can reduce this risk. Also the operation of pumps P101 and P204 is critical for the proper operation of the absorbers, and therefore they are spared.
The use of analysis controllers warns the operators when the gas mixture approaches explosive limits.

10.2.2. The oxygen mixing station

In the oxygen mixing station explosive limits are locally exceeded. Unfortunately this can not be avoided, only minimized by use of proper mixing equipment. A gas analyzer placed behind the mixing station will monitor the gas composition, warning of the possibility of the occurrence of explosive gas compositions behind this mixing point. Because of the potential risk of explosion the mixing station is placed inside a concrete bunker.

10.2.3. The carbonate and hydrolysis MEG reactors

The concentration of chemically explosive gases in the carbonate reactor is low enough to make potentially dangerous situations improbable. A decrease in available absorbent could result in relatively higher gas flows and concentrations, but this situation can only occur when the primary EO absorber (C101) fails, and the results hereof are addressed elsewhere. In the hydrolysis reactors explosive dangers are avoided by the absence of oxygen and EO.

10.2.4. Piping

One potential hazard concerning the use of carbon steel for piping is the formation of iron oxide. EO in vapor form can react exothermally to form ethylene, hydrogen and CO_2 easily when catalyzed by iron oxide. When this occurs local temperatures can exceed the decomposition temperature of EO. This has been the reason of many industrial accidents with EO [ref. 31]. For this reason all piping through which EO in vapor passes should be stainless steel, or other ways protected.

10.2.5. Vessels

Both steam drums V101 and V102 present a hazard because of the large amounts of potential energy stored in the high temperature liquid water. A boiling liquid expanding vapor explosion (BLEVE) should be prevented. Overfilling of vessels, a main reason for the occurrence of BLEVEs, must be monitored carefully. The grounding of vessels, isolation from potential fire hazards, and the use of pressure relief devices help reduce this risk. In this design a liquid level controller is also installed. Further consideration of safety implementations are beyond the scope of this design.

10.2.6. Columns

The operation of the main EO absorption column (C101) is critical for the stable operation of the complete plant. Therefore this column should be carefully monitored because off-normal operation creates build up of chemicals in many streams creating potential dangerous situations. Also the EO heavy ends separation column (C404) is potentially dangerous because of the high EO concentration in the column and top stream. Explosion limits are unfortunately exceeded here but can under no circumstances be avoided because pure EO is one of the end products of this plant. The exceeding of explosion limits here justifies under no circumstances the exceeding of explosion limits in other places.

10.3. Dow Fire & Explosion Index Analysis

The Dow Fire and Explosion Index (F&EI) analysis provides key information to help evaluate the overall risk from fire and explosion for a design. The calculations are based on quantifiable data, resulting in realistic maximum losses that could actually be experienced under the most adverse operating conditions, instead of theoretically based risk figures. After calculating the F&EI, calculations can be carried out resulting in the maximum probable property damage (MPPD) and the business interruption (BI), as well as other safety related results. A proper F&EI analysis requires a complete process flow sheet and replacement costs of equipment, as well as the F&EI publications$^{\text{ref. 30}}$.
The procedure to be followed basically requires taking one pertinent part of the manufacturing unit (a process unit), consisting of, for example, a reactor and its peripheral equipment, and taking the single most hazardous chemical in the unit. In general, the most hazardous chemical should be present in more than 5% concentration. When these parameters have been set, all aspects of the chemical and process units are factored into a final F&EI, as shown below.

F&EI = General Process Hazards Factor (F_1) × Special Process Hazards Factor (F_2) × Material Factor (MF)

The process units and chemicals this analysis has been carried out for are shown in table 10.3

Table 10.3 Fire and Explosion indices for pertinent process units

Process unit	Chemical	MF	F_3	F&EI	Degree of hazard
EO reactors, including recycle compressor	ethylene	24	5	126	intermediate
EO absorber C101	EC/MEG mix.	5	1.8	9	light
EO purification column C403	EO	29	2.9	84	moderate
MEG purification columns C301 and C302	MEG	5	1.8	9	light

The relation between degree of hazard and the F&EI range is shown in table 10.4.

Table 10.4 Degree of hazard for F&EI (ref : F&EI 7th edition)

Degree of Hazard for F&EI	
F&EI Index Range	Degree of Hazard
1 – 60	Light
61 – 96	Moderate
97 – 127	Intermediate
128 – 158	Heavy
159 – up	Severe

The computation of the various factors is based on the conceptual design presented in this report. These calculations only consider the aspects that have been implemented in the design up to this point and issues that are so obvious that they can be considered in these calculations without serious deliberations. For example, drainage and access for

emergency services are not included, but the presence of EO in the reactor gas mixture is. This approach is considered to be the most valid at this point. Basic technological insight is required at every step to ensure reasonable and usable results.
Such results as the business interruption (BI) were also considered beyond the scope of this document.

All calculations and results are included in appendix 14.

11. Economy

In this chapter an estimate for the total costs for the EO/MEG plant (CAPEX) as well as an estimate for the operating expenses (OPEX) are calculated.

11.1. Fixed Capital

Fixed capital is the total costs of the plant ready for start-up. It is the cost paid to the contractors. It includes the costs of:

- Design, and other engineering and construction supervision.
- All items of equipment and their installation.
- All piping, instrumentation and control systems.
- Buildings and structures.
- Auxiliary facilities, like land and civil engineering work.

The equipment costs are calculated as accurately as possible using costing data from various sources. The costs of all other items in the previous list are estimated using the Lang factorial method [ref. 20, pg 219]. This means they are calculated as a fraction of the purchased equipment costs. It is therefore important to have an accurate estimate of the purchased equipment costs to prevent huge build-up of errors.

11.2. Estimation of Purchased Equipment Costs

Because the cost of purchased equipment (PCE) is used as the basis of the factorial method it must be estimated as accurately as possible. Many sources for equipment cost data are available, especially the journals Chemical Engineering and Hydrocarbon Processing publish costs lists on a regular basis. The problem with these lists is that fairly accurate knowledge of the size, shell thickness, construction material, etc of the equipment is necessary. Usually this accuracy is not included in a conceptual design. Fortunately, other methods are available. For the EO/MEG plant PCE Coulson and Richardson's Chemical Engineering [ref.20] is used, except for the costs of heat exchangers and reactors for which the Conceptual Cost Estimation Manual is used [ref. 25]. The Cost Estimation manual is used because the data from Coulson & Richardson's was not detailed enough for a reasonable estimate of these costs. The costs in both sources are not given in 1999 US dollars. To convert data from both sources to US dollars in 1999 an exchange rate of 1.6 $/£ is used. Furthermore, an index of 1.04 is used for the devaluation of money each year.

The calculations are included in appendix 15 and result in the equipment costs tabulated in table 11.1:

Table 11.1 Equipment costs

Equipment	Costs		Time base
Total columns costs	k$	1,836	1999
Total heat exchanger costs	k$	8,995	1999
Total compressor costs	k$	2,050	1999
Total pump costs	k$	556	1999
Total reactor costs	k$	8,872	1999
Total vessel costs	k$	136	1999
C102 CO2 removal system	k$	7,000	1999
Total purchased equipment costs (PCE)	**k$**	**29,310**	1999

11.3. Calculation of Fixed Capital

As mentioned before, the fixed capital is estimated by multiplying the PCE with a Lang factor. This means that fixed capital costs are calculated using formula:

$$C_f = f_L * C_e$$

Where:

	C_f	fixed capital costs
	f_L	the Lang factor
	C_e	costs of all major equipment

For a predominantly fluids processing plant, as the EO/MEG plant is, a Lang-factor of 4.7 is usually taken. To make a more accurate cost estimate, the cost factors that are compounded into the Lang-factor can be considered individually. The direct-costs items included in the fixed capital in addition to the costs of all major equipment and their individual Lang factor are demonstrated in table 11.2.

Table 11.2 Lang factors and costs

Item			Lang factor	Costs	
PCE		All major equipment	PCE	k$	29,310
	f_1	Equipment erection	0.40	k$	11,724
	f_2	Piping	0.70	k$	20,517
	f_3	Instrumentation	0.20	k$	5,862
	f_4	Electrical	0.10	k$	2,931
	f_5	Buildings, process	0.15	k$	4,396
	f_6	Site development	0.05	k$	1,465
	f_7	Utilities	0.50	k$	14,655
	f_8	Storages	0.15	k$	4,396
PPC		Total physical plant costs = PCE*(f_1+..+f_8)	PCE * 3.25	k$	95,256
	f_9	Design and engineering	0.30	k$	28,577
	f_{10}	Contractor's fee	0.05	k$	4,763
	f_{11}	Contingency	0.10	k$	9,526
FC		Fixed capital	PPC * 1.45	k$	138,121

PCE × 4.71

With this detailed cost estimation, a Lang factor of 4.55 is compounded. The values given above are derived from historical cost data for fluids type processes[ref 20]. Historical data solely from EO/MEG facilities is unfortunately not available.

11.4. Comparison

A comparison of the costs of this design to designs licensed by Scientific Design is presented in table 11.3. The Scientific Design EO process values are based on a production of solely 200,000 t/a EO[ref 35] and the Scientific Design EO/MEG process values are based on a production of solely 400,000 t/a of MEG[ref 36]. The values of this design are based on a production of 100,000 t/a EO and 200,000 t/a MEG. It seems strange that this design consumes 28 kt/a of steam and both Scientific Design processes ~~produce~~ steam. The reason for this is that the selectivity of both Scientific Design processes is approximately 0.72 mol EO/mol ethylene, and this design has a selectivity of 86%. The heat generated by the combustion of the extra ethylene produces approximately 375 kt of steam per annum. This leaves a difference of approximately 75 kt/a between the EO/MEG plant of this design and the Scientific Design EO and MEG designs – that is, all three processes produce an excess of steam, but the SD designs produce more. A possible explanation is that the design described in this document is based on the production of both EO and MEG, while the designs it is compared ~~to~~ with produce either EO or MEG.

Table 11.3 Comparison of costing data

Category	This design	Scientific Design EO process	Scientific Design EO/MEG process
Production	100 kt/a EO 200 kt/a MEG	200 kt/a EO	400 kt/a MEG
PCE	29.3 mil. USD	39.9 mil. USD	63.0 mil. USD
Raw materials E + O_2	377.7 kt/a	359 kt/a	523 kt/a
Electric power	69.8 $*10^6$ kWh	31.9 $*10^6$ kWh	85.7 $*10^6$ kWh
Cooling + chilled water	79.9 mil. t/a	65.6 mil. t/a	95.7 mil. t/a
High Pressure Steam	280 kt/a	- 197 kt/a (EOR)	-169 kt/a (EOR)

11.4.1.Operating costs

The costs of producing EO and MEG can be divided into two groups:
Fixed operating costs, costs that do not vary with the production rate
Variable operating costs, costs that do vary with the production rate

Fixed costs include

- Maintenance (labor and materials)
- Operating labor
- Laboratory costs
- Supervision
- Plant overheads
- Capital charges
- Rates (and other local taxes)
- Insurance
- License fees and royalty payments

Variable costs include

- Raw materials
- Utilities
- Shipping & packaging
- Miscellaneous operating materials

Raw material and utility costs are easily calculated from flow sheet data. The only problem occurs with the catalyst costs. The costs listed in table 11.4 are costs of fresh catalyst per year, although the used silver as well as the ion exchange resin MSA-1 can probably be easily regenerated. The costs of this regeneration is unknown and therefore this option is not taken into account.
Miscellaneous materials are materials required for operation of the plant but not included in maintenance and raw materials, such as safety clothing, gaskets, etc. 10% of the maintenance costs is used as estimate for these costs.

Shipping and packaging is not a large factor in the EO/MEG plant; in this design they are neglected.
For the operating labor only personnel directly involved with plant operation is taken into account. Plant personnel costs approximately $3000 per person per month, 20 persons per shift with 5 shifts a day then comes to the total operating labor.
The other fixed costs are calculated as a percentage of the fixed capital. Historical data was again used to determine those percentages.

Table 11.4 Operating costs

Fixed capital (FC)					**k$**	**138,121**	
Raw materials	price		use				
Ethylene	520.00	$/ton	193,198	ton	k$	100,463	✓
Oxygen	50.00	$/ton	184,584	ton	k$	9,229	
Methane	120.00	$/ton	11,592	ton	k$	1,391	
Methylene chloride	750.00	$/ton	12	ton	k$	9	
Water	2.50	$/ton	65,143	ton	k$	163	
Catalyst EO reaction, alumina carrier	10352.94	$/m3	42	m3	k$	435	
Silver for catalyst	5.18	$/ounce	176,056	ounce	k$	912	
Catalyst based on DOW MSA-1	11764.71	$/m3	327	m3	k$	3,843	
Total raw materials					**k$**	**116,445**	
Utilities							
Chilled water	0.21	$/ton			k$	5,323	
Cooling water	0.02	$/ton			k$	1,090	
Steam	9.28	$/ton			k$	2,595	✓
Electricity	0.04	$/kWh	69,823,689	kWh	k$	2,793	
Total Utilities					**k$**	**11,801**	
Miscellaneous materials	10% of maintenance				k$	1,381	
Maintenance	10% of fixed capital				k$	13,812	
Operating labour					k$	2,700	
Laboratory costs	20% of operating labor				k$	540	
Supervision	20% of operating labor				k$	540	
Plant overheads	50% of operating labor				k$	1,350	
Capital charges	15% of fixed capital				k$	20,718	
Insurance	1% of fixed capital				k$	1,381	
Local taxes	2% of fixed capital				k$	2,762	
Royalties	1% of fixed capital				k$	1,381	
Annual production costs					**k$**	**174,812**	✓

11.5. Profitability Measures

Several profitability measures have been calculated for the EO/MEG design. In order to do this a few assumptions are made. The total investment is the earlier calculated fixed capital. It is assumed that this will be paid at the beginning of plant operation, e.g. the end of year 0.
When plant life is taken to be 15 years, and the time value of money discount rate 10%, the net present worth of future earnings is 400 million dollar, or 2.9 times the original investment and the payback time is just under 2.5 years.
When a plant life of 15 years is assumed the discounted cash flow rate of return (DCFRR) is 51%. When a plant-life of 10 years is assumed the DCFRR is 38%. The

DCFRR is a measure of the maximum rate that the project could pay and still break even at the end of the project life. The DCFRR provides a useful way of comparing the performance of capital for different projects, independent of the amount of capital used.

Table 11.5 Summary and data for profitability measures

End of Year	Investment	Operating costs	Forecast sales	Net cash flow	Cumulative cash flow (Net Future Worth)	Discounted cash flow at 10%	Cumulative discounted cash flow (Net Present Worth)
0	k$ 138,121	k$ -	k$ -	k$ - 138,121	k$ - 138,121	k$ - 138,121	k$ - 138,121
1	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ - 67,366	k$ 64,323	k$ - 73,799
2	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 3,389	k$ 58,475	k$ - 15,323
3	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 74,144	k$ 53,159	k$ 37,836
4	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 144,899	k$ 48,327	k$ 86,163
5	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 215,654	k$ 43,933	k$ 130,096
6	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 286,409	k$ 39,939	k$ 170,035
7	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 357,164	k$ 36,309	k$ 206,344
8	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 427,919	k$ 33,008	k$ 239,352
9	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 498,674	k$ 30,007	k$ 269,359
10	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 569,429	k$ 27,279	k$ 296,638
11	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 640,185	k$ 24,799	k$ 321,437
12	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 710,940	k$ 22,545	k$ 343,982
13	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 781,695	k$ 20,495	k$ 364,477
14	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 852,450	k$ 18,632	k$ 383,109
15	k$ -	k$ 174,812	k$ 245,567	k$ 70,755	k$ 923,205	k$ 16,938	k$ 400,047

Table 11.6 Financial scenarios for the EO/MEG process

End of Year	Project NPW at 55% discount rate (15 years)	Project NPW at 51% discount rate (15 years)	Project NPW at 45% discount rate (15 years)	Project NPW at 40% discount rate (10 years)	Project NPW at 38% discount rate (10 years)	Project NPW at 35% discount rate (10 years)
0	k$ - 138,121	k$ - 138,121	k$ - 138,121	k$ - 138,121	k$ - 138,121	k$ - 138,121
1	k$ - 92,473	k$ - 91,264	k$ - 89,325	k$ - 87,582	k$ - 86,850	k$ - 85,710
2	k$ - 63,022	k$ - 60,232	k$ - 55,672	k$ - 56,373	k$ - 54,110	k$ - 50,502
3	k$ - 44,022	k$ - 39,682	k$ - 32,463	k$ - 37,237	k$ - 33,309	k$ - 26,914
4	k$ - 31,764	k$ - 26,072	k$ - 16,457	k$ - 25,604	k$ - 20,172	k$ - 11,161
5	k$ - 23,855	k$ - 17,059	k$ - 5,418	k$ - 18,608	k$ - 11,935	k$ - 678
6	k$ - 18,753	k$ - 11,090	k$ 2,195	k$ - 14,458	k$ - 6,814	k$ 6,270
7	k$ - 15,461	k$ - 7,137	k$ 7,445	k$ - 12,041	k$ - 3,666	k$ 10,853
8	k$ - 13,337	k$ - 4,519	k$ 11,066	k$ - 10,667	k$ - 1,758	k$ 13,858
9	k$ - 11,967	k$ - 2,785	k$ 13,563	k$ - 9,913	k$ - 621	k$ 15,816
10	k$ - 11,083	k$ - 1,637	k$ 15,285	k$ - 9,521	k$ 39	k$ 17,082
11	k$ - 10,513	k$ - 877	k$ 16,473			
12	k$ - 10,145	k$ - 373	k$ 17,292			
13	k$ - 9,907	k$ - 40	k$ 17,857			
14	k$ - 9,754	k$ 181	k$ 18,246			
15	k$ - 9,655	k$ 327	k$ 18,515			

11.6. Sensitivity analysis

As can be seen in table 11.6, using current EO and MEG prices the EO/MEG process is economically viable. However, it is based on current EO/MEG prices and estimated EO/MEG production rates. Without going too deeply into the matter, possible future scenarios are disappointing production rates, decrease of sales prices, increase of raw material prices, etc.
The cash flow is a very important measure of profitability. Its calculation is also very easily made. Therefore in the following analysis this is taken as the main measure for profitability.

Table 11.7 Sensitivity analysis (only ethylene, oxygen and methane have been considered as raw materials)

	OPEX		Sales		Cashflow	
Normal case	k$	174,812	k$	245,567	k$	70,755
Production decrease of 10%	k$	174,812	k$	221,010	k$	46,198
Sales price decrease of 10%	k$	174,812	k$	221,010	k$	46,198
Sales price and production decrease of 10%	k$	174,812	k$	198,909	k$	24,097
Raw material price increase of 10%	k$	185,920	k$	245,567	k$	59,647
Sales price and production decrease of 10% and a raw material price increase of 10%	k$	185,920	k$	198,909	k$	12,989

In all cases the cash flow remains positive, not accounting for taxes, costs of loans, etc. This means the process is still making money. The process is thus very stable.

12. Conclusions and Recommendations

This document presents the conceptual design of an EO/MEG manufacturing plant, producing 100,000 t/a direct sales specification EO and 200,000 t/a fiber grade MEG. The design must be economically and technologically feasible, based on state-of-the-art technology, while not discarding the value of proven technology.
After literature research, a Shell/CRI based EO production plant and a Mitsubishi 2-step MEG production plant were selected. The most important advantages of such a design are high selectivity and minimal energy consumption. The absence of water as absorbent is responsible for effective energy usage, while the 2-step MEG production without large excesses of water ensures higher selectivity than conventional processes. Anti-freeze grade MEG or higher order ethylene glycols are not produced in any significant amount. At first sight, this design seems to consume a substantial amount of steam compared to other designs by Scientific Design which produce a large amount of steam. This design was expected to produce steam as well. This difference in steam production is due to a much lower, 72% in the SD designs vs. 86% of this design, selectivity and thus more ethylene combustion of those designs.

The deliverables, a flow sheet simulation, mass and heat balances, major equipment sizing and cost estimation, and limited safety analysis, are all complete. The flow sheet diagram is included in appendix 1, mass and heat balances are presented in chapter 7, and the plant economics in chapter 11.

The fixed capital calculated is 138 million USD. Annual raw material costs are 116 million USD. Annual utility costs are 12 million USD. Total annual production costs are 175 million USD. Income from the sales of EO and MEG products is 246 million USD annually. The annual margin is thus 71 million USD; this results in a payback time of just under 2.5 years, a sound economic plan. A stability analysis shows that possible fluctuations in prices and production do not have lethal influence on the economic prospects of this design.

Most information contained in this conceptual design is based either on basic engineering insight, or information available in the public domain, notably patent literature for the manufacturing processes. Although such literature and insight may occasionally not be wholly true, much effort has been made to verify the information that has any significance for this design. Therefore, the designers feel that the concepts herein laid out are a good basis for a detailed design.

13. List of Abbreviations

BFW	Boiler feed water
BOD	Basis of design
CAPEX	Capital expenses
CW	Cooling water
CHW	Chilled water
DCFRR	Discounted cash flow rate of return
DEG	Diethylene glycol
EC	Ethylene carbonate
EG	Ethylene glycol
EO	Ethylene oxide
EP	End point
F&EI	The Dow Fire and Explosion Index
HPS	High pressure steam
HSS	High selection switch
IBP	Initial boiling point
LC	Level controller
LPS	Low pressure steam
MAINTEX	Maintenance expenses
MEG	Monoethylene glycol
Mmt	Mega metric tons
MPS	Medium pressure steam
Mt/a	Millions of metric tons/year
NFW	Net future worth
NPW	Net present worth
OPEX	Operational expenses
OSBL	Outside battery limit
PC	Pressure controller
PC	Propylene carbonate
POT	Pay out time
PRV	Pressure relieve valve
PVDF	Polyvinylidenedifluoride
PFD	Process flow diagram
RC	Ratio controller
ROI	Return on investment
ROR	Rate of return
t/a	Metric tons/year
TC	Temperature controller
TEG	Triethylene glycol

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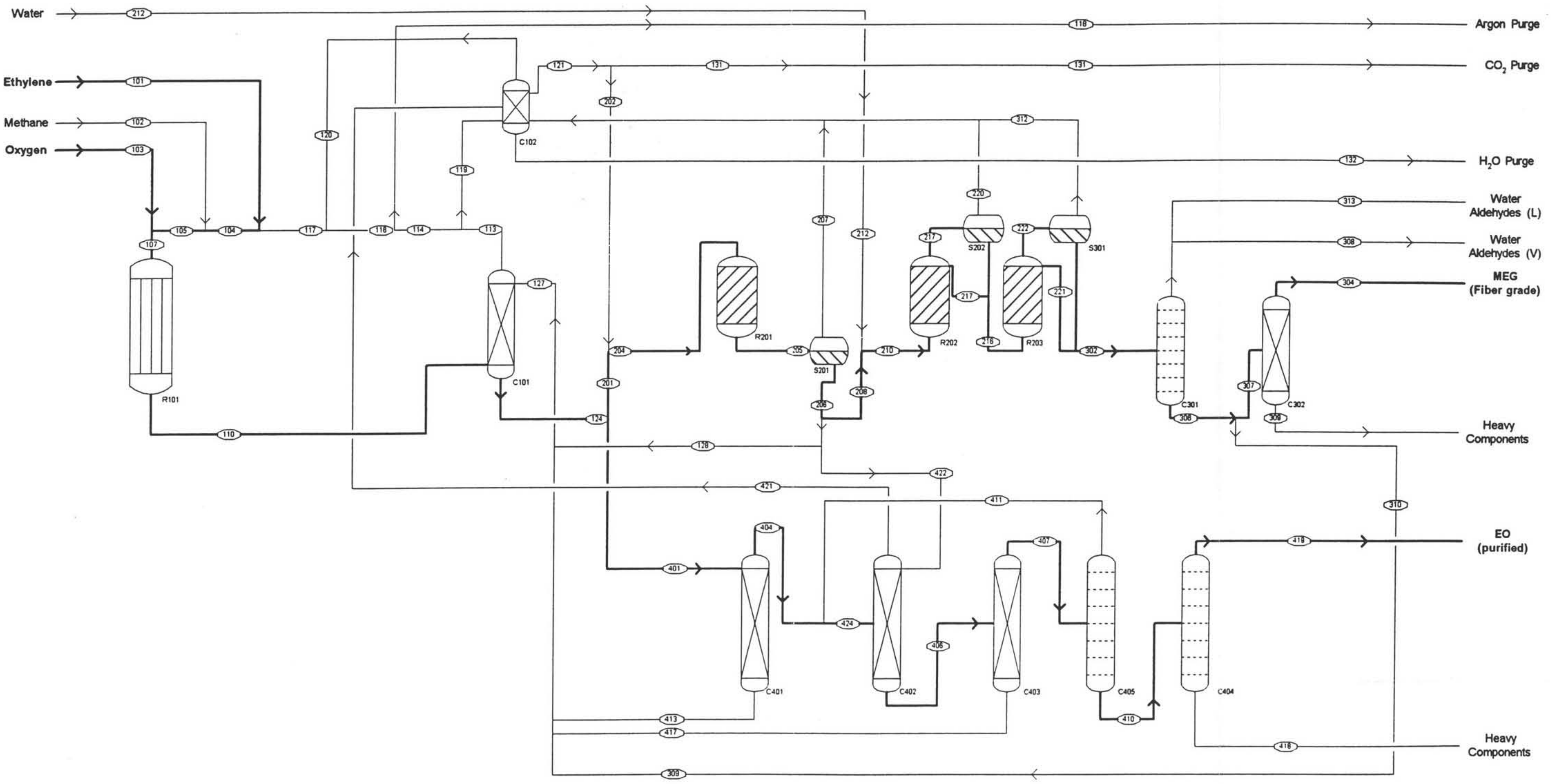
Ethylene Oxide and Ethylene Glycol from Ethylene

Process Unit Block Scheme

V3.00



Unit Operation summary

C101: EO Absorber
C102: CO$_2$ Removal system
C301: Water Removal
C302: MEG Purification
C401: Raw EO Stripper
C402: Sec. EO Absorber

C403: Sec. EO Stripper
C404: Heavy end Seperator
C405: Light end Seperator
R101: EO Reactor
R201: Carbonate Reactor
R202: Pri. Meg Reactor
R203: Sec. Meg Reactor

S201: CO2 Flashdrum
S202: Flashdrum
S301: CO$_2$ Flashdrum

Designers

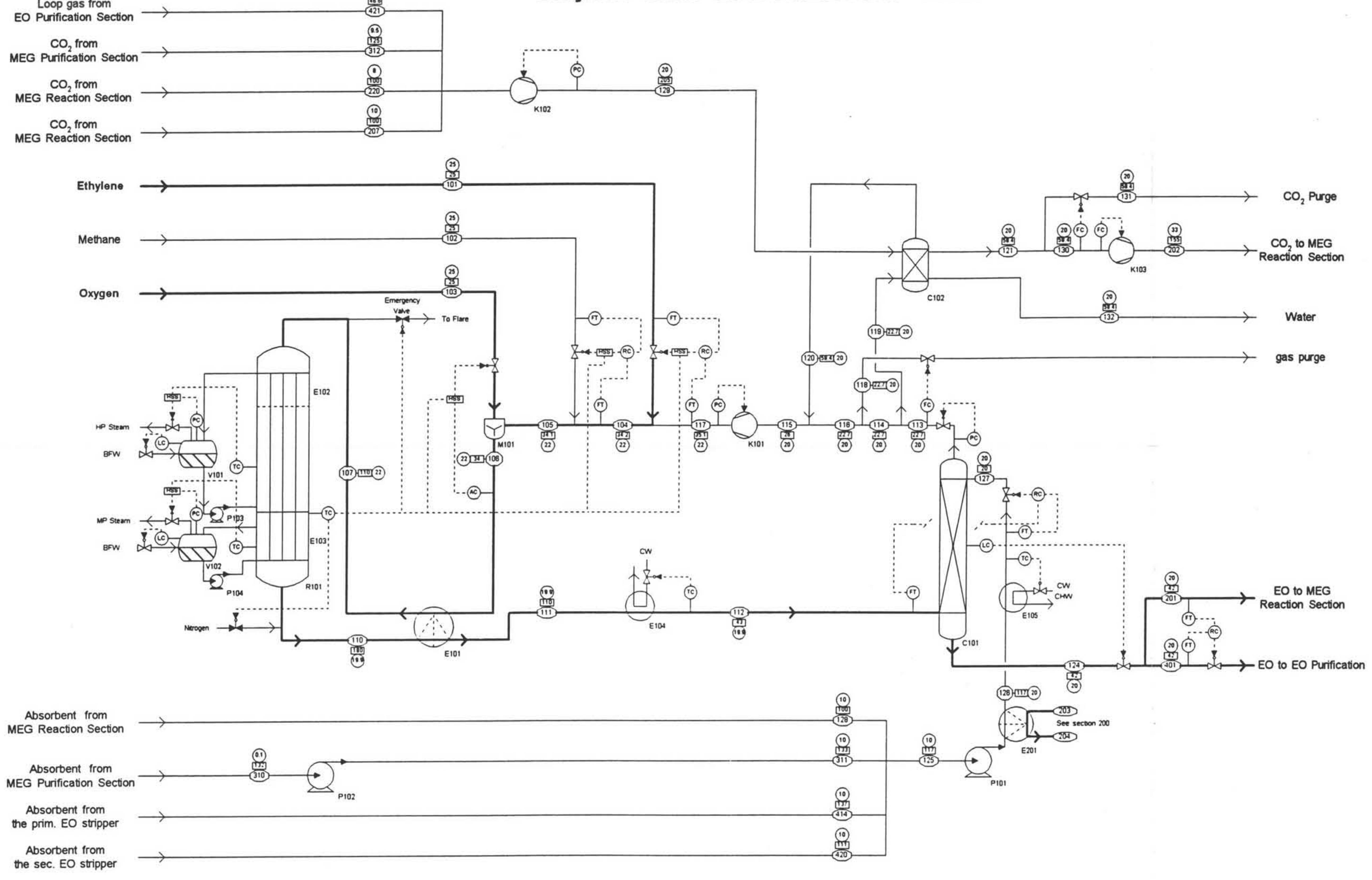
Frank Sheldon
Flip Kleyn van Willigen
Raoul Boucke
Menno Sorgedrager

Process Unit Block Scheme

Project:	Integrated MEG and EO production from ethylene and purified oxygen
Project ID:	**CPD 3244** Internal project ID: 80511 CA30
Completion date:	January 1st 2000
Stream number:	201

Ethylene Oxide and Ethylene Glycol from Ethylene

Ethylene Oxide Reaction Section -U100



Unit Operation summary		
C101: EO Absorber	K101: Loop Gas Compressor	P103: Coolant Pump
C102: CO₂ Removal system	K102: Recycle Compressor	P104: Coolant Pump
E101: Feed Preheater	K103: CO₂ Compressor	R101: EO Reactor
E102: Feed Heater	M101: Oxygen Mixing Station	V101: Steam drum
E103: Quench	P101: Main Absorbent Pump	V102: Steam drum
E104: Effluent Cooler	P102: Absorbent Pump	
E105: Absorbent Cooler		

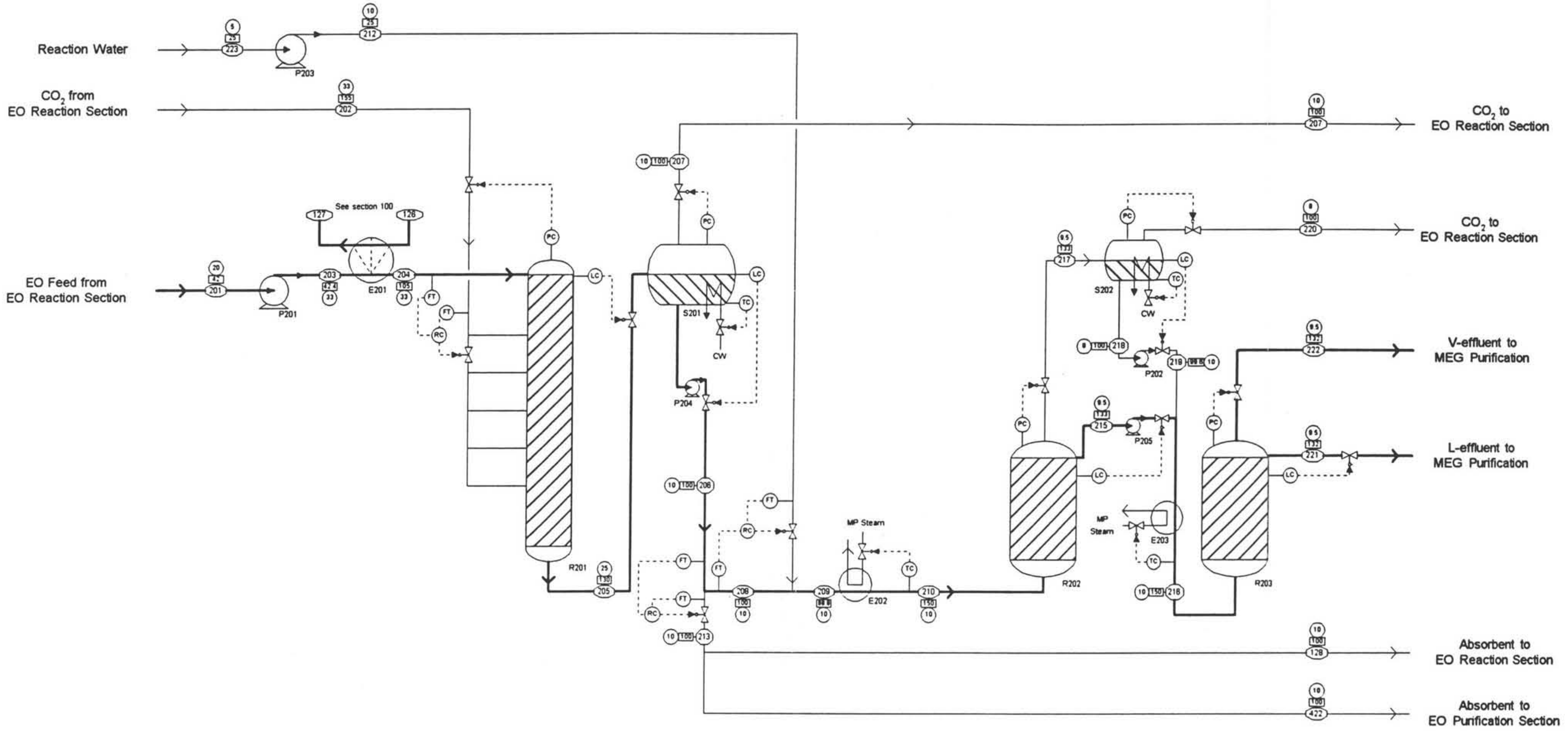
Abbreviations	Designers
TC: Temperature controler	Frank Sheldon
PC: Pressure controler	Flip Kleyn van Willigen
RC: Ratio controler	Raoul Boucke
FC: Flow controler	Menno Sorgedrager
FT: Flow measure/transmit	
HSS: High selection switch	
CW: Cooling water	
CHW: Chilled water	
BFW: Boiler feed water	

Process Flow Scheme

Project:	Integrated MEG and EO production from ethylene and purified oxygen
Section:	Ethylene Oxide Reaction Section U-100
Project ID:	**CPD 3244 / P80511 CA30**
Completion date:	January 1st 2000
Stream number:	201 Temperature (C): 20 Pressure (Bar): ③

Ethylene Oxide and Ethylene Glycol from Ethylene

MEG Reaction Section - U200



Unit Operation summary

E201: Preheater R201
E202: Preheater R202
E203: Preheater R203

P201: EO Feed Pump
P202: Liquid phase Pump S202
P203: Water Pump
P204: Liquid phase Pump S201
P205: Liquid phase Pump R202

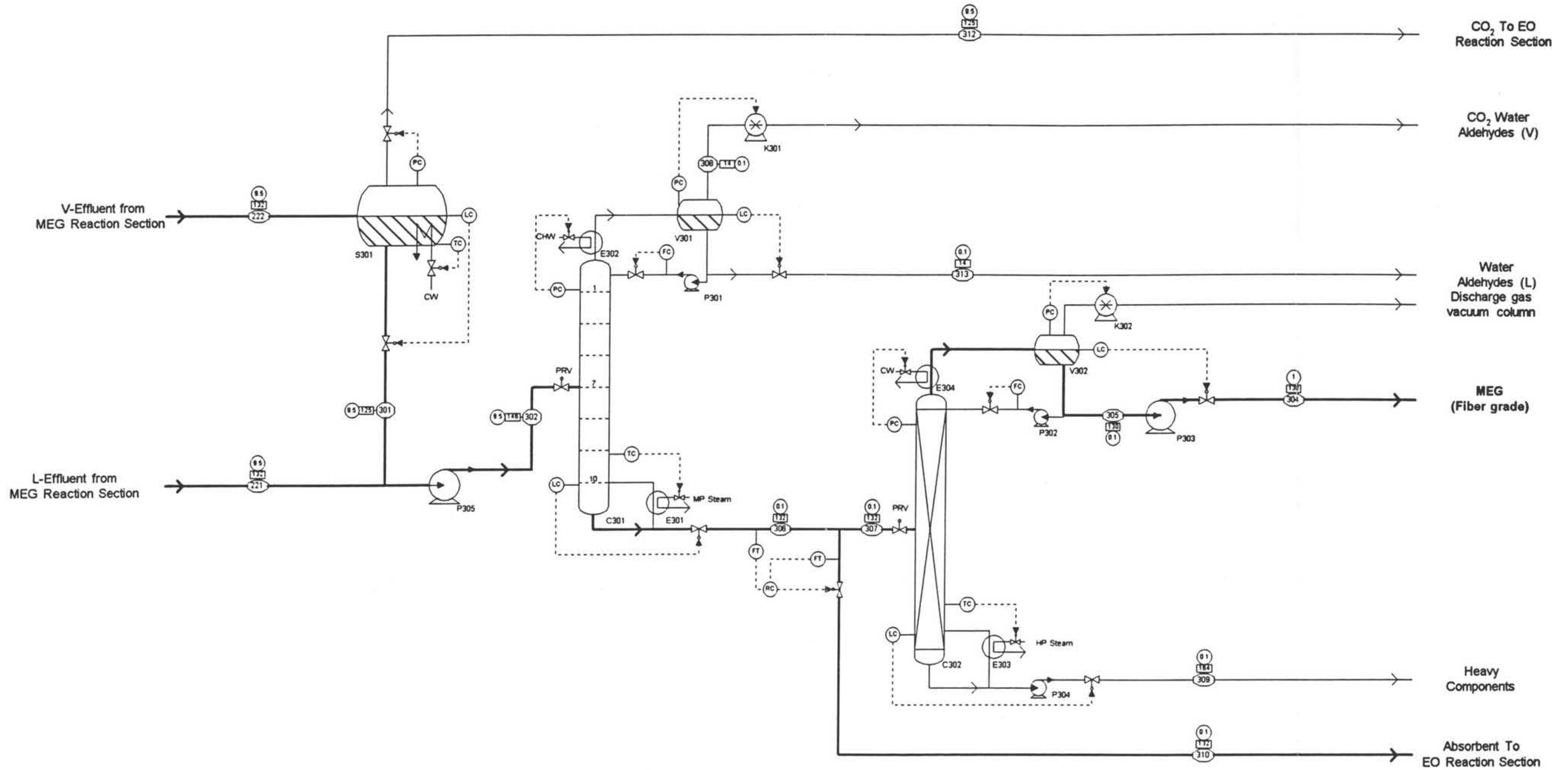
R201: Carbonate Reactor
R202: Pri. MEG Reactor
R203: Sec. MEG Reactor
S201: CO2 Flashdrum
S202: Flashdrum

Abbreviations	Designers
TC: Temperature controler PC: Pressure controler RC: Ratio controler FC: Flow controler FT: Flow measure/transmit HSS: High selection swtich CW: Cooling water CHW: Chilled water BFW: Boiler feed water	Frank Sheldon Flip Kleyn van Willigen Raoul Boucke Menno Sorgedrager

Process Flow Scheme

Project:	Integrated MEG and EO production from ethylene and purified oxygen		
Section:	MEG Reaction Section U200		
Project ID:	**CPD 3244 / P80511 CA30**		
Completion date:	January 1st 2000		
Stream number:	201	Temperature (C): 20	Pressure (Bar): ③

Ethylene Oxide and Ethylene Glycol from Ethylene
MEG Purification Section - U300



Unit Operation summary

C301: Water Removal
C302: MEG Purification
E301: Reboiler C301
E302: Condenser C301
E303: Reboiler C302
E304: Condenser C302

K301: Vacuum compressor C301
K302: Vacuum compressor C302
P301: Reflux pump C301
P302: Reflux pump C302
P303: MEG Pump
P304: Heavy component Pump
P305: Raw MEG Pump

S301: CO2 Flashdrum
V301: V/L Seperator C301
V302: V/L Seperator C302

Abbreviations

TC: Temperature controler
PC: Pressure controler
RC: Ratio controler
FC: Flow controler
FT: Flow measure/transmit
HSS: High selection switch
CW: Cooling water
CHW: Chilled water
BFW: Boiler feed water

Designers

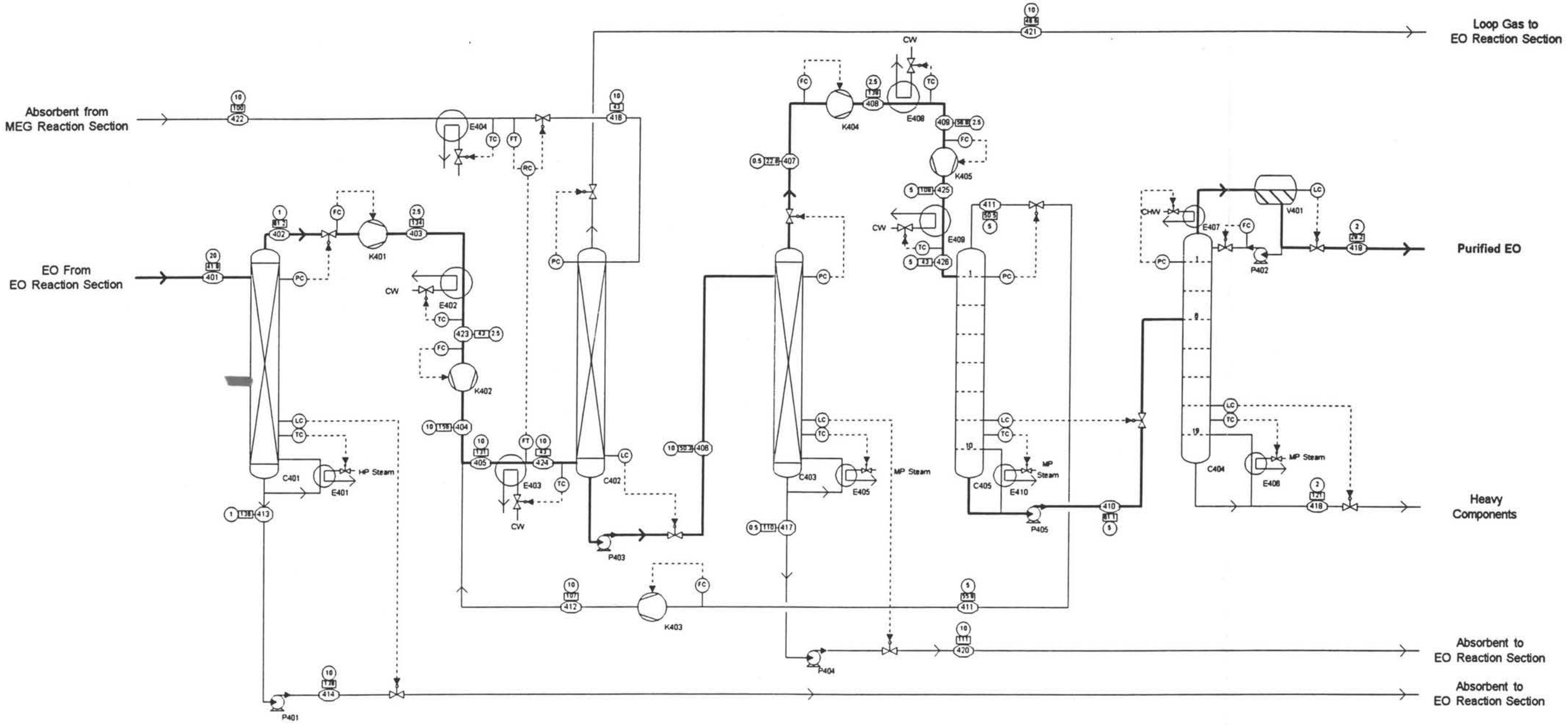
Frank Sheldon
Flip Kleyn van Willigen
Raoul Boucke
Menno Sorgedrager

Process Flow Scheme

Project:	Integrated MEG and EO production from ethylene and purified oxygen
Section:	MEG Purification Section U300
Project ID:	**CPD 3244 / P80511 CA30**
Completion date:	January 1st 2000

Stream number: 201 Temperature (C): 20 Pressure (Bar): ③

Ethylene Oxide and Ethylene Glycol from Ethylene
EO Purification Section - U 400



Unit Operation summary

C401: Raw EO Stripper
C402: Sec. EO Absorber
C403: Sec. EO Stripper
C404: Heavy end Seperator
C405: Light end Seperator
E401: Reboiler C401
E402: Post-Cooler K401

E403: Post-Cooler K402/403
E404: Absorbent Cooler
E405: Reboiler C403
E406: Reboiler C404
E407: Codenser C404
E408: Post-Cooler K404
E409: Post-Cooler K405

E410: Reboiler C405
K401: EO Pri. Compressor
K402: EO sec.Compressor
K403: EO Recycle Compressor
K404: Raw EO Pri. Compressor
K405: Raw EO Sec Compressor
P401: Absorbent Pump

P402: Reflux Pump C404
P403: Bottom Pump C402
P404: absorbent Pump
P405: Bottom Pump C405
V401: V/L Seperator C404

Abbreviations

TC: Temperature controler
PC: Pressure controler
RC: Ratio controler
FC: Flow controler
FT: Flow measure/transmit
HSS: High selection switch
CW: Cooling water
CHW: Chilled water
BFW: Boiler feed water

Designers

Frank Sheldon
Flip Kleyn van Willigen
Raoul Boucke
Menno Sorgedrage

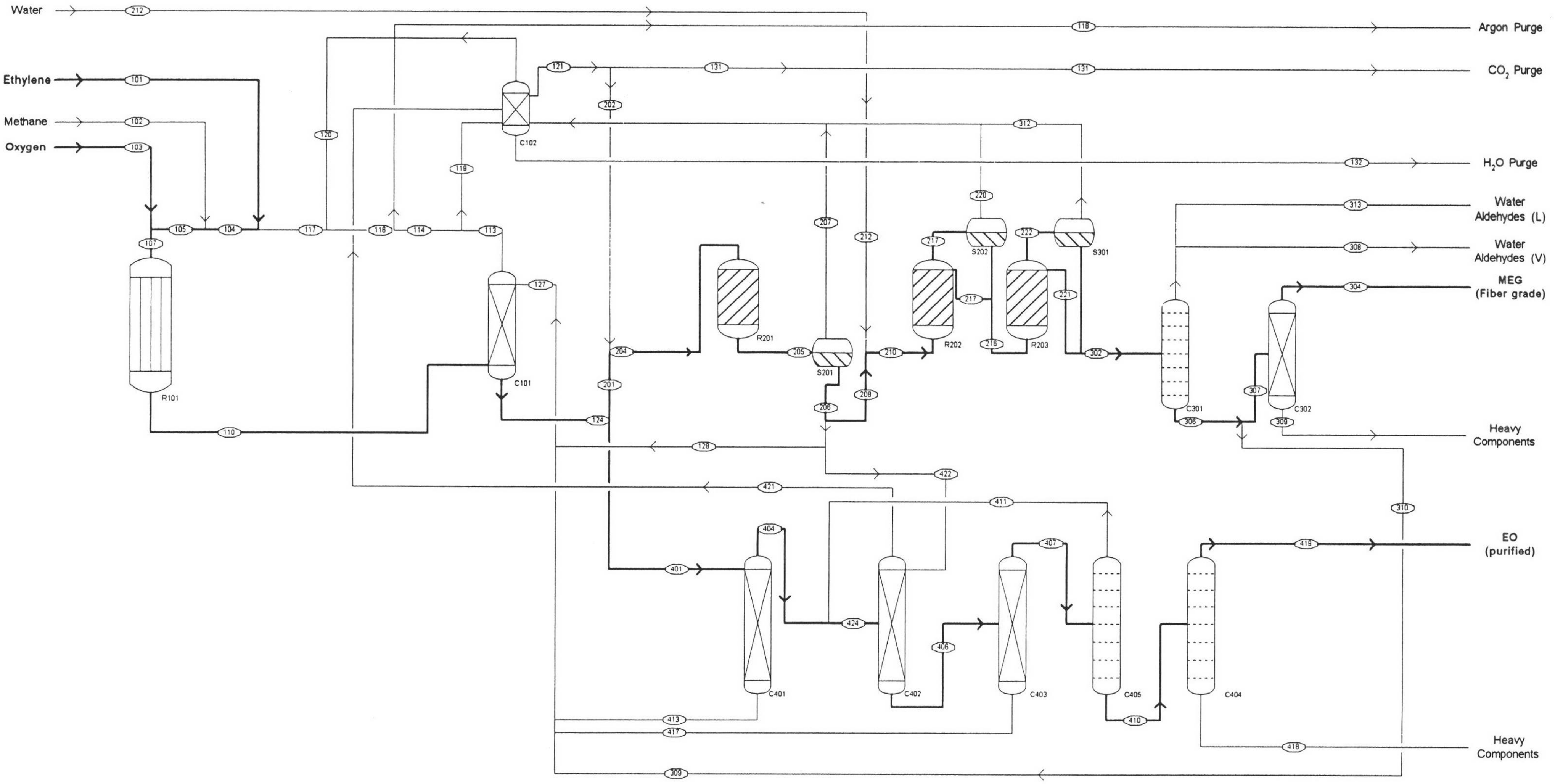
Process Flow Scheme

Project:	Integrated MEG and EO production from ethylene and purified oxygen				
Section:	Ethylene Oxide Purification section - U400				
Project ID:	**CPD 3244 / P80511 CA30**				
Completion date:	January 1st 2000				
Stream number:	201	Temperature (C):	20	Pressure (Bar):	③

Ethylene Oxide and Ethylene Glycol from Ethylene

v3.00

Process Unit Block Scheme



Unit Operation summary

C101: EO Absorber
C102: CO_2 Removal system
C301: Water Removal
C302: MEG Purification
C401: Raw EO Stripper
C402: Sec. EO Absorber
C403: Sec. EO Stripper
C404: Heavy end Seperator
C405: Light end Seperator
R101: EO Reactor
R201: Carbonate Reactor
R202: Pri. Meg Reactor
R203: Sec. Meg Reactor
S201: CO2 Flashdrum
S202: Flashdrum
S301: CO_2 Flashdrum

Designers

Frank Sheldon
Flip Kleyn van Willigen
Raoul Boucke
Menno Sorgedrager

Process Unit Block Scheme

Project:	Integrated MEG and EO production from ethylene and purified oxygen
Project ID:	**CPD 3244** Internal project ID: 80511 CA30
Completion date:	January 1st 2000
Stream number:	201

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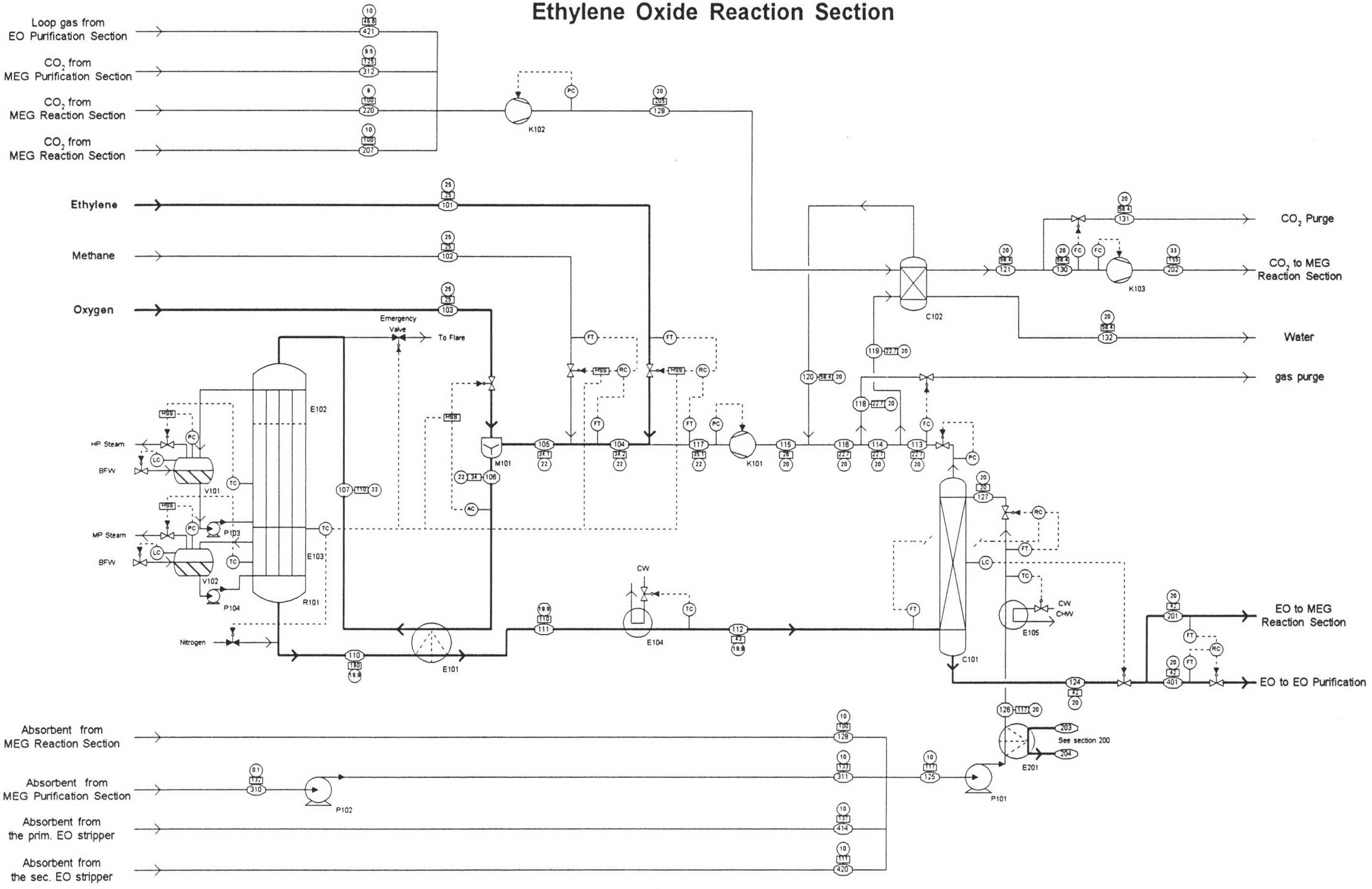
Ethylene Oxide and Ethylene Glycol from Ethylene

Ethylene Oxide Reaction Section



Unit Operation summary		
C101: EO Absorber	K101: Loop Gas Compressor	P103: Coolant Pump
C102: CO_2 Removal system	K102: Recycle Compressor	P104: Coolant Pump
E101: Feed Preheater	K103: CO_2 Compressor	R101: EO Reactor
E102: Feed Heater	M101: Oxygen Mixing Station	V101: Steam drum
E103: Quench	P101: Main Absorbent Pump	V102: Steam drum
E104: Effluent Cooler	P102: Absorbent Pump	
E105: Absorbent Cooler		

Abbreviations	Designers
TC: Temperature controler	Frank Sheldon
PC: Pressure controler	Flip Kleyn van Willigen
RC: Ratio controler	Raoul Boucke
FC: Flow controler	Menno Sorgedrager
FT: Flow measure/transmit	
HSS: High selection switch	
CW: Cooling water	
CHW: Chilled water	
BFW: Boiler feed water	

Process Flow Scheme

Project: Integrated MEG and EO production from ethylene and purified oxygen
Section: Ethylene Oxide Reaction Section
Project ID: **CPD 3244 / P80511 CA30**
Completion date: January 1st 2000
Stream number: (201) Temperature (C): [20] Pressure (Bar): ③

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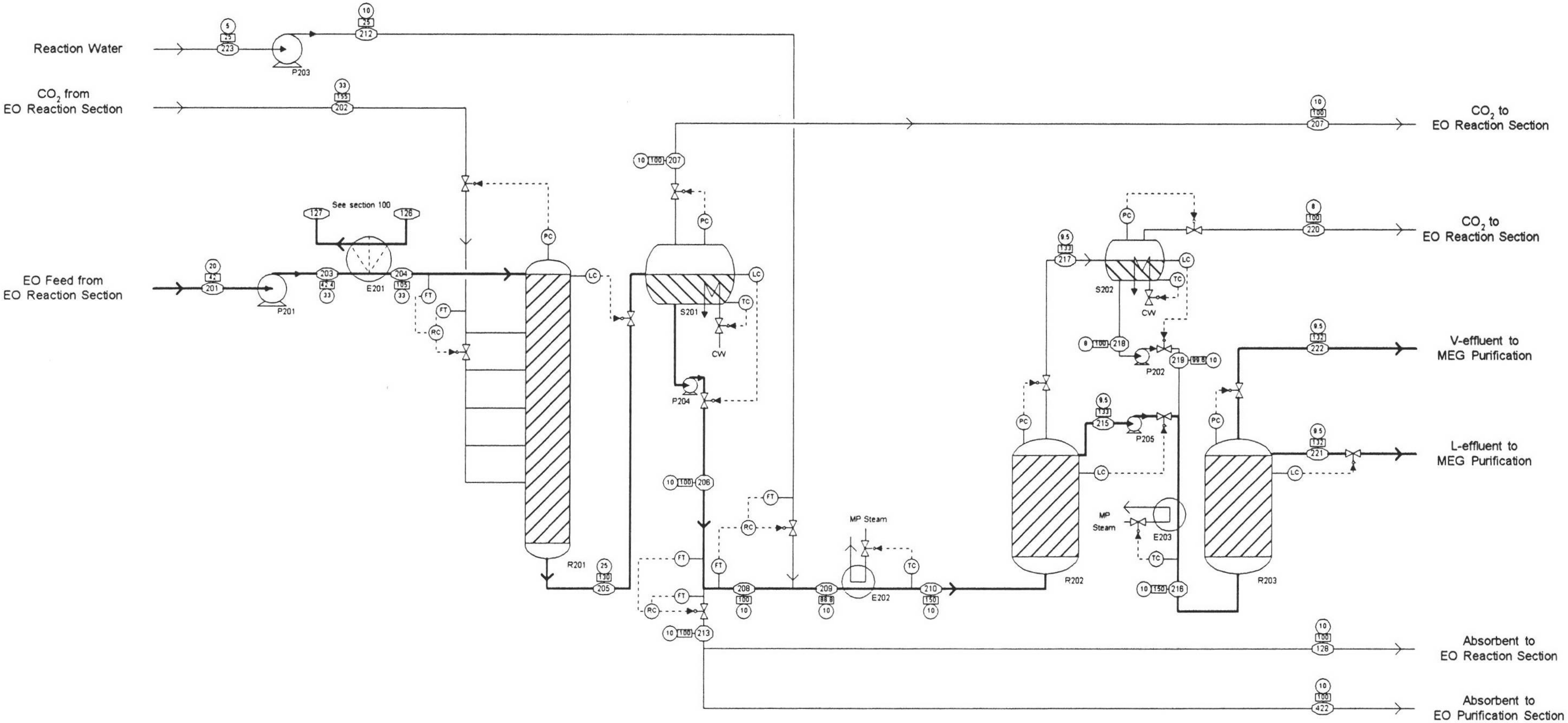
Ethylene Oxide and Ethylene Glycol from Ethylene

MEG Reaction Section



Unit Operation summary		
E201: Preheater R201 E202: Preheater R202 E203: Preheater R203	P201: EO Feed Pump P202: Liquid phase Pump S202 P203: Water Pump P204: Liquid phase Pump S201 P205: Liquid phase Pump R202	R201: Carbonate Reactor R202: Pri. MEG Reactor R203: Sec. MEG Reactor S201: CO2 Flashdrum S202: Flashdrum

Abbreviations	Designers
TC: Temperature controler PC: Pressure controler RC: Ratio controler FC: Flow controler FT: Flow measure/transmit HSS: High selection switch CW: Cooling water CHW: Chilled water BFW: Boiler feed water	Frank Sheldon Flip Kleyn van Willigen Raoul Boucke Menno Sorgedrager

Process Flow Scheme			
Project:	Integrated MEG and EO production from ethylene and purified oxygen		
Section:	MEG Reaction Section		
Project ID:	**CPD 3244 / P80511 CA30**		
Completion date:	January 1st 2000		
Stream number:	201	Temperature (C): 20	Pressure (Bar): 3

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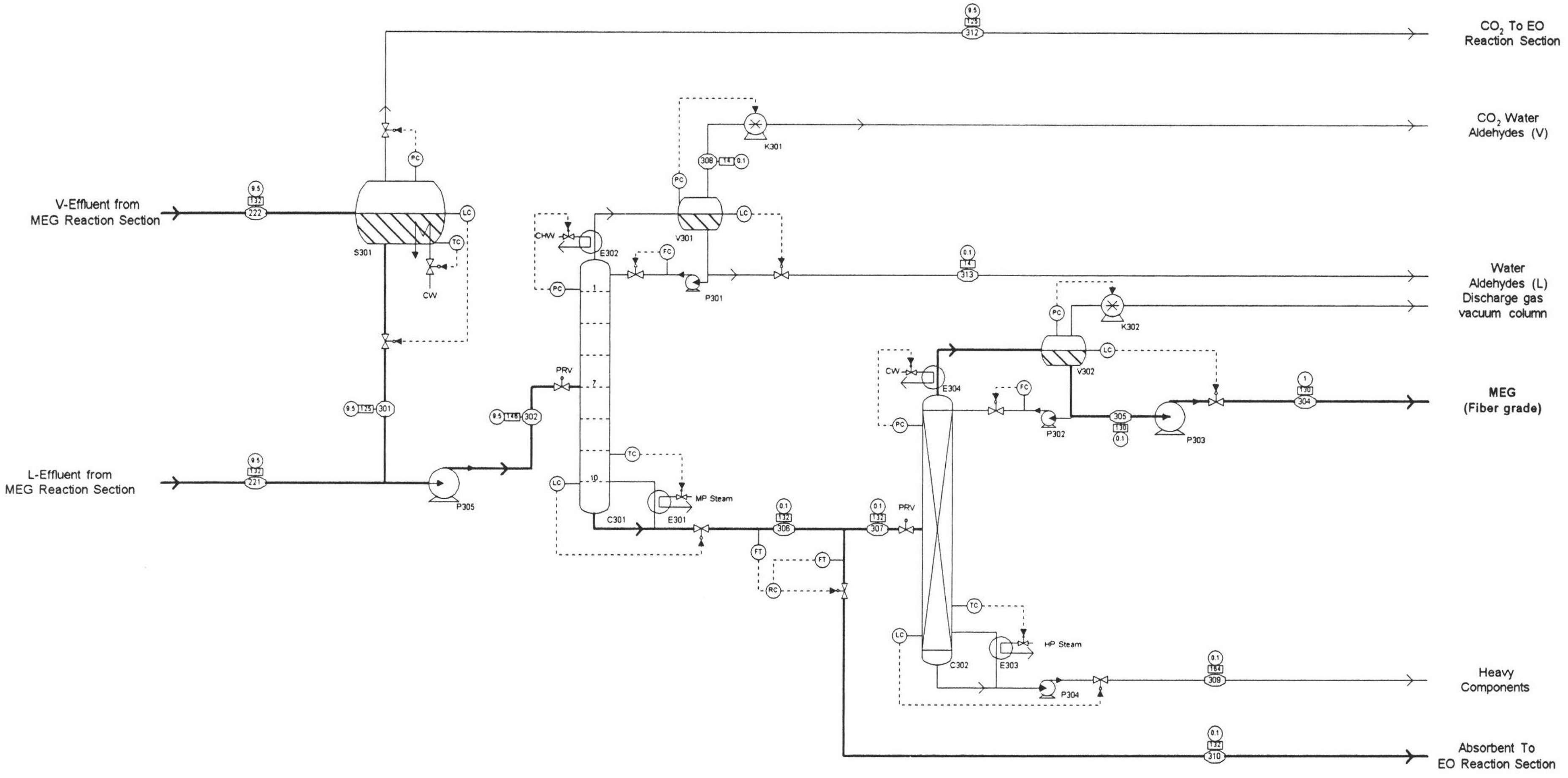
Ethylene Oxide and Ethylene Glycol from Ethylene

MEG Purification Section



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Unit Operation summary

C301: Water Removal
C302: MEG Purification
E301: Reboiler C301
E302: Condenser C301
E303: Reboiler C302
E304: Condenser C302
K301: Vacuum compressor C301
K302: Vacuum compressor C302
P301: Reflux pump C301
P302: Reflux pump C302
P303: MEG Pump
P304: Heavy component Pump
P305: Raw MEG Pump
S301: CO2 Flashdrum
V301: V/L Seperator C301
V302: V/L Seperator C302

Abbreviations

TC: Temperature controler
PC: Pressure controler
RC: Ratio controler
FC: Flow controler
FT: Flow measure/transmit
HSS: High selection switch
CW: Cooling water
CHW: Chilled water
BFW: Boiler feed water

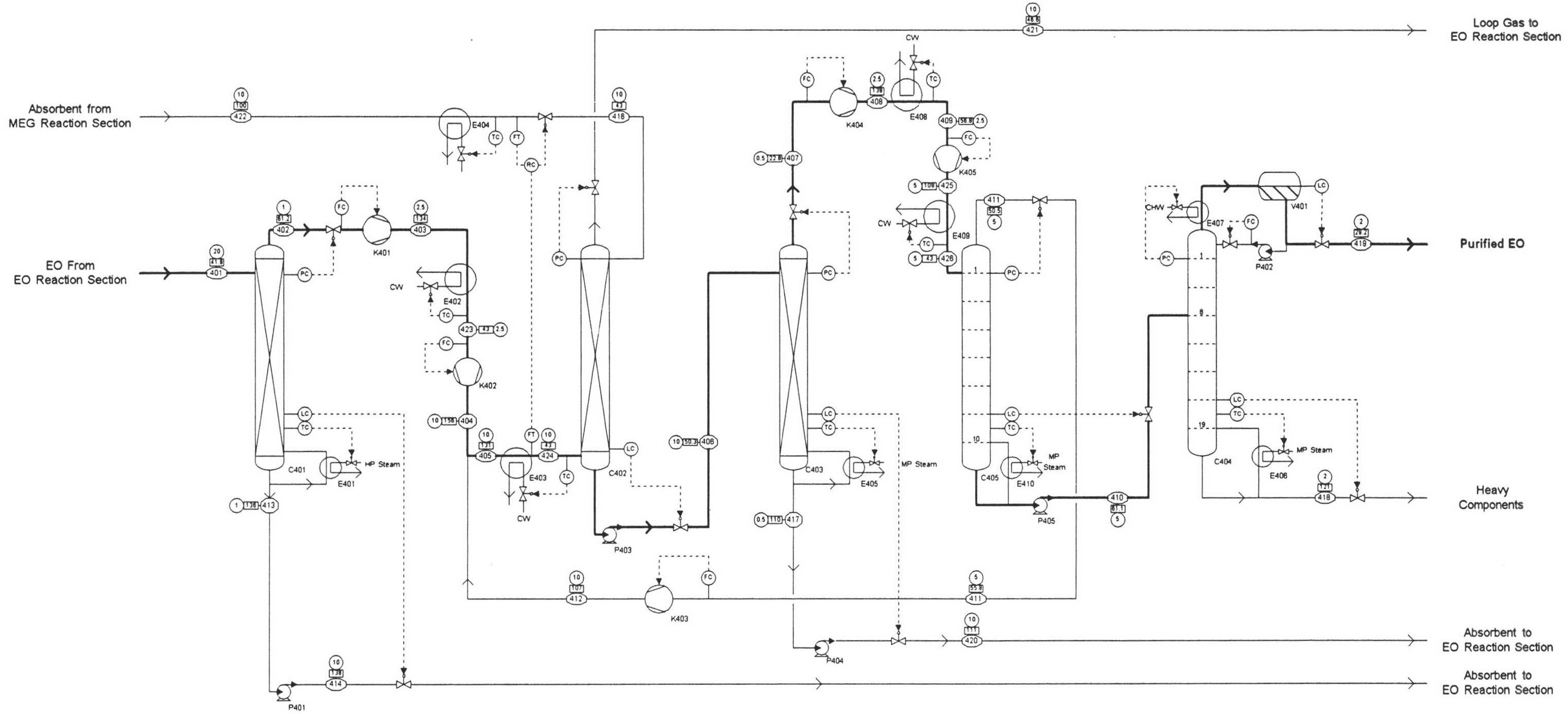
Designers

Frank Sheldon
Flip Kleyn van Willigen
Raoul Boucke
Menno Sorgedrager

Process Flow Scheme

Project: Integrated MEG and EO production from ethylene and purified oxygen
Section: MEG Purification Section
Project ID: **CPD 3244 / P80511 CA30**
Completion date: January 1st 2000
Stream number: 201 Temperature (C): 20 Pressure (Bar): ③

Ethylene Oxide and Ethylene Glycol from Ethylene
EO Purification Section



Unit Operation summary

C401: Raw EO Stripper
C402: Sec. EO Absorber
C403: Sec. EO Stripper
C404: Heavy end Seperator
C405: Light end Seperator
E401: Reboiler C401
E402: Post-Cooler K401
E403: Post-Cooler K402/403
E404: Absorbent Cooler
E405: Reboiler C403
E406: Reboiler C404
E407: Codenser C404
E408: Post-Cooler K404
E409: Post-Cooler K405
E410: Reboiler C405
K401: EO Pri. Compressor
K402: EO sec.Compressor
K403: EO Recycle Compressor
K404: Raw EO Pri. Compressor
K405: Raw EO Sec Compressor
P401: Absorbent Pump
P402: Reflux Pump C404
P403: Bottom Pump C402
P404: absorbent Pump
P405: Bottom Pump C405
V401: V/L Seperator C404

Abbreviations

TC: Temperature controler
PC: Pressure controler
RC: Ratio controler
FC: Flow controler
FT: Flow measure/transmit
HSS: High selection switch
CW: Cooling water
CHW: Chilled water
BFW: Boiler feed water

Designers

Frank Sheldon
Flip Kleyn van Willigen
Raoul Boucke
Menno Sorgedrage

Process Flow Scheme

Project: Integrated MEG and EO production from ethylene and purified oxygen
Section: Ethylene Oxide Purification section
Project ID: **CPD 3244 / P80511 CA30**
Completion date: January 1st 2000
Stream number: (201) Temperature (C): 20 Pressure (Bar): ③

Appendix 2 - FEEDSTOCK SPECIFICATIONS

- **Ethylene**

Units Spec!

Source: Pipeline from OSBL	
Pressure at battery limits (kg/cm^{2}g min)	25
Price, US$/MT	520

Analysis

ethylene, vol% min.	99.9
Ethane, vol% max.	0.05
Hydrogen, vol ppm max.	5
C_3^+, vol ppm max.	10
Acetylene, vol ppm max.	5
Carbon monoxide, vol ppm max.	10
Carbon dioxide, vol ppm max.	10
Oxygen, vol ppm max.	1
Sulphur (H_2S), vol ppm max.	1
Methanol, vol ppm max.	10
water, vol ppm max.	2
Chlorine (Cl), vol ppm max.	1

etc

- **Oxygen**

Source: Pipeline OSBL	
Pressure at battery limits (kg/cm^{2}g min)	27
Price, US$/MT	50

Analysis

Oxygen, vol% min.	99.8
Nitrogen + Argon, vol% max.	0.2

- **Methane**

Source: Pipeline OSBL	
Pressure at battery limits (kg/cm^{2}g min)	25
Price, US$/MT	120

Analysis

Methane, vol% min.	90.0
Ethylene, vol% max.	10
Hydrogen, vol% max.	5.0
Ethane, vol% max.	1.0
Carbon monoxide, vol% max.	1.0
C_3^+, vol% max.	0.1
Acetylene, vol ppm max.	100
Sulphur compounds, wt ppm, max.	5
Halogen compounds, vol ppm, max	5

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Appendix 3 - PRODUCT SPECIFICATIONS

- **Fibergrade monoethylene glycol (MEG)**

Market value, US$/MT		650
Appearance		Clear, colorless
Specific gravity at 20/20 °C		1.1151 to 1.1156
MEG content, wt% min.		99.8
DEG content, wt% max.		0.05
Water, wt% max.		0.05
Acidity (as acitic), wt% max.		0.005
Chlorides (Cl), wt% max.		0.1
Iron, wt ppm max.		0.1
Ash, wt% max.		0.001
aldehydes (formaldehyde), wt ppm max.		8
Color, Pt-Co units		
Before heating		5
after 4 hours boiling		10
Destillation range at 760 mm. Hg		
IBP, °C min	197.3°	196.0
EP, °C max.		199.0
Vol% range, °C max.		1.0

- **Ethylene oxide (EO)**

Market value, US$/MT		1080
Appearance		Clear, colorless
Purity, wt% min.		99.99
Color, APHA max.		5
Carbon dioxide, wt% max.		0.001
Water, wt% max.	50 ppm.	0.005
Aldehydes (acetaldehyde), wt ppm max.		10
Acidity, wt ppm max.		20
Residue, wt% max.		0.001

- **Diethylene glycol (DEG)**

Market value, US$/MT	551
Appearance	Clear, colorless
Specific gravity at 20/20°C	1.117 to 1.120
DEG content, wt% min.	99.8
MEG content, wt% max.	0.05
TEG content, wt% max.	0.05
Water, wt% max.	0.1
Acidity (as acetic), wt% max.	0.005
Ash, wt% max.	0.005
Color, Pt-Co units max.	10
Destillation range at 760 mm. Hg	
IBP to DP	4°C including 246°C

Confidential.

Appendix 4 - UTILITY SPECIFICATIONS

- **Steam**

Price, US$/MT	
High pressure (for start up), kg/cm^{2}g min	32
High pressure, kg/cm^{2}g min	25
Medium pressure, kg/cm^{2}g min	14
Low pressure, kg/cm^{2}g min	4

- **Power**

Price, US$/kWh	0.04
1 to 225 kilowatts, volts	415
above 225 kilowatts, volts	6600

- **Cooling water**

Price, US$/M^3	0.02
Supply pressure at grade, kg/cm^{2}g	7
Return pressure at grade, kg/cm^{2}g	2
Inlet temperature, °C	33
Temperature difference, °C	10

- **Process water**

Price, US$/nM^3	
Quality	Demineralized
Pressure, kg/cm^{2}g	5
Temperature	Ambient
pH, units	7.5 to 8.5
Conductivity, $\mu\Omega$/cm	0.5 to 1.0
SiO_2, wt ppm max.	0.1

- **Nitrogen**

Price, US$/nM^3	0.02
Pressure, kg/cm^{2}g	7
Temperature	Ambient
Purity, vol% min.	99.9
Oxygen, vol ppm max.	10

- **Instrument air**

Price, US$/NM^3	0.02
Pressure, kg/cm^{2}g	5
Temperature	Ambient
Quality	Dry, Oil-free
Dew point, °C	-40

CATALYST SPECIFICATIONS

- **EO catalyst**

Price, US$/MT	
composition	silver on alumina catalyst
shape	Spherical particles
bulk density (kg/MT)	850

- **MEG catalyst**

Price, US$/NM^3	
Catalyst	Dow MSA-1
composition	Tributylmethylamonium chloride
Shape	resin on spherical particles

Appendix 5 - Catalyst Selection

Company	date	patent #	selectivity	conversion ethylene	conversion oxygen	modifier=inhibitor	promoter	lifetime	diluent/inhibitor ratio	surface area m^2/g B.E.T.	porosimetry	support
ICI	16/8/89	357292	Discarded - ethylene concentration must be higher than 35 vol%, which is higher than our anticipated process conditions									
ICI	14/1/82	57066	76.3-76.2	5-20%	40%	Cl, best=(m)ethyl chloride or VC	K, Rb, Cs	90 days tested	ethane/cl: 5000, 2.4% vol	1.5-5	Hg: 35-65%, .3-4 mu, bimodal pore size dist.	alpha alumina
Mitsubishi	25/9/96	764464 A2	81.6		40%		Li, Cs	7 days tested		.9-1.55	H2O: 33-43%, 1.1-1.4 mu	alpha alumina, 0-6% Si
Mitsubishi	4-12-1994	624398 A1	83		40%	VC 4 ppm, 6% CO2	W, Cs, Na	7 days tested		1.02	H2O: 34.54%, .01-.4 mu	alpha alumina, Si
Mitsubishi	5-8-1987	247414 A2	82.1			ethylene dichloride	Na, K, Rb, Vs			.6-2	H2O: 20-50%	alpha alumina, .5-12% Si
Nippon Shokubai	25/8/99	937498 A1	81	10			Cs only					alpha alumina
Scientific Design	4-11-1999	99/11367	81.5-84.1	160 kg EO/hr/m^3 kat			Cs, S, F, (P, Sb, Bi)			.5-1.3	Hg: 20-70 %	alpha alumina, <15% Si
Scientific Design	4-11-1999	99/11370	84.6,83.8	160kg/hr/m^3 kat			Cs, F, S, Ge, Sn			.5-1.3	Hg: 20-70 %	
Scientific Design	4-11-1999	99/11371	~84.3	160kg/hr/m^3 kat			Cs, F, S, Lanthanide (La, Ce, Pr, Nd, Sm, Eu, Gd, Tb, Dy, Ho, Er, Tm, Yb, Lu)			.5-1.3	Hg: 20-70 %	
Shell	3-2-1999	95/05896	Only startup			ethyl chloride		10 hrs			H2O: 46%, 6.2 mu	alpha alumina, .8% Si
Shell	5-9-1996	96/13329	85		40%	ethyl chloride	Cs, Re, P, B	2-3 days		.1-3	H2O: 75%	alpha alumina,
Shell	5-11-1994	625370 A1	86% initial, decrease measured			ethyl chloride	Cs, (Gd, Ce, Yb)			.42-.51	H2O: 37%, 3.4 mu	alpha alumina, 1.3% Si
Shell	17/1/92	496470 A1	86 (ex2)			ethyl chloride		100 days, dec. to 80%				alpha alumina, 1.3% Si
Shell	30/10/87	266015	86			VC 4.5-5.5 ppm		60, sel. const!		0.42	H2O: 36%, Hg:2.7 mu	alpha alumina, 2% Si
UCC	10-4-1982	76504 A1	71-75			ethyl chloride						alpha alumina, .74% Si

Appendix 5 - Catalyst Selection

Company	wt% Ag	Li	Cs	Others	ethylene/O2 ratio mole %	diluent	T Celsius	P bar		
ICI										
ICI	6-28				.05-100	40-70% methane, .3-5% ethane	201-207			
Mitsubishi	8-20%	500-900 ppm, in 2 steps	300-800 ppm, in 2 steps		3.5	1-70% methane or N2	227-236	8	Li/Cs .1:1-4:4 wt%	Hydrocarbon halide, .1-50 ppm, prevents hot spots
Mitsubishi	5-50%		250-2000 ppm	W:5-700 ppm, Na: 92 ppm	3.50	60% N2	235-255	7	rings? Saddle's? High conversion	Other metals, see source
Mitsubishi				Na: 0.8-2%			215	15	Carrier (Si) and Na analysis	
Nippon Shokubai					2.69		240	20	Carrier must be washed with H2O	
Scientific Design	8-15%		600 ppm	S: 34 ppm, Best with Sb: 130 ppm		78% N2+CO2	240	20.5	Equivalent diam. 4-8mm for supports	
Scientific Design			640,580	F:180,75, S:34,34, Ge:75,0, Sn:0,125		N2, CO2	240	20.4		
Scientific Design			600-800	F:75, S: ~40, Lan:175		N2, CO2	240	20.4		
Shell										
Shell			340,460	Re: 280, P:10,31;B:4,42	3.53	5% CO2, 54.5 % N2	230	15.5		
Shell			430,460	Re:280, S:48, Rare earth:80	3.53	5% CO2, 54.5 % N2		15.5		
Shell	13-14%	1.5-2+4 umol/g	500-700	Re: 1.5-2 umol/g	3.53	5% CO2, 54.5 % CH4	250-265	15.5	Check exact specs of ex2	
Shell	13.5		500	Re: 260, S:35	3.53	7% CO2, 54.5% N2	250-256	14.5	ex 10	
UCC	13.1		.00906 wt%	K:.00268wt%	0.75	N2	250-260			

OUT OUT

	101	102	103	104	105	106	107	108	109	110	111	112	113	114	115	116	117	118	119	120	121	124	125	126	127	128	129	130	131	132
Temperature C	25.0	25.0	25.0	34.2	34.1	34.0	109.7	220.0	240.0	180.0	110.0	43.0	22.7	22.7	26.0	22.7	35.1	22.7	22.7	58.4	100.0	42.2	116.7	117.2	20.0	100.0	204.8	58.4	58.4	58.4
Pressure atm	25.0	25.0	25.0	22.0	22.0	22.0	22.0	22.0	19.9	19.9	19.9	19.9	20.0	20.0	20.0	20.0	22.0	20.0	20.0	20.0	20.0	20.0	10.0	20.0	20.0	10.0	20.0	20.0	20.0	20.0
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	1.00	1.00	0.94	0.94
Mole Flow kmol/hr	861	90	721	35599	35690	36410	36410	36410	36056	36056	36056	36056	34962	32340	34738	32200	34738	139	2622	2538	455	18715	17621	17621	17621	9243	632	455	224	47
Mass Flow kg/hr	24150	1449	23073	777618	779067	802140	802140	802142	802142	802142	802142	802142	762255	705086	753468	702054	753468	3032	57169	51414	20024	1148270	1108380	1108380	1108380	583963	24526	20024	9406	851
Mass Flow kg/sec	6.7	0.4	6.4	216.0	216.4	222.8	222.8	222.8	222.8	222.8	222.8	222.8	211.7	195.9	209.3	195.0	209.3	0.8	15.9	14.3	5.6	319.0	307.9	307.9	307.9	162.2	6.8	5.6	2.6	0.2
Volume Flow cum/hr	871	102	832	40751	40853	41695	53629	58992	76293	67064	55987	44876	40292	37270	40800	37110	38063	160	3022	3450	697	1283	1007	1008	929	670	1220	488	229	1
Enthalpy GJ/hr	44.30	-6.77	-0.13	-2230.47	-2237.23	-2237.36	-2122.51	-1938.01	-2133.13	-2239.44	-2354.28	-2460.99	-2371.55	-2193.68	-2285.78	-2184.25	-2274.76	-9.43	-177.87	-101.53	-177.72	-8294.94	-7971.55	-7970.36	-8205.48	-4212.69	-198.32	-177.68	-48.85	-48.85
Density kg/cum	28	14	28	19	19	19	15	14	11	12	14	18	19	18	18	18	19	18	18	14	29	895	1100	1100	1194	1120	18	41	41	1000
Mass Flow kg/hr																														
EO				0.3	0.3	0.3	0.3	0.3	31494.1	31494.1	31494.1	31494.1	0.3	0.3	0.3	0.3	0.3	0.0	0.0	0.0		31493.8	0.0	0.0	0.0		< 0.001			
MEG				139.9	139.9	139.9	139.9	139.9					11.4	10.5	139.9	10.5	139.9	0.0	0.9	129.4		472552.1	472563.5	472563.5	472563.5	234039.1	128.6			
ETHYLENE	24149.7			233206.7	233206.7	233206.7	233206.7	233206.7	209886.0	209886.0	209886.0	209886.0	208396.9	192767.1	209057.0	191938.2	209057.0	828.9	15629.8	17118.8		2089.1	599.9	599.9	599.9	599.9	1489.1			
WATER				465.5	465.5	465.5	465.5	465.5	4780.5	4780.5	4780.5	4780.5	456.8	422.5	465.5	420.7	465.5	1.8	34.3	44.8		66433.8	62110.1	62110.1	62110.1	32868.5	861.4			850.8
METHANE		1449.0		362808.6	364257.6	364257.6	364257.6	364252.6	364252.6	364252.6	364252.6	364252.6	363569.6	336301.9	362808.6	334855.8	362808.6	1446.1	27267.7	27952.8		798.0	113.0	113.0	113.0	113.0	685.1			
OXYGEN			23016.3	58373.2	58373.2	81389.6	81389.6	81397.2	58606.7	58606.7	58606.7	58606.7	58517.5	54128.7	58373.2	53895.9	58373.2	232.8	4388.8	4477.3		100.2	11.7	11.7	11.7	11.7	88.5			
ARGON			56.7	14190.0	14190.0	14246.6	14246.6	14246.6	14246.6	14246.6	14246.6	14246.6	14225.7	13158.8	14190.0	13102.2	14190.0	56.6	1066.9	1087.8		23.6	2.7	2.7	2.7	2.7	20.8			
EC				3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	1.5	1.4	3.0	1.4	3.0	0.0	0.1	1.6		566776.9	566775.4	566775.4	566775.4	311890.4	1.5			
CO2				108430.0	108430.0	108430.0	108430.0	108429.8	118871.9	118871.9	118871.9	118871.9	117075.7	108295.0	108430.0	107829.3	108430.0	465.7	8780.7	600.6	20024.5	4496.9	2700.5	2700.5	2700.5	2700.5	21250.8	20024.5	9406.0	
DEG				0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.0	0.0	0.4	0.0	0.4	< 0.001	0.0	0.4		3507.4	3507.0	3507.0	3507.0	1737.1	0.4			
ACETA				0.1	0.1	0.1	0.1	0.1	0.2	0.2	0.2	0.2	0.1	0.1	0.1	0.1	0.1	< 0.001	0.0	0.0		0.2	0.1	0.1	0.1	0.1	0.0			
Mass Frac																														
EO				384 PPB	383 PPB	372 PPB	372 PPB	372 PPB	0.039	0.039	0.039	0.039	393 PPB	393 PPB	396 PPB	393 PPB	396 PPB	393 PPB	393 PPB	446 PPB		0.027	14 PPB	14 PPB	14 PPB		18 PPB			
MEG				180 PPM	180 PPM	174 PPM	174 PPM	174 PPM					15 PPM	15 PPM	186 PPM	15 PPM	186 PPM	15 PPM	15 PPM	0.003		0.412	0.426	0.426	0.426	0.401	0.005			
ETHYLENE	1.000			0.300	0.299	0.291	0.291	0.291	0.262	0.262	0.262	0.262	0.273	0.273	0.277	0.273	0.277	0.273	0.273	0.333		0.002	541 PPM	541 PPM	541 PPM	0.001	0.061			
WATER				599 PPM	597 PPM	580 PPM	580 PPM	580 PPM	0.006	0.006	0.006	0.006	599 PPM	599 PPM	618 PPM	599 PPM	618 PPM	599 PPM	599 PPM	871 PPM		0.058	0.056	0.056	0.056	0.056	0.035			1.000
METHANE		1.000		0.467	0.468	0.454	0.454	0.454	0.454	0.454	0.454	0.454	0.477	0.477	0.482	0.477	0.482	0.477	0.477	0.544		695 PPM	102 PPM	102 PPM	102 PPM	193 PPM	0.028			
OXYGEN			0.998	0.075	0.075	0.101	0.101	0.101	0.073	0.073	0.073	0.073	0.077	0.077	0.077	0.077	0.077	0.077	0.077	0.087		87 PPM	11 PPM	11 PPM	11 PPM	20 PPM	0.004			
ARGON			0.002	0.018	0.018	0.018	0.018	0.018	0.018	0.018	0.018	0.018	0.019	0.019	0.019	0.019	0.019	0.019	0.019	0.021		21 PPM	2 PPM	2 PPM	2 PPM	5 PPM	850 PPM			
EC				4 PPM	4 PPM	4 PPM	4 PPM	4 PPM	4 PPM	4 PPM	4 PPM	4 PPM	2 PPM	2 PPM	4 PPM	2 PPM	4 PPM	2 PPM	2 PPM	31 PPM		0.494	0.511	0.511	0.511	0.534	61 PPM			
CO2				0.139	0.139	0.135	0.135	0.135	0.148	0.148	0.148	0.148	0.154	0.154	0.144	0.154	0.144	0.154	0.154	0.012	1.000	0.004	0.002	0.002	0.002	0.005	0.866	1.000	1.000	
DEG				524 PPB	523 PPB	508 PPB	508 PPB	508 PPB	508 PPB	508 PPB	508 PPB	508 PPB	22 PPB	22 PPB	541 PPB	22 PPB	541 PPB	22 PPB	22 PPB	8 PPM		0.003	0.003	0.003	0.003	0.003	16 PPM			
ACETA				146 PPB	145 PPB	141 PPB	141 PPB	141 PPB	278 PPB	278 PPB	278 PPB	278 PPB	127 PPB	127 PPB	150 PPB	127 PPB	150 PPB	127 PPB	127 PPB	472 PPB		214 PPB	107 PPB	107 PPB	107 PPB	203 PPB	694 PPB			
Mole Flow kmol/hr																														
EO				0.0	0.0	0.0	0.0	0.0	714.9	714.9	714.9	714.9	0.0	0.0	0.0	0.0	0.0	< 0.001	0.0	0.0		714.9	< 0.001	< 0.001	< 0.001		< 0.001			
MEG				2.3	2.3	2.3	2.3	2.3					0.2	0.2	2.3	0.2	2.3	0.0	0.0	2.1		7613.4	7613.6	7613.6	7613.6	3770.7	2.1			
ETHYLENE	860.8			8312.9	8312.9	8312.9	8312.9	8312.9	7481.6	7481.6	7481.6	7481.6	7428.5	6871.3	7452.0	6841.8	7452.0	29.5	557.1	610.2		74.5	21.4	21.4	21.4	21.4	53.1			
WATER				25.8	25.8	25.8	25.8	25.8	265.4	265.4	265.4	265.4	25.4	23.5	25.8	23.4	25.8	0.1	1.9	2.5		3687.6	3447.6	3447.6	3447.6	1824.5	47.8			47.2
METHANE		90.3		22615.1	22705.4	22705.4	22705.4	22705.1	22705.1	22705.1	22705.1	22705.1	22662.5	20962.8	22615.1	20872.7	22615.1	90.1	1699.7	1742.4		49.7	7.0	7.0	7.0	7.0	42.7			
OXYGEN			719.3	1824.2	1824.2	2543.5	2543.5	2543.8	1831.5	1831.5	1831.5	1831.5	1828.7	1691.6	1824.2	1684.3	1824.2	7.3	137.2	139.9		3.1	0.4	0.4	0.4	0.4	2.8			
ARGON			1.4	355.2	355.2	356.6	356.6	356.6	356.6	356.6	356.6	356.6	356.1	329.4	355.2	328.0	355.2	1.4	26.7	27.2		0.6	0.1	0.1	0.1	0.1	0.5			
EC				0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	< 0.001	0.0	0.0		6436.0	6436.0	6436.0	6436.0	3541.7	0.0			
CO2				2463.8	2463.8	2463.8	2463.8	2463.8	2701.0	2701.0	2701.0	2701.0	2660.2	2460.7	2463.8	2450.1	2463.8	10.6	199.5	13.6	455.0	102.2	61.4	61.4	61.4	61.4	482.9	455.0	224.0	
DEG				0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	< 0.001	< 0.001	0.0	< 0.001	0.0	trace	< 0.001	0.0		33.1	33.0	33.0	33.0	16.4	0.0			
ACETA				0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	trace	< 0.001	0.0		0.0	0.0	0.0	0.0	0.0	< 0.001			
Mole Frac																														
EO				190 PPB	190 PPB	186 PPB	186 PPB	186 PPB	0.020	0.020	0.020	0.020	194 PPB	194 PPB	195 PPB	194 PPB	195 PPB	194 PPB	194 PPB	205 PPB		0.038	20 PPB	20 PPB	20 PPB		16 PPB			
MEG				63 PPM	63 PPM	62 PPM	62 PPM	62 PPM					5 PPM	5 PPM	65 PPM	5 PPM	65 PPM	5 PPM	5 PPM	822 PPM		0.407	0.432	0.432	0.432	0.408	0.003			
ETHYLENE	1.000			0.234	0.233	0.228	0.228	0.228	0.207	0.207	0.207	0.207	0.212	0.212	0.215	0.212	0.215	0.212	0.212	0.240		0.004	0.001	0.001	0.001	0.002	0.084			
WATER				726 PPM	724 PPM	710 PPM	710 PPM	710 PPM	0.007	0.007	0.007	0.007	725 PPM	725 PPM	744 PPM	725 PPM	744 PPM	725 PPM	725 PPM	979 PPM		0.197	0.196	0.196	0.196	0.197	0.076			1.000
METHANE		1.000		0.635	0.636	0.624	0.624	0.624	0.630	0.630	0.630	0.630	0.648	0.648	0.651	0.648	0.651	0.648	0.648	0.687		0.003	400 PPM	400 PPM	400 PPM	762 PPM	0.068			
OXYGEN			0.998	0.051	0.051	0.070	0.070	0.070	0.051	0.051	0.051	0.051	0.052	0.052	0.053	0.052	0.053	0.052	0.052	0.055		167 PPM	21 PPM	21 PPM	21 PPM	40 PPM	0.004			
ARGON			0.002	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.010	0.011		32 PPM	4 PPM	4 PPM	4 PPM	7 PPM	826 PPM			
EC				949 PPB	947 PPB	928 PPB	928 PPB	928 PPB	937 PPB	937 PPB	937 PPB	937 PPB	483 PPB	483 PPB	973 PPB	483 PPB	973 PPB	483 PPB	483 PPB	7 PPM		0.344	0.365	0.365	0.365	0.383	27 PPM			
CO2				0.069	0.069	0.068	0.068	0.068	0.075	0.075	0.075	0.075	0.076	0.076	0.071	0.076	0.071	0.076	0.076	0.005	1.000	0.005	0.003	0.003	0.003	0.007	0.764	1.000	1.000	
DEG				108 PPB	108 PPB	106 PPB	106 PPB	106 PPB	107 PPB	107 PPB	107 PPB	107 PPB	4 PPB	4 PPB	111 PPB	4 PPB	111 PPB	4 PPB	4 PPB	1 PPM		0.002	0.002	0.002	0.002	0.002	6 PPM			
ACETA				72 PPB	72 PPB	71 PPB	71 PPB	71 PPB	141 PPB	141 PPB	141 PPB	141 PPB	63 PPB	63 PPB	74 PPB	63 PPB	74 PPB	63 PPB	63 PPB	217 PPB		297 PPB	153 PPB	153 PPB	153 PPB	291 PPB	611 PPB			

			OUT				OUT	OUT				OUT
	301	302	304	305	306	307	308	309	310	311	312	313
Temperature C	125.0	145.6	129.8	129.8	132.3	132.3	14.000	164.4	132.3	132.9	125.0	14.000
Pressure atm	9.5	9.5	1.0	0.1	0.1	0.1	0.100	0.1	0.1	10.0	9.5	0.100
Vapor Frac	0.00	0.00	0.00	0.00	0.00	0.00	1.000	0.00	0.00	0.00	1.00	0.000
Mole Flow kmol/hr	1	1084	414	414	867	416	45	2	451	451	232	172
Mass Flow kg/hr	45	58199	25666	25666	53908	25876	1152	210	28032	28032	9732	3139 ←
Mass Flow kg/sec	0.0	16.2	7.1	7.1	15.0	7.2	0.3	0.1	7.8	7.8	2.7	0.9
Volume Flow cum/hr	0	59	25	25	53	26	9961	0	28	28	780	3
Enthalpy GJ/hr	-0.35	-442.89	-182.85	-182.86	-383.16	-183.92	-14.153	-1.21	-199.25	-199.20	-87.40	-48.790
Density kg/cum	823	991	1013	1013	1010	1010	0	930	1010	1010	12	978
Mass Flow kg/hr												
EO												
MEG	40.9	53524.7	25640.5	25640.5	53471.2	25666.2		25.7	27805.0	27805.0	108.5	53.5
ETHYLENE	< 0.001	0.1			trace	trace	0.1		trace	trace	2.1	
WATER	3.6	3694.0	10.6	10.6	22.2	10.6	586.4	trace	11.5	11.5	341.8	3085.4
METHANE	trace	0.0			trace	trace	0.0		trace	trace	0.1	
OXYGEN	trace	< 0.001			trace	trace	< 0.001		trace	trace	0.0	
ARGON	trace	< 0.001			trace	trace	< 0.001		trace	trace	0.0	
EC	0.0	18.5	3.8	3.8	18.5	8.9		5.1	9.6	9.6	0.0	
CO2	0.5	565.5			trace	trace	565.5		trace	trace	9279.5	
DEG	0.2	395.9	11.0	11.0	395.9	190.0		179.0	205.9	205.9	0.2	
ACETA	trace	0.0			trace	trace			trace	trace	0.0	0.0
Mass Frac												
EO							0.000					0.000
MEG	0.906	0.920	0.999	0.999	0.992	0.992	0.000	0.122	0.992	0.992	0.011	0.017
ETHYLENE	999 PPB	1 PPM			trace	trace	0.000		trace	trace	214 PPM	0.000
WATER	0.079	0.063	415 PPM	415 PPM	411 PPM	411 PPM	0.509	trace	411 PPM	411 PPM	0.035	0.983
METHANE	22 PPB	23 PPB			trace	trace	0.000		trace	trace	14 PPM	0.000
OXYGEN	1 PPB	2 PPB			trace	trace			trace	trace	1 PPM	0.000
ARGON	trace	trace			trace	trace			trace	trace	279 PPB	0.000
EC	140 PPM	318 PPM	147 PPM	147 PPM	343 PPM	343 PPM	0.000	0.024	343 PPM	343 PPM	309 PPB	0.000
CO2	0.010	0.010			trace	trace	0.491		trace	trace	0.953	0.000
DEG	0.005	0.007	429 PPM	429 PPM	0.007	0.007	0.000	0.853	0.007	0.007	21 PPM	0.000
ACETA	37 PPB	35 PPB			trace	trace	0.000		trace	trace	353 PPB	0.000
Mole Flow kmol/hr												
EO												
MEG	0.7	862.4	413.1	413.1	861.5	413.5		0.4	448.0	448.0	1.7	0.9
ETHYLENE	trace	0.0			trace	trace	0.0		trace	trace	0.1	
WATER	0.2	205.0	0.6	0.6	1.2	0.6	32.5	trace	0.6	0.6	19.0	171.3
METHANE	trace	< 0.001			trace	trace	< 0.001		trace	trace	0.0	
OXYGEN	trace	trace			trace	trace	trace		trace	trace	< 0.001	
ARGON	trace	trace			trace	trace	trace		trace	trace	< 0.001	
EC	< 0.001	0.2	0.0	0.0	0.2	0.1		0.1	0.1	0.1	< 0.001	
CO2	0.0	12.8			trace	trace	12.8		trace	trace	210.9	
DEG	0.0	3.7	0.1	0.1	3.7	1.8		1.7	1.9	1.9	0.0	
ACETA	trace	< 0.001			trace	trace	< 0.001		trace	trace	< 0.001	
Mole Frac												
EO							0.000					0.000
MEG	0.758	0.795	0.998	0.998	0.994	0.994	0.000	0.192	0.994	0.994	0.008	0.005
ETHYLENE	2 PPM	2 PPM			trace	trace	0.000		trace	trace	321 PPM	0.000
WATER	0.228	0.189	0.001	0.001	0.001	0.001	0.717	trace	0.001	0.001	0.082	0.995
METHANE	70 PPB	76 PPB			trace	trace			trace	trace	37 PPM	0.000
OXYGEN	2 PPB	3 PPB			trace	trace			trace	trace	2 PPM	0.000
ARGON	trace	trace			trace	trace			trace	trace	293 PPB	0.000
EC	83 PPM	194 PPM	104 PPM	104 PPM	243 PPM	243 PPM	0.000	0.027	243 PPM	243 PPM	147 PPB	0.000
CO2	0.012	0.012			trace	trace	0.283		trace	trace	0.910	0.000
DEG	0.002	0.003	251 PPM	251 PPM	0.004	0.004	0.000	0.782	0.004	0.004	9 PPM	0.000
ACETA	44 PPB	42 PPB			trace	trace			trace	trace	336 PPB	0.000

	401	402	403	404	405	406	↓ 407	408	409	↓ 410	411	412	413	414	416	417	OUT 418	OUT 419	420	421	422	423	424	425	426
Temperature C	41.8	61.2	133.7	155.7	130.8	50.3	22.8	137.9	56.8	61.1	55.8	107.0	135.8	137.1	43.0	109.6	120.6	29.2	110.9	46.6	100.0	43.0	43.0	107.9	43.0
Pressure atm	20.0	1.0	2.5	10.0	10.0	10.0	0.5	2.5	2.5	5.0	5.0	10.0	1.0	10.0	10.0	0.5	2.0	2.0	10.0	10.0	10.0	2.5	10.0	5.0	5.0
Vapor Frac	0.00	1.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00	0.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00	1.00	0.19	1.00	0.09
Mole Flow kmol/hr	7711	428	428	428	694	1226	582	582	582	316	266	266	7282	7282	633	644	21	295	644	101	633	428	694	582	582
Mass Flow kg/hr	473088	16793	16793	16793	28352	65016	24922	24922	24922	13363	11559	11559	456295	456295	40000	40094	387	12976	40094	3337	40000	16793	28352	24922	24922
Mass Flow kg/sec	131.4	4.7	4.7	4.7	7.9	18.1	6.9	6.9	6.9	3.7	3.2	3.2	126.7	126.7	11.1	11.1	0.1	3.6	11.1	0.9	11.1	4.7	7.9	6.9	6.9
Volume Flow cum/hr	409	11628	5637	1433	2301	78	28008	7851	6302	16	1435	830	539	422	34	46	0	15	36	257	46	4444	365	3639	289
Enthalpy GJ/hr	-3417.52	-40.79	-39.18	-38.81	-59.91	-351.88	-43.02	-39.34	-42.01	-27.72	-21.85	-21.11	-3269.47	-3268.87	-293.45	-290.87	-5.95	-22.74	-290.80	-18.15	-288.56	-41.13	-76.59	-40.39	-55.27
Density kg/cum	1157	1	3	12	12	919	4	727	4	809	8	14	847	1081	1175	858	26	26	1106	9	1120	4	80	7	834
Mass Flow kg/hr																									
EO	12975.5	12975.5	12975.5	12975.5	23194.8	23194.8	23194.8	23194.8	23194.8	12975.4	10219.3	10219.3	< 0.001	< 0.001		0.0	0.1	12975.3	0.0	< 0.001		12975.5	23194.8	23194.8	23194.8
MEG	194691.5	27.7	27.7	27.7	27.7	16058.4	2.9	2.9	2.9	2.9	0.0	0.0	194663.8	194663.8	16031.1	16055.6	2.9	< 0.001	16055.6	0.3	16031.1	27.7	27.7	2.9	2.9
ETHYLENE	860.7	860.7	860.7	860.7	1041.3	180.7	180.7	180.7	180.7	trace	180.7	180.7	trace	trace	41.1	trace			trace	901.8	41.1	860.7	1041.3	180.7	180.7
WATER	27370.7	692.5	692.5	692.5	710.7	2954.0	402.2	402.2	402.2	384.0	18.2	18.2	26678.2	26678.2	2251.4	2551.8	383.9	0.1	2551.8	8.1	2251.4	692.5	710.7	402.2	402.2
METHANE	328.8	328.8	328.8	328.8	343.2	14.4	14.4	14.4	14.4	trace	14.4	14.4	trace	trace	7.7	trace			trace	336.5	7.7	328.8	343.2	14.4	14.4
OXYGEN	41.3	41.3	41.3	41.3	42.5	1.2	1.2	1.2	1.2	trace	1.2	1.2	trace	trace	0.8	trace			trace	42.1	0.8	41.3	42.5	1.2	1.2
ARGON	9.7	9.7	9.7	9.7	10.0	0.3	0.3	0.3	0.3	trace	0.3	0.3	trace	trace	0.2	trace			trace	9.9	0.2	9.7	10.0	0.3	0.3
EC	233512.1	3.9	3.9	3.9	3.9	21367.6	0.4	0.4	0.4	0.4	< 0.001	< 0.001	233508.1	233508.1	21363.7	21367.2	0.4	trace	21367.2	0.0	21363.7	3.9	3.9	0.4	0.4
CO2	1852.7	1852.7	1852.7	1852.7	2978.0	1125.2	1125.2	1125.2	1125.2	trace	1125.2	1125.2	trace	trace	185.0	trace			trace	2037.7	185.0	1852.7	2978.0	1125.2	1125.2
DEG	1445.1	0.0	0.0	0.0	0.0	119.0	0.0	0.0	0.0	0.0	trace	trace	1445.0	1445.0	119.0	119.0	0.0	trace	119.0	< 0.001	119.0	0.0	0.0	0.0	0.0
ACETA	0.1	0.1	0.1	0.1	0.2	0.2	0.2	0.2	0.2	0.1	0.1	0.1	trace	trace	0.0	trace	trace	0.1	trace	0.0	0.0	0.1	0.2	0.2	0.2
Mass Frac																									
EO	0.027	0.773	0.773	0.773	0.818	0.357	0.931	0.931	0.931	0.971	0.884	0.884	trace	trace		381 PPB	335 PPM	1.000	381 PPB	135 PPB		0.773	0.818	0.931	0.931
MEG	0.412	0.002	0.002	0.002	976 PPM	0.247	116 PPM	116 PPM	116 PPM	217 PPM	67 PPB	67 PPB	0.427	0.427	0.401	0.400	0.007	3 PPB	0.400	95 PPM	0.401	0.002	976 PPM	116 PPM	116 PPM
ETHYLENE	0.002	0.051	0.051	0.051	0.037	0.003	0.007	0.007	0.007	trace	0.016	0.016	trace	trace	0.001	trace			trace	0.270	0.001	0.051	0.037	0.007	0.007
WATER	0.058	0.041	0.041	0.041	0.025	0.045	0.016	0.016	0.016	0.029	0.002	0.002	0.058	0.058	0.056	0.064	0.991	9 PPM	0.064	0.002	0.056	0.041	0.025	0.016	0.016
METHANE	695 PPM	0.020	0.020	0.020	0.012	222 PPM	579 PPM	579 PPM	579 PPM	trace	0.001	0.001	trace	trace	193 PPM	trace			trace	0.101	193 PPM	0.020	0.012	579 PPM	579 PPM
OXYGEN	87 PPM	0.002	0.002	0.002	0.001	19 PPM	50 PPM	50 PPM	50 PPM	trace	107 PPM	107 PPM	trace	trace	20 PPM	trace			trace	0.013	20 PPM	0.002	0.001	50 PPM	50 PPM
ARGON	21 PPM	579 PPM	579 PPM	579 PPM	353 PPM	4 PPM	11 PPM	11 PPM	11 PPM	trace	25 PPM	25 PPM	trace	trace	5 PPM	trace			trace	0.003	5 PPM	579 PPM	353 PPM	11 PPM	11 PPM
EC	0.494	235 PPM	235 PPM	235 PPM	139 PPM	0.329	15 PPM	15 PPM	15 PPM	28 PPM	1 PPB	1 PPB	0.512	0.512	0.534	0.533	968 PPM	trace	0.533	15 PPM	0.534	235 PPM	139 PPM	15 PPM	15 PPM
CO2	0.004	0.110	0.110	0.110	0.105	0.017	0.045	0.045	0.045	trace	0.097	0.097	trace	trace	0.005	trace			trace	0.611	0.005	0.110	0.105	0.045	0.045
DEG	0.003	2 PPM	2 PPM	2 PPM	1 PPM	0.002	146 PPB	146 PPB	146 PPB	272 PPB	trace	trace	0.003	0.003	0.003	0.003	9 PPM	trace	0.003	143 PPB	0.003	2 PPM	1 PPM	146 PPB	146 PPB
ACETA	214 PPB	6 PPM	6 PPM	6 PPM	6 PPM	3 PPM	7 PPM	7 PPM	7 PPM	8 PPM	7 PPM	7 PPM	trace	trace	203 PPB	trace	7 PPB	8 PPM	trace	504 PPB	203 PPB	6 PPM	6 PPM	7 PPM	7 PPM
Mole Flow kmol/hr																									
EO	294.5	294.5	294.5	294.5	526.5	526.5	526.5	526.5	526.5	294.5	232.0	232.0	trace	trace		< 0.001	0.0	294.5	< 0.001	< 0.001		294.5	526.5	526.5	526.5
MEG	3136.7	0.4	0.4	0.4	0.4	258.7	0.0	0.0	0.0	0.0	< 0.001	< 0.001	3136.3	3136.3	258.3	258.7	0.0	trace	258.7	0.0	258.3	0.4	0.4	0.0	0.0
ETHYLENE	30.7	30.7	30.7	30.7	37.1	6.4	6.4	6.4	6.4	trace	6.4	6.4	trace	trace	1.5	trace			trace	32.1	1.5	30.7	37.1	6.4	6.4
WATER	1519.3	38.4	38.4	38.4	39.4	164.0	22.3	22.3	22.3	21.3	1.0	1.0	1480.9	1480.9	125.0	141.6	21.3	0.0	141.6	0.5	125.0	38.4	39.4	22.3	22.3
METHANE	20.5	20.5	20.5	20.5	21.4	0.9	0.9	0.9	0.9	trace	0.9	0.9	trace	trace	0.5	trace			trace	21.0	0.5	20.5	21.4	0.9	0.9
OXYGEN	1.3	1.3	1.3	1.3	1.3	0.0	0.0	0.0	0.0	trace	0.0	0.0	trace	trace	0.0	trace			trace	1.3	0.0	1.3	1.3	0.0	0.0
ARGON	0.2	0.2	0.2	0.2	0.3	0.0	0.0	0.0	0.0	trace	0.0	0.0	trace	trace	0.0	trace			trace	0.2	0.0	0.2	0.3	0.0	0.0
EC	2651.6	0.0	0.0	0.0	0.0	242.6	0.0	0.0	0.0	0.0	trace	trace	2651.6	2651.6	242.6	242.6	0.0	trace	242.6	0.0	242.6	0.0	0.0	0.0	0.0
CO2	42.1	42.1	42.1	42.1	67.7	25.6	25.6	25.6	25.6	trace	25.6	25.6	trace	trace	4.2	trace			trace	46.3	4.2	42.1	67.7	25.6	25.6
DEG	13.6	< 0.001	< 0.001	< 0.001	< 0.001	1.1	< 0.001	< 0.001	< 0.001	< 0.001	trace	trace	13.6	13.6	1.1	1.1	< 0.001	trace	1.1	trace	1.1	< 0.001	< 0.001	< 0.001	< 0.001
ACETA	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	trace	trace	< 0.001	trace	trace	0.0	trace	< 0.001	< 0.001	0.0	0.0	0.0	0.0
Mole Frac																									
EO	0.038	0.688	0.688	0.688	0.758	0.429	0.905	0.905	0.905	0.932	0.872	0.872	trace	trace		538 PPB	138 PPM	1.000	538 PPB	101 PPB		0.688	0.758	0.905	0.905
MEG	0.407	0.001	0.001	0.001	642 PPM	0.211	80 PPM	80 PPM	80 PPM	148 PPM	47 PPB	47 PPB	0.431	0.431	0.408	0.402	0.002	2 PPB	0.402	50 PPM	0.408	0.001	642 PPM	80 PPM	80 PPM
ETHYLENE	0.004	0.072	0.072	0.072	0.053	0.005	0.011	0.011	0.011	trace	0.024	0.024	trace	trace	0.002	trace			trace	0.317	0.002	0.072	0.053	0.011	0.011
WATER	0.197	0.090	0.090	0.090	0.057	0.134	0.038	0.038	0.038	0.067	0.004	0.004	0.203	0.203	0.197	0.220	0.997	23 PPM	0.220	0.004	0.197	0.090	0.057	0.038	0.038
METHANE	0.003	0.048	0.048	0.048	0.031	733 PPM	0.002	0.002	0.002	trace	0.003	0.003	trace	trace	762 PPM	trace			trace	0.207	762 PPM	0.048	0.031	0.002	0.002
OXYGEN	167 PPM	0.003	0.003	0.003	0.002	32 PPM	66 PPM	66 PPM	66 PPM	trace	145 PPM	145 PPM	trace	trace	40 PPM	trace			trace	0.013	40 PPM	0.003	0.002	66 PPM	66 PPM
ARGON	32 PPM	568 PPM	568 PPM	568 PPM	361 PPM	6 PPM	12 PPM	12 PPM	12 PPM	trace	27 PPM	27 PPM	trace	trace	7 PPM	trace			trace	0.002	7 PPM	568 PPM	361 PPM	12 PPM	12 PPM
EC	0.344	105 PPM	105 PPM	105 PPM	65 PPM	0.198	7 PPM	7 PPM	7 PPM	13 PPM	trace	trace	0.364	0.364	0.383	0.377	199 PPM	trace	0.377	6 PPM	0.383	105 PPM	65 PPM	7 PPM	7 PPM
CO2	0.005	0.098	0.098	0.098	0.097	0.021	0.044	0.044	0.044	trace	0.096	0.096	trace	trace	0.007	trace			trace	0.456	0.007	0.098	0.097	0.044	0.044
DEG	0.002	897 PPB	897 PPB	897 PPB	553 PPB	915 PPM	59 PPB	59 PPB	59 PPB	109 PPB	trace	trace	0.002	0.002	0.002	0.002	2 PPM	trace	0.002	44 PPB	0.002	897 PPB	553 PPB	59 PPB	59 PPB
ACETA	297 PPB	5 PPM	5 PPM	5 PPM	6 PPM	3 PPM	7 PPM	7 PPM	7 PPM	8 PPM	6 PPM	6 PPM	trace	trace	291 PPB	trace	3 PPB	8 PPM	trace	377 PPB	291 PPB	5 PPM	6 PPM	7 PPM	7 PPM

Appendix 7

Summary of Reactors,Columns and Vessels

equipment No.	Name	Operating temperature °C	Operating pressure bara	Size dia. x length mm	Internals			Materials of construction	
					Type	trays (nr).	packing	Shell	internals/tube
R101	EO reactor	230	22	4670 x 11540	multitubular		Ag on aluminium	CS	SS
R201	Carbonate reactor	120	33	4000 x 16040	packed bed		DOW MSA1	SS	
R202	Primary MEG reactor	150	10	4000 x 6680	packed bed		DOW MSA1	CS	
R203	Secondary MEG reactor	150	10	4000 x 6640	packed bed		DOW MSA1	CS	
C101	EO absorber	43	20.0	4720 x 31500	packed column	15	mellapak 250X	CS	PP
C301	Water removal	145	0.1	2770 x 6100	tray column	sieve (10)		CS	CS
C302	MEG purification	130	0.09	2920 x 23100	packed column	15	mellapak 250X	CS	SS
C401	Raw EO stripper	42	1	3380 x 20300	packed column	10	mellapak 250Y	SS	PVDF
C402	EO secondary absorber	43	10	790 x 23200	packed column	9	mellapak 250Y	SS	PP
C403	EO secondary stripper	50	0.5	1770 x 11600	packed column	5	mellapak 250Y	SS	PVDF
C404	Heavy end seperator	61	2.0	1220 x 11582	tray column	sieve (19)		SS	SS
C405	Light end seperator	57	5	910 x 6100	tray column	sieve (10)		SS	SS
V101*	Steam drum	230	27	volume: 8.8 m^3	vessel			CS	
V102*	Steam drum	160	8	volume: 4.3 m^3	vessel			CS	
V301	V/L seperator C301	44	0.1	volume: 4.5 m^3	vessel			CS	
V302	V/L seperator C302	130	0.09	volume: 2.5 m^3	vessel			CS	
V401	V/L seperator C404	30	2	volume: 1.3 m^3	vessel			CS	
S201	CO_2 flashdrum	100	10	volume: 53.5 m^3	flashvessel			CS	
S202	flashdrum	100	8.0	volume: 70 m^3	flashvessel			CS	
S301	CO_2 flashdrum	125	9.5	volume: 66.5 m^3	flashvessel			CS	

Remarks: * Steam drums are installed two times, because two EO reactors are placed parallel

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Summary of Reactors,Columns and Vessels

Summary of Pumps and Compressors

Pump No.	Name	Spared	Capacity Normal m^3/h	Capacity Maximum m^3/h	Suction Pressure bara	discharge Pressure bara	Pumping temp. °C	Density at pumping temp. kg/m^3	Theoretical power kW	effeciency	Brake power kW	Type
P101	Main absorbent pump	Yes	1007.2	1107.92	10	24	117	1100	397.65	0.86	465.1	Centrifugal
P102	Absorbent pump		27.8	30.58	0.1	11.5	130	1010	8.94	0.56	15.9	Centrifugal
P103*	Coolant pump	Yes	106.6	117.26	28	29.5	230	1000	4.51	0.72	6.3	Centrifugal
P104*	Coolant pump	Yes	51.117	56.2287	28	29.5	230	1000	2.16	0.72	3.0	Centrifugal
P201	EO feed pump		583.4	641.74	20	34.5	42	1157	238.56	0.83	287.4	Centrifugal
P202	liquid phase pump S202		0.5	0.55	8	11.5	100	965	0.05	0.30	small	positive displacement
P203	Water pump		8.2	9.02	5	11.5	25	994	1.50	0.72	small	positive displacement
P204	liquid phase pump S201		795.8	875.38	10	11.5	100	1086	33.66	0.72	46.8	Centrifugal
P205	liquid phase pump R202		82.3	90.53	9.5	11.5	133	822	4.64	0.72	6.4	Centrifugal
P301	Reflux pump C301		37.5	41.25	0.1	1.6	44	1000	1.59	0.72	small	positive displacement
P302	Reflux pump C302		11.4	12.54	0.1	1.6	130	1013	0.48	0.72	small	positive displacement
P303	MEG pump		25.3	27.83	0.1	2.5	130	1013	1.71	0.72	small	positive displacement
P304	Heavy component pump		0.2	0.22	0.1	1.6	164	930	0.01	0.72	small	positive displacement
P305	Raw MEG pump		58.8	64.68	9.5	11	146	991	2.49	0.72	3.5	Centrifugal
P401	Absorbent pump		539	592.9	1	11.5	137	1081	159.60	0.62	257.0	Centrifugal
P402	Reflux pump C404		2.1	2.31	2	3.5	29		0.09	0.72	small	positive displacement
P403	Bottom pump C402		77.6	85.36	10	13	50	919	6.57	0.72	9.1	Centrifugal
P404	Absorbent pump		46.4	51.04	0.5	11.5	111	1106	14.39	0.62	23.1	Centrifugal
P405	Bottom pump C405		16.4	18.04	5	8	61	809	1.39	0.72	small	positive displacement
K101	Loop gas compressor		40800.2	44880.22	20	22	35.1	19	2202.2	0.72	3058.6	Rotary
K102	Recycle compressor		1219.5	1341.45	10	20	204.8	18	536.7	0.72	745.4	Rotary
K103	CO2 compressor		696.6	766.26	20	33	155	41	206.6	0.72	286.9	Rotary
K301	Vacuum compressor C301		9961.4	10957.54	0.1	1.6	44	14	340	0.5	680.0	Liquid ring
K302	Vacuum compressor C302		100	110	0.1	1.6	44	14	3.5	0.5	7.0	Liquid ring
K401	EO. pri. compressor		11628	12790.8	1	2.5	134	3	322.9	0.72	448.5	Rotary
K402	EO. sec. compressor		4444.2	4888.62	2.5	10	156	12	463.7	0.72	644.0	Rotary
K403	EO recycle compressor		1435.4	1578.94	5	10	107	14	148	0.72	205.6	Rotary
K404	Raw EO pri. compressor		28008.3	30809.13	0.5	2.5	138	4	768.69	0.72	1067.6	Rotary
K405	Raw EO sec. compressor		24922.2	27414.42	2.5	5	108	7	324.8	0.72	451.1	Rotary

Remarks: * Pumps are installed twice, because two EO reactors are placed parallel

Small means power < 2.5 kW, on/off operation employed

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Summary of Pumps and Compressors

Appendix 9

Stork Engineers & Contractors

Summary of Heatexchangers

HEAT EXCHANGER SUMMARY

Equipment No.	Number of exchangers	Parallel	In series	Service	Type of equipment 1)	Duty	Total (bare) outside surface	Overall heat transfer coef. 2)	Heat Flux	Shell	Design conditions						Materials of construction	
											Shell ST			Tubes ST+AC				
E						kW	m²	W/m².K	W/m²	Diam.	Tin °C	Tout °C	Press. bar ga	Tin °C	Tout °C	Press. bar ga	Shell/ headers	Tubes
101	4	*		Exchanger	ST	31901	993	110	-	2.13	34	109.7	21.0	180	110	19.9	ss	ss
102	2	*		Exchanger	ST	51290	1928	300	-	4.7	230	230	27.6	110	220	25.0	cs	ss
103	2	*		Exchanger	ST	29551	1138	300	-	4.7	160	160	7.8	240	180	25.0	cs	ss
104	4	*		Exchanger	ST	29665	825	300	-	2.68	33	43	1.0	110	43	19.9	cs	ss
105a	1			Exchanger	ST	15466	477	900	-	2.21	4	10	1.0	43	20	20.0	cs	ss
105b	1	*		Exchanger	ST	23787	825	900	-	2.84	78	43	1.0	33	43	20.0	cs	ss
201	3	*		Exchanger	ST	26108	983	400	-	1.52	42	105	20.0	117	78	33.0	cs	ss
202	1			Exchanger	ST	3473	111	1000	-	1.09	160	160	7.8	89	150	10.0	cs	ss
203	1			Exchanger	ST	406	24	1000	-	0.55	160	160	7.8	133	150	10.0	cs	ss
301	1			Reboiler	ST	3500	100	-	35000	1.04	160	160	7.8	-	132	0.1	cs	cs
302	1			Condenser	ST	3389	253	2000	-	1.57	33	40	1.0	-	44	0.1	cs	cs
303	1			Reboiler	ST	8889	148	-	60000	1.24	230	230	27.6	-	164	0.1	cs	cs
304	1			Condenser	ST	8944	256	-	35000	1.58	120	120	1.9	-	130	0.1	cs	cs

1) ST = shell-and-tube heat exchanger

AC = air-cooled equipment

2) Based on bare outside surface

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Appendix 9

Stork Engineers & Contractors

Summary of Heatexchangers

Equipment No. E	Number of exchangers	Parallel	In series	Service	Type of equipment 1)	Duty kW	Total (bare) outside surface m^2	Overall heat transfer coef. 2) $W/m^2.K$	Heat Flux W/m^2	Shell Diam.	Design conditions Shell ST Tin °C	Shell ST Tout °C	Shell ST Press. bar ga	Tubes ST+AC Tin °C	Tubes ST+AC Tout °C	Tubes ST+AC Press. bar ga	Materials of construction Shell/ headers	Tubes
401	1			Reboiler	ST	29806	497	-	60000	2.13	230	230	27.6	-	136	1.0	cs	ss
402	1			Exchanger	ST	542	49	300	-	0.76	33	43	1.0	134	43	2.5	cs	cs
403	1			Exchanger	ST	4635	431	300	-	2	33	43	1.0	131	43	10.0	cs	ss
404	1			Exchanger	ST	1359	67	750	-	0.87	33	43	1.0	100	43	10.0	cs	ss
405	1			Reboiler	ST	5000	143	-	35000	1.22	160	160	7.8	-	109	0.5	cs	ss
406	1			Reboiler	ST	2031	51	-	40000	0.77	160	160	7.8	-	121	2.0	cs	ss
407	1			Condenser	ST	2300	52	2000	-	0.78	4	10	1.0	-	29	2.0	cs	ss
408	1			Exchanger	ST	743	48	300	-	0.78	33	43	1.0	138	57	2.5	cs	ss
409	1			Exchanger	ST	4135	188	750	-	188	33	43	1.0	108	43	5.0	cs	ss
410	1			Reboiler	ST	1583	6	-	60000	26	160	160	7.8	-	61	5.0	cs	ss

1) ST = shell-and-tube heat exchanger

AC = air-cooled equipment

2) Based on bare outside surface

Rev.	Date	Sign.

HEAT EXCHANGER SUMMARY

Design book No. :

page

Appendix 10

REACTOR - SPECIFICATION SHEET

Equipment number :	R101 A/B	Parallel:
Name :	Ethylene oxide reactor	2

General Data

Type :	Cooled multitubular reactor			
Catalyst :	Silver doped on alumina			
Space time :	5.9 sec. In reaction section			
Volume :	90.8 m^3			
Tubes:		Shell:		
- number :	6265	- inner diameter	:	4.66 m
- length :	11.5 m	- outer diameter	:	4.70 m
- inner diameter :	40 mm	- material	:	CS
- outer diameter :	44 mm			
- pitch :	55 mm			
- material :	SS ?			

Process Conditions					Cooling Reaction section	
Stream details	Feed 108		Product 109		Coolant inlet	Coolant outlet
Temp. [°C]	220		240		230	230
Pressure [bar]	22		19.9		27.6	27.6
Density [kg/m^3]	14.0		11.0			
Mass flow [kg/h]	802142.2		802142.2		106557	106557
Vapor fraction	1		1		0	1
Composition (frac)	mol	wt	mol	wt		
EO	186 PPB	372 PPB	0.02	0.039		
MEG	62 PPM	174 PPM				
Ethylene	0.228	0.291	0.207	0.262		
Water	710 PPM	580 PPM	0.007	0.006		
Methane	0.624	0.454	0.63	0.454		
Oxygen	0.07	0.101	0.051	0.073		
Argon	0.01	0.018	0.01	0.018		
EC	928 PPB	4 PPM	937 PPB	4 PPM		
CO_2	0.068	0.135	0.075	0.148		
DEG	106 PPB	508 PPB	107 PPB	508 PPB		
Acetaldehydes	71 PPB	141 PPB	141 PPB	278 PPB		

Coolant medium : BFW
Coolant area : 6929 m^2
Overall U : 770 W/m^2 K
Coolant capacity : 53625 kW

Remarks: The preheating and cooling sections are included in the reactor volume and tube length. The space time is only concerns the reaction section.

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen, R.M. Boucke M.J. Sorgedrager, F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 10

REACTOR - SPECIFICATION SHEET

Equipment number	:	R201			Parallel:
Name	:	Carbonate reactor			1
		General Data			
Type	:	Trickle flow reactor			
Catalyst	:	Dow MSA-1			
Space time	:	23.4 min			
Volume	:	201.5 m^3			
Tubes:			Shell:		
- number	:		- inner diameter	:	4
- length/height	:	16 m	- outer diameter	:	4.04
- inner diameter	:		- material	:	SS
- outer diameter	:				
- pitch	:				
- material	:				

Process Conditions							
Stream details	Feed 204		Product 205		CO2 Feed 202		
Temp. [°C]	105		130		155		
Pressure [bara]	33		25		33		
Density [kg/m³]	1095		1086		41		
Mass flow [kg/h]	675184.0		695208.5		20024.5		
Vapor fraction	0		0		1		
Composition (frac)	mol	Wt	mol	wt	mol	wt	
EO	0.038	0.027					
MEG	0.407	0.412	0.406	0.4			
Ethylene	0.004	0.002	0.004	0.002			
Water	0.197	0.058	0.196	0.056			
Methane	0.003	695 PPM	0.003	675 PPM			
Oxygen	167 PPM	87 PPM	167 PPM	85 PPM			
Argon	32 PPM	21 PPM	31 PPM	20 PPM			
EC	0.344	0.494	0.381	0.533			
CO_2	0.005	0.004	0.009	0.006	20024.5	1	
DEG	0.002	0.003	0.002	0.003			
Acetaldehydes	297 PPB	214 PPB	296 PPB	207 PPB			

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 10

REACTOR - SPECIFICATION SHEET

Equipment number	:	R202		Parallel:
Name	:	Primary MEG reactor		1

General Data

Type : Packed bed reactor with chimney tray on top for V/L separation
Catalyst : Dow MSA-1
Space time : 58.6 min
Volume : 83.5 m^3

Tubes:			Shell:		
- number	:		- inner diameter	:	4m
- length/height	:	6.6 m	- outer diameter	:	4.03 m
- inner diameter	:		- material	:	CS
- outer diameter	:				
- pitch	:				
- material	:				

Process Conditions

Stream details	Feed 210		Product (V) 217		Product (L) 215		
Temp. [°C]	150		133		133		
Pressure [bara]	10		9.5		9.5		
Density [kg/m^3]	1041		12		822		
Mass flow [ks/h]	77472.1		9812.3		67659.9		
Vapor fraction	0		1		0		
Composition (frac)	mol	wt	mol	wt	mol	wt	
EO							
MEG	0.289	0.359	0.007	0.011	0.5	0.599	
Ethylene	0.002	919 PPM	0.01	0.007	59 PPM	32 PPM	
Water	0.432	0.155	0.146	0.065	0.325	0.113	
Methane	539 PPM	173 PPM	0.003	0.001	7 PPM	2 PPM	
Oxygen	28 PPM	18 PPM	176 PPM	141 PPM	278 PPB	172 PPB	
Argon	5 PPM	4 PPM	33 PPM	33 PPM	52 PPB	40 PPB	
EC	0.271	0.478	410 PPM	901 PPM	0.161	0.274	
CO_2	0.005	0.004	0.833	0.914	0.01	0.009	
DEG	0.001	0.003	16 PPM	41 PPM	0.003	0.006	
Acetaldehydes	206 PPB	182 PPB	802 PPB	881 PPB	95 PPB	80 PPB	

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 10

REACTOR - SPECIFICATION SHEET

Equipment number : R203 Parallel: 1
Name : Secondary MEG reactor

General Data

Type : Packed bed reactor with chimney tray on top for V/L separation
Catalyst : Dow MSA-1
Space time : 65.2 min
Volume : 83.1m^3

Tubes:		Shell:	
- number :		- inner diameter :	4 m
- length/height :	6.6 m	- outer diameter :	4.03 m
- inner diameter :		- material :	CS
- outer diameter :			
- pitch :			
- material :			

Process Conditions

Stream details	Feed 216		Product (V) 222		Product (L) 221		
Temp. [°C]	150		132		132		
Pressure [bara]	10		9.5		9.5		
Density [kg/m^3]	1018		12		823		
Mass flow [ks/h]	67930.9		9777.4		58153.5		
Vapor fraction	0		1		0		
Composition (frac)	mol	wt	mol	wt	mol	wt	
EO							
MEG	0.497	0.598	0.01	0.015	0.795	0.92	
Ethylene	58 PPM	32 PPM	320 PPM	213 PPM	2 PPM	1 PPM	
Water	0.33	0.115	0.082	0.035	0.189	0.063	
Methane	7 PPM	2 PPM	37 PPM	14 PPM	76 PPB	23 PPB	
Oxygen	276 PPB	171 PPB	2 PPM	1 PPM	3 PPB	2 PPB	
Argon	52 PPB	40 PPB	292 PPB	278 PPB	trace	trace	
EC	0.16	0.273	456 PPB	955 PPB	194 PPM	318 PPM	
CO_2	0.01	0.009	0.907	0.949	0.012	0.01	
DEG	0.003	0.006	17 PPM	42 PPM	0.003	0.007	
Acetaldehydes	94 PPB	80 PPB	335 PPB	351 PPB	42 PPB	35 PPB	

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

Equipment number	: C101			rev 3.00
Name	: EO Absorber			
	General Data			
Service	: absorbtion			
Column type	: structured packing			
Tray / Packing type	: mellapak 250X			
Stages				
- Theoretical	: 15			
- Feed	: 1, 15			
HETP [m]	: 2.1	Tray material	:	PP
Column diameter [m]	: 4.72	Column material	:	CS
Column height [m]	: 31.5			
Heating	: none			

Process Conditions

Stream details	Feed 112	Top 113	Bottom 124	Absorbent 127
Temp. [°C]	43.0	22.7	42.2	20
Pressure [bara]	19.9	20	20	20
Density [kg/m^3]	18	19	992	929
Vapor frac [-]	.997	1	0	0
Mass flow [kg/s]	223	212	319	308

Composition (frac)	Feed mol	Feed wt	Top mol	Top wt	Bottom mol	Bottom wt	Absorbent mol	Absorbent wt
EO	0.02	0.039	194 PPB	393 PPB	0.038	0.027	20 PPB	14 PPB
MEG			5 PPM	15 PPM	0.407	0.412	0.432	0.426
Ethylene	0.207	0.262	0.212	0.273	0.004	0.002	0.001	541 PPM
Water	0.007	0.006	725 PPM	599 PPM	0.197	0.058	0.196	0.056
Methane	0.63	0.454	0.648	0.477	0.003	695 PPM	400 PPM	102 PPM
Oxygen	0.051	0.073	0.052	0.077	167 PPM	87 PPM	21 PPM	11 PPM
Argon	0.01	0.018	0.01	0.019	32 PPM	21 PPM	4 PPM	2 PPM
EC	937 PPB	4 PPM	483 PPB	2 PPM	0.344	0.494	0.365	0.511
CO_2	0.075	0.148	0.076	0.154	0.005	0.004	0.003	0.002
DEG	107 PPB	508 PPB	4 PPB	22 PPB	0.002	0.003	0.002	0.003
Acetaldehydes	141 PPB	278 PPB	63 PPB	127 PPB	297 PPB	214 PPB	153 PPB	107 PPB

Column Internals

Packing

Type : Mellapak 250X

Material : PP

Fractional approach to max. capacity : 0.9

Max. capacity factor : 0.094 [m/sec]

Surface area : 250 [m^2/m^3]

Average p-drop/height : 3.31 [mbar/m]

Remarks: section pressure drop : 0.103 atm

Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

Equipment number : C301 rev 3.00
Name : Water removal

General Data

Service : distillation
Column type : tray
Tray / Packing type : sieve tray
Stages
- Theoretical : 6
- Actual : 10
- Feed : 7

Tray spacing [m] : .6096 Tray material : CS
Column diameter [m] : 2.77 Column material : CS
Column height [m] : 6.10
Heating : thermosyphon

Process Conditions

Stream details	Feed 302		Top (l) 313		Bottom 306		Top (g) 308	
Temp. [°C]	145.6		43.8		132.3		43.8	
Pressure [bara]	9.5		0.1		0.1		0.1	
Density [kg/m^3]	991		981		1011		<1	
Mass flow [kg/s]	16.16		4.52		7.13		4.52	
Composition (frac)	mol	Wt	mol	wt	mol	wt	mol	wt
EO								
MEG	0.795	0.92	0.004	0.012	0.994	0.992	0.004	0.012
Ethylene	2 PPM	1 PPM	10 PPM	14 PPM	trace	trace	10 PPM	14 PPM
Water	0.189	0.063	0.937	0.856	0.001	411 PPM	0.937	0.856
Methane	76 PPB	23 PPB	377 PPB	307 PPB	trace	trace	377 PPB	307 PPB
Oxygen	3 PPB	2 PPB	13 PPB	20 PPB	trace	trace	13 PPB	20 PPB
Argon	trace	Trace	2 PPB	5 PPB	trace	trace	2 PPB	5 PPB
EC	194 PPM	318 PPM			243 PPM	343 PPM		
CO_2	0.012	0.01	0.059	0.132	trace	trace	0.059	0.132
DEG	0.003	0.007			0.004	0.007		
Acetaldehydes	42 PPB	35 PPB	210 PPB	470 PPB	trace	trace	210 PPB	470 PPB

Column Internals

Trays
Sieve hole diameter [cm] : 1.27
Sieve hole area to active area ratio : 0.12
Weir length [m] : 2.02
Murphree eff. : 0.60
Downcomer area/column area : 0.10

Remarks:
Section pressure drop= 0.025 atm Reboiler E301 Reflux ratio = 0.57
Distillate vapor fraction= 50 mol% Condenser E302

Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

STORK® Stork Engineers & Contractors

Equipment number	: C302		rev 3.00
Name	: MEG purification		
	General Data		
Service	: Distillation, total condensor		
Column type	: structured packing		
Tray / Packing type	: Mellapak 250X		
Stages	:		
- Theoretical	: 11		
- Feed (actual)	: 3		
HETP [m]	: 2.1	Packing material:	SS
Column diameter [m]	: 2.92	Column material :	CS
Column height [m]	: 23.10		
Heating	: thermosyphon		

Process Conditions

Stream details	Feed 307		Top 305		Bottom 309			
Temp. [°C]	132.3		129.8		164.4			
Pressure [bara]	0.1		0.09		0.09			
Density [kg/m^3]	1010		1013		930			
Mass flow [kg/s]	7.188		7.129		0.058			
Composition (frac)	mol	Wt	mol	wt	mol	wt	mol	wt
EO								
MEG	0.994	0.992	0.998	0.999	0.192	0.122		
Ethylene	trace	Trace						
Water	0.001	411 PPM	0.001	415 PPM	trace	trace		
Methane	trace	Trace						
Oxygen	trace	Trace						
Argon	trace	Trace						
EC	243 PPM	343 PPM	104 PPM	147 PPM	0.027	0.024		
CO_2	trace	Trace						
DEG	0.004	0.007	251 PPM	429 PPM	0.782	0.853		
Acetaldehydes	trace	Trace						

Column Internals

Packing

Type : Mellapak 250X

Material : SS

Fractional approach to max. capacity : 0.62

Max. capacity factor : 0.108 [m/sec]

Surface area : 250 [m^2/m^3]

Average p-drop/height : 1.28 [mbar/m]

Remarks:

section pressure drop= 0.024 atm   Reboiler E303

Reflux ratio = 0.314   Condenser E304

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

Stork Engineers & Contractors

Equipment number	:	C401		rev 3.00
Name	:	Raw EO Stripper		
		General Data		
Service	:	heat stripping, pressure swing		
Column type	:	structured packing		
Tray / Packing type	:	Mellapak 250Y		
Stages				
- Theoretical	:	10		
- Feed	:	1		
HETP [m]	:	2.9	Tray material :	PVDF
Column diameter [m]	:	3.38	Column material :	SS
Column height [m]	:	20.30		
Heating	:	thermosyphon		

Process Conditions

Stream details	Feed 401		Top 402		Bottom 413		Reflux	
Temp. [°C]	41.8		62.2		137.1			
Pressure [bara]	20		1		1			
Density [kg/m^3]	1.122		.001		1.064			
Mass flow [kg/s]	131.4		4.665		126.75			
Composition (frac)	mol	wt	Mol	wt	mol	wt	mol	wt
EO	0.038	0.027	0.688	0.773	trace	trace		
MEG	0.407	0.412	0.001	0.002	0.431	0.427		
Ethylene	0.004	0.002	0.072	0.051	trace	trace		
Water	0.197	0.058	0.09	0.041	0.203	0.058		
Methane	0.003	695 PPM	0.048	0.02	trace	trace		
Oxygen	167 PPM	87 PPM	0.003	0.002	trace	trace		
Argon	32 PPM	21 PPM	568 PPM	579 PPM	trace	trace		
EC	0.344	0.494	105 PPM	235 PPM	0.364	0.512		
CO_2	0.005	0.004	0.098	0.11	trace	trace		
DEG	0.002	0.003	897 PPB	2 PPM	0.002	0.003		
Acetaldehydes	297 PPB	214 PPB	5 PPM	6 PPM	trace	trace		

Column Internals

Packing

Type : Mellapak 250Y
Material : PVDF
Fractional approach to max. capacity : 0.8
Max. capacity factor : 0.059 [m/sec]
Surface area : 250 [m^2/m^3]
Average p-drop/height : 1.13 [mbar/m]

Remarks:
pressure drop= 0.022 atm Reboiler E401
boilup ratio= 0.308

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

Stork Engineers & Contractors

Equipment number	: C402			rev 3.00
Name	: secondary EO absorber			
	General Data			
Service	: absorbtion			
Column type	: structured packing			
Tray / Packing type	: Mellapak 250Y			
Stages				
- Theoretical	: 9			
- Feed	: 1, 9			
HETP [m]	: 2.9	Tray material	:	PP
Column diameter [m]	: 0.79	Column material :		ss
Column height [m]	: 23.2			
Heating	: none			

Process Conditions

Stream details	Feed 424		Top 421		Bottom 406		Absorbent 416	
Temp. [°C]	43		46.6		50.3		43	
Pressure [bara]	10		10		10.0		10	
Density [kg/m^3]	80		9		919.0		1175	
Mass flow [kg/s]	7.876		0.927		18.060		11.111	
Composition (frac)	mol	Wt	mol	wt	mol	wt	mol	wt
EO	0.758	0.818	101 PPB	135 PPB	0.429	0.357		
MEG	642 PPM	976 PPM	50 PPM	95 PPM	0.211	0.247	0.408	0.401
Ethylene	0.053	0.037	0.317	0.27	0.005	0.003	0.002	0.001
Water	0.057	0.025	0.004	0.002	0.134	0.045	0.197	0.056
Methane	0.031	0.012	0.207	0.101	733 PPM	222 PPM	762 PPM	193 PPM
Oxygen	0.002	0.001	0.013	0.013	32 PPM	19 PPM	40 PPM	20 PPM
Argon	361 PPM	353 PPM	0.002	0.003	6 PPM	4 PPM	7 PPM	5 PPM
EC	65 PPM	139 PPM	6 PPM	15 PPM	0.198	0.329	0.383	0.534
CO_2	0.097	0.105	0.456	0.611	0.021	0.017	0.007	0.005
DEG	553 PPB	1 PPM	44 PPB	143 PPB	915 PPM	0.002	0.002	0.003
Acetaldehydes	6 PPM	6 PPM	377 PPB	504 PPB	3 PPM	3 PPM	291 PPB	203 PPB

Column Internals

Packing

Type : Mellapak 250Y

Material : PP

Fractional approach to max. capacity : 0.8

Max. capacity factor : 0.025 [m/sec]

Surface area : 250 [m^2/m^3]

Average p-drop/height : 0.18 [mbar/m]

Remarks: Pressure drop: 0.005 atm

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

Equipment number	:	C403		rev 3.00
Name	:	secondary EO stripper		

General Data

Service	:	Stripping w/ heat			
Column type	:	Structured packing			
Tray / Packing type	:	Mellapak 250Y			
Stages					
- Theoretical	:	5			
- Feed	:	1			
HETP [m]	:	2.9	Tray material	:	PP
Column diameter [m]	:	1.77	Column material	:	ss
Column height [m]	:	11.60			
Heating	:	thermosyphon			

Stork Engineers & Contractors

Process Conditions

Stream details	Feed 406		Top 407		Bottom 417			
Temp. [°C]	50.3		22.8		109.6			
Pressure [bara]	10		0.5		0.5			
Density [kg/m^3]	919.0		4.0		858.0			
Mass flow [kg/s]	18.060		6.923		11.137			
Composition (frac)	mol	wt	Mol	wt	mol	wt	mol	wt
EO	0.429	0.357	0.905	0.931	538 PPB	381 PPB		
MEG	0.211	0.247	80 PPM	116 PPM	0.402	0.4		
Ethylene	0.005	0.003	0.011	0.007	trace	trace		
Water	0.134	0.045	0.038	0.016	0.22	0.064		
Methane	733 PPM	222 PPM	0.002	579 PPM	trace	trace		
Oxygen	32 PPM	19 PPM	66 PPM	50 PPM	trace	trace		
Argon	6 PPM	4 PPM	12 PPM	11 PPM	trace	trace		
EC	0.198	0.329	7 PPM	15 PPM	0.377	0.533		
CO_2	0.021	0.017	0.044	0.045	trace	trace		
DEG	915 PPM	0.002	59 PPB	146 PPB	0.002	0.003		
Acetaldehydes	3 PPM	3 PPM	7 PPM	7 PPM	trace	trace		

Column Internals

Packing

Type : Mellapak 250Y

Material : PVDF

Fractional approach to max. capacity : 0.62

Max. capacity factor : 0.056 [m/sec]

Surface area : 250 [m^2/m^3]

Average p-drop/height : 0.68 [mbar/m]

Remarks:

pressure drop= 0.0077 atm Reboiler E405

boilup ratio= 0.36

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 11

DISTILLATION COLUMN - SPECIFICATION SHEET

Stork Engineers & Contractors

Equipment number : C404 rev 3.00
Name : HE separator

General Data

Service : distillation, total condensor
Column type : Tray
Tray / Packing type : sieve
Stages - Actual : 19
- Theoretical : 15
- Feed (actual) : 8
Tray height [m] : .6096 Tray material : SS
Column diameter [m] : 1.22 Column material : SS
Column height [m] : 11.58
Heating : thermosyphon

Process Conditions

Stream details	Feed 410	Top 419	Bottom 418	Reflux
Temp. [°C]	61.1	29.2	120.6	
Pressure [bara]	5	2.0	2.0	
Density [kg/m^3]	813	855	896	
Mass flow [kg/s]	3.712	3.604	0.108	

Composition (frac)	mol	Wt	mol	wt	mol	wt	mol	wt
EO	0.932	0.971	1	1	138 PPM	335 PPM		
MEG	148 PPM	217 PPM	2 PPB	3 PPB	0.002	0.007		
Ethylene	trace	Trace						
Water	0.067	0.029	23 PPM	9 PPM	0.997	0.991		
Methane	trace	Trace						
Oxygen	trace	Trace						
Argon	trace	Trace						
EC	13 PPM	28 PPM	trace	trace	199 PPM	968 PPM		
CO_2	trace	Trace						
DEG	109 PPB	272 PPB	trace	trace	2 PPM	9 PPM		
Acetaldehydes	8 PPM	8 PPM	8 PPM	8 PPM	3 PPB	7 PPB		

Column Internals

Trays

Number of trays : 19
Murphree efficiency : 0.8
Sieve hole area to active area ratio : 0.12
Weir length [m] : 0.88
Downcomer area/column area 0.1
Side downcomer vel. [m/sec] 0.0044

Remarks: pressure drop= 0.041 atm
reflux ratio: 0.121 Reboiler E406
boilup ratio: 7.27 Condenser E407

Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

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DISTILLATION COLUMN - SPECIFICATION SHEET

Equipment number	:	C405		rev 3.00
Name	:	LE separator		
		General Data		
Service	:	distillation, no condensor		
Column type	:	tray		
Tray / Packing type	:	sieve		
Stages - Actual	:	10		
- Theoretical	:	8		
- Feed	:	1		
Tray spacing [m]	:	.6096	Tray material :	SS
Column diameter [m]	:	0.91	Column material :	SS
Column height [m]	:	6.10		
Heating	:	thermosyphon		

Process Conditions

Stream details	Feed 409		Top 411		Bottom 410			
Temp. [°C]	56.8		55.8		61.1			
Pressure [bara]	2.5		5.0		5.0			
Density [kg/m^3]	4		8		809			
Mass flow [kg/s]	6.923		3.211		3.712			
Composition (frac)	mol	Wt	mol	wt	mol	wt	mol	wt
EO	0.905	0.931	0.872	0.884	0.932	0.971		
MEG	80 PPM	116 PPM	47 PPB	67 PPB	148 PPM	217 PPM		
Ethylene	0.011	0.007	0.024	0.016	trace	trace		
Water	0.038	0.016	0.004	0.002	0.067	0.029		
Methane	0.002	579 PPM	0.003	0.001	trace	trace		
Oxygen	66 PPM	50 PPM	145 PPM	107 PPM	trace	trace		
Argon	12 PPM	11 PPM	27 PPM	25 PPM	trace	trace		
EC	7 PPM	15 PPM	trace	1 PPB	13 PPM	28 PPM		
CO_2	0.044	0.045	0.096	0.097	trace	trace		
DEG	59 PPB	146 PPB	trace	trace	109 PPB	272 PPB		
Acetaldehydes	7 PPM	7 PPM	6 PPM	7 PPM	8 PPM	8 PPM		

Column Internals

Trays

Number of trays	:	10
Murphree eff.	:	0.80
Sieve hole area to active area ratio	:	0.12
Weir length [m]	:	0.69
Downcomer area/column area		0.11
Side downcomer vel. [m/sec]		0.110

Remarks:
boil-up ratio: 0.78 Reboiler E410
section pressure drop= 0.083 atm

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

STORK®
Stork Engineers & Contractors

Equipment number	: E101	In series :	
Name	: Feed Preheater	In parallel :	4

General Data

Service : Heat exchanger
Type : Fixed tubes
Position : Horizontal
Capacity : 31901 kW
Heat exchange area : 3973 m^2
Overall heat transfer coefficient : 110 W/m^2 K
Log. mean temp. difference (LMTD) : 73 K
Passes tube side : 1
Passes shell side : 1
Correction factor LMTD : 1

Process Conditions

			Shell side	Tube side
Medium		:	Reactor out	Reactor in
Temp. in	[°C]		180	34
Temp.	[°C]	:	110	109.6
Pressure	[bara]	:	19.9	19.9
Density	[kg/m^3]	:	12	19
Mass flow	[kg/hr]	:	802142	802130
Vapor fraction		:	1	1
Average specific heat	[kJ/kg K]	:	2.132	1.836
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	ss	ss

Remarks:
Four exchangers parallel

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E104	In series	:
Name	: Effluent Cooler	In parallel :	4

General Data

Service	: Heat exchanger
Type	: Fixed tubes
Position	: Horizontal
Capacity	: 29665 kW
Heat exchange area	: 825 m^2
Overall heat transfer coefficient	: 300 $W/m^2 K$
Log. mean temp. difference (LMTD)	: 30 K
Passes tube side	: 1
Passes shell side	: 1
Correction factor LMTD	: 1

STORK®
Stork Engineers & Contractors

Process Conditions

			Shell side	Tube side
Medium		:	Cooling water	liquid
Temp. in	[°C]		33	110
Temp.	[°C]	:	43	43
Pressure	[bara]	:	1	19.9
Density	[kg/m^3]	:	998	14.0
Mass flow	[kg/hr]	:	2554898	1021230
Vapor fraction		:	0	1
Average specific heat	[kJ/kg K]	:	4.2	21.961
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:
Four exchangers parallel

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Stork Engineers & Contractors

Equipment number	:	E105a	In series :	1
Name	:	Absorbent cooler	In parallel :	~~1~~

General Data

Service	:	Heat exchanger
Type	:	Fixed tubes
Position	:	Horizontal
Capacity	:	15466 kW
Heat exchange area	:	477 m^2
Overall heat transfer coefficient	:	900 W/m^2 K
Log. mean temp. difference (LMTD)	:	36 K
Passes tube side	:	1
Passes shell side	:	1
Correction factor LMTD	:	1

Process Conditions

			Shell side	Tube side
Medium		:	Chilled water	liquid
Temp. in	[°C]		4	43
Temp.	[°C]	:	10	20
Pressure	[bara]	:	1	20
Density	[kg/m^3]	:	998	1100
Mass flow	[kg/hr]	:	2210340	1108380
Vapor fraction		:	0	0
Average specific heat	[kJ/kg K]	:	4.2	2.369
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

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HEAT EXCHANGER - SPECIFICATION SHEET

STORK®
Stork Engineers & Contractors

Equipment number : E105b
Name : Absorbent cooler
In series : 1
In parallel : ~~1~~

General Data

Service : Heat exchanger
Type : Fixed tubes
Position : Horizontal
Capacity : 23707 kW
Heat exchange area : 825 m^2
Overall heat transfer coefficient : 900 W/m^2 K
Log. mean temp. difference (LMTD) : 32 K
Passes tube side : 1
Passes shell side : 1
Correction factor LMTD : 1

Process Conditions

			Shell side	Tube side
Medium		:	Cooling water	liquid
Temp. in	[°C]	:	33	117
Temp.	[°C]	:	43	43
Pressure	[bara]	:	1	20
Density	[kg/m^3]	:	998	1100
Mass flow	[kg/hr]	:	2048650	1108380
Vapor fraction		:	0	0
Average specific heat	[kJ/kg K]	:	4.2	2.369
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen, M.J. Sorgedrager, R.M. Boucke, F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

STORK®
Stork Engineers & Contractors

Equipment number	: E201	In series	:
Name	: Preheater	In parallel	: 3

General Data

Service	: Heat exchanger
Type	: Fixed Tube
Position	: Horizontal
Capacity	: 26108 kW
Heat exchange area	: 2950m^2
Overall heat transfer coefficient	: 400 W/m^2 K
Log. mean temp. difference (LMTD)	: 22
Passes tube side	: 1
Passes shell side	: 1
Correction factor LMTD	: 1

Process Conditions

			Shell side	Tube side
Medium		:	liquid	liquid
Temp. in	[°C]	:	42	117
Temp. out	[°C]	:	105	78
Pressure	[bara]	:	20	33
Density	[kg/m^3]	:	1100	1157
Mass flow	[kg/hr]	:	1108380	675184
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.100
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks: 3 exchangers
parallel

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

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HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E202	In series	:	
Name	: Preheater	In parallel	:	1

General Data

Service	: Heat Exchanger
Type	: Fixed Tube
Position	: Horizontal
Capacity	: 3473 kW
Heat exchange area	: 111 m^2
Overall heat transfer coefficient	: 1000 $W/m^2 K$
Log. mean temp. difference (LMTD)	: 31
Passes tube side	: 1
Passes shell side	: 1
Correction factor LMTD	: 1

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	160	89
Temp. out	[°C]	:	160	150
Pressure	[bara]	:	7.8	10
Density	[kg/m^3]	:	-	1105
Mass flow	[kg/hr]	:	6008	77472
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.495
Heat of evap./cond.	[kJ/kg]	:	2084.3	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

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HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E203	In series	: 1
Name	: Preheater	In parallel	: 1

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General Data

Service	: Heat exchanger
Type	: Fixed Tube
Position	: Horizontal
Capacity	: 406 kW
Heat exchange area	: 24 m^2
Overall heat transfer coefficient	: 1000 W/m^2 K
Log. mean temp. difference (LMTD)	: 17
Passes tube side	: 1
Passes shell side	: 1
Correction factor LMTD	: 1

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	160	133
Temp. out	[°C]	:	160	150
Pressure	[bara]	:	7.8	10
Density	[kg/m^3]	:	-	822
Mass flow	[kg/hr]	:	703	67931
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	-
Heat of evap./cond.	[kJ/kg]	:	2084.3	-
Material		:	ss	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

STORK®
Stork Engineers & Contractors

Equipment number	: E301	In series	: 1
Name	: Reboiler C301	In parallel	: ~~1~~

General Data	
Service	: Reboiler
Type	: Thermosyphon
Position	: Vertical
Capacity	: 3500 kW
Heat exchange area	: 100 m^2
Reduced temperature	: 0.41
Heat flux	: 35000 W/m^2

Process Conditions				
			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	160	-
Temp. out	[°C]	:	160	132
Pressure	[bara]	:	7.8	0.1
Density	[kg/m^3]	:	-	1010
Mass flow	[kg/hr]	:	6054	84636
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:		2.819
Heat of evap./cond.	[kJ/kg]	:	2084.3	-
Material		:	cs	cs

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Stork Engineers & Contractors

Equipment number	: E302	In series	: 1
Name	: Condenser C301	In parallel	: ~~1~~

General Data

Service	: Condenser
Type	: Fixed tubes
Position	: vertical
Capacity	: 4413 kW
Heat exchange area	: 337 m^2
Overall heat transfer coefficient	: 2000 W/m^2 K

Process Conditions

			Shell side	Tube side
Medium		:	Chilled water	Liquid
Temp. in	[°C]	:	4	-
Temp. out	[°C]	:	10	14
Pressure	[bara]	:	1	0.1
Density	[kg/m^3]	:	998	-
Mass flow	[kg/hr]	:	630687	4291
Vapor fraction		:	0	0.5
Average specific	[kJ/kg K]	:	4.2	2.704
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	cs

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E303	In series	: 1
Name	: Reboiler C302	In parallel	: ~~1~~

STORK®
Stork Engineers & Contractors

General Data

Service	: Reboiler
Type	: Thermosyphon
Position	: Vertical
Capacity	: 8889 kW
Heat exchange area	: 148 m^2
Reduced temperature	: 0.55
Heat flux	: 60000 W/m^2

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	230	-
Temp. out	[°C]	:	230	164
Pressure	[bara]	:	27.6	0.1
Density	[kg/m^3]	:	-	930
Mass flow	[kg/hr]	:	17663	275.7
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.898
Heat of evap./cond.	[kJ/kg]	:	1785.1	-
Material		:	cs	cs

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E304	In series	: 1
Name	: Condenser C302	In parallel	: 1

General Data

Service	: Condenser
Type	: Vertical
Position	: Fixed Tubes
Capacity	: 8944 kW
Heat exchange area	: 256 m^2
Reduced Temperature	: 0.35 W/m^2 K
Heat flux	: 35000

Process Conditions

Medium		:	Evaporating BFW water	Liquid
Temp. in	[°C]	:	120	-
Temp. out	[°C]	:	120	130
Pressure	[bara]	:	1.9	0.1
Density	[kg/m^3]	:	-	-
Mass flow	[kg/hr]	:	11899	33622.4
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	4.2	2.807
Heat of evap./cond.	[kJ/kg]	:	2204.67	-
Material		:	cs	cs

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen, R.M. Boucke M.J. Sorgedrager, F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E401	In series	: 1
Name	: Reboiler C401	In parallel	: 1

STORK
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General Data

Service	: Reboiler
Type	: Thermosyphon
Position	: Vertical
Capacity	: 29806 kW
Heat exchange area	: 497 m^2
Reduced temperature	: 0.51
Heat flux	: 60000 W/m^2

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	230	-
Temp. out	[°C]	:	230	136
Pressure	[bara]	:	27.6	1.0
Density	[kg/m^3]	:	-	847
Mass flow	[kg/hr]	:	59226	6762
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.422
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen R.M. Boucke M.J. Sorgedrager F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

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Equipment number	:	E402	In series :	1
Name	:	Post- Cooler	In parallel :	1

General Data

Service	:	Cooler
Type	:	Fixed tube
Position	:	Horizontal
Capacity	:	542 kW
Heat exchange area	:	49 m^2
Overall heat transfer coefficient	:	300 W/m^2 K
Log. mean temp. difference (LMTD)	:	37
Passes tube side	:	1
Passes shell side	:	1
Correction factor LMTD	:	1

Process Conditions

			Shell side	Tube side
Medium		:	Cooling water	Vapor
Temp. in	[°C]	:	33	134
Temp. out	[°C]	:	43	43
Pressure	[bara]	:	1	2.5
Density	[kg/m^3]	:	998	3
Mass flow	[kg/hr]	:	46711	16793
Vapor fraction		:	0	1
Average specific heat	[kJ/kg K]	:	4.2	1.469
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E403	In series	: 1
Name	: Post cooler	In parallel	: 1

General Data

Service	: Cooler
Type	: Fixed tube
Position	: Horizontal
Capacity	: 4636kW
Heat exchange area	: 431 m^2
Overall heat transfer coefficient	: 300 W/ m^2 K
Log. mean temp. difference (LMTD)	: 36
Passes tube side	: 1
Passes shell side	: 1
Correction factor LMTD	: 1

Process Conditions

			Shell side	Tube side
Medium		:	Cooling water	Vapor
Temp. in	[°C]	:	33	131
Temp. out	[°C]	:	43	43
Pressure	[bara]	:	1	10
Density	[kg/m^3]	:	998	12
Mass flow	[kg/hr]	:	399253	28352
Vapor fraction		:	0	1
Average specific	[kJ/kg K]	:	4.2	1.462
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	:	E404	In series	: 1
Name	:	Absorbent cooler	In parallel	: ~~1~~

General Data

Service	:	Cooler
Type	:	Fixed tube
Position	:	Horizontal
Capacity	:	1359 kW
Heat exchange area	:	67 m^2
Overall heat transfer coefficient	:	750 W/ m^2 K
Log. mean temp. difference (LMTD)	:	27
Passes tube side	:	1
Passes shell side	:	1
Correction factor LMTD	:	1

Process Conditions

			Shell side	Tube side
Medium		:	Cooling water	Liquid
Temp. in	[°C]	:	33	100
Temp. out	[°C]	:	43	43
Pressure	[bara]	:	1	10
Density	[kg/m^3]	:	998	1120
Mass flow	[kg/hr]	:	117079	40000
Vapor fraction		:	0	0
Average specific heat	[kJ/kg K]	:	4.2	2.275
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

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HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E405	In series	: 1
Name	: Reboiler C403	In parallel	: 1

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General Data

Service	: Reboiler
Type	: Thermosyphon
Position	: Vertical
Capacity	: 5000 kW
Heat exchange area	: 143 m^2
Reduced temperature	: 0.37
Heat flux	: 35000 W/m^2

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	160	-
Temp. out	[°C]	:	160	109
Pressure	[bara]	:	7.8	0.5
Density	[kg/m^3]	:	-	858
Mass flow	[kg/hr]	:	8649	40094
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.279
Heat of evap./cond.	[kJ/kg]	:	2084.3	-
Material		:	cs	ss

Remarks:

Case identification : normal	Design
Overcapacity :	Revision :

Designers: F. Kleijn van Willigen	R.M. Boucke	Project ID: CPD3244 / P80511 CA30
M.J. Sorgedrager	F.A. Sheldon	Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

STORK
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Equipment number	: E406	In series	: 1
Name	: Reboiler C404	In parallel :	1

General Data

Service	: Reboiler
Type	: Thermosyphon
Position	: Vertical
Capacity	: 2031 kW
Heat exchange area	: 51 m^2
Reduced temperature	: 0.39
Heat flux	: 40000 W/m^2

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	160	-
Temp. out	[°C]	:	160	121
Pressure	[bara]	:	7.8	2
Density	[kg/m^3]	:	-	896
Mass flow	[kg/hr]	:	3512	387
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.208
Heat of evap./cond.	[kJ/kg]	:	2084.3	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

STORK®
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Equipment number	: E407		In series	: 1
Name	: Condenser C404		In parallel	: 1

General Data

Service	:	Condenser
Type	:	Fixed Tubes
Position	:	Vertical
Capacity	:	2300 kW
Heat exchange area		52 m^2
Overall heat transfer coefficient	:	2000 W/m^2 K

Process Conditions

			Shell side	Tube side
Medium		:	Chilled water	Liquid
Temp. in	[°C]		4	-
Temp. out	[°C]	:	10	29
Pressure	[bara]	:	1	2
Density	[kg/m^3]	:	998	26
Mass flow	[kg/hr]	:	328702	12976
Vapor fraction		:	0	0
Average specific heat	[kJ/kg K]	:	4.2	2.208
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E408	In series	: 1
Name	: Post cooler K404	In parallel	: 1

General Data

Service	: Heat exchanger
Type	: Fixed Tube
Position	: Horizontal
Capacity	: 744 kW
Heat exchange area	: 48 m^2
Overall heat transfer coefficient	: 300 W/m^2 K
Log. mean temp. difference (LMTD)	: 51 K
Passes tube side	: 1
Passes shell side	: 1
Correction factor LMTD	: 1

Process Conditions

			Shell side	Tube side
Medium		:	Cooling water	Vapor
Temp. in	[°C]	:	33	138
Temp. out	[°C]	:	43	57
Pressure	[bara]	:	1	2.5
Density	[kg/m^3]	:	998	3
Mass flow	[kg/hr]	:	64053	24922
Vapor fraction		:	0	1.0
Average specific heat	[kJ/kg K]	:	4.2	2.072
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00
Designers: F. Kleijn van Willigen, R.M. Boucke M.J. Sorgedrager, F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E409	In series :	1
Name	: Post cooler K404	In parallel :	

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General Data	
Service	: Reboiler
Type	: Thermosyphon
Position	: Horizontal
Capacity	: 4135kW
Heat exchange area	: 188 m^2
Overall heat transfer coefficient	: 750W/m^2 K
Log. mean temp. difference (LMTD)	: 29 K

Process Conditions

			Shell side	Tube side
Medium		:	water	liquid
Temp. in	[°C]	:	33	108
Temp.	[°C]	:	43	43
Pressure	[bara]	:	1	5
Density	[kg/m^3]	:	998	7
Mass flow	[kg/hr]	:	356150	24922
Vapor fraction		:	0	1
Average specific heat	[kJ/kg K]	:	4.2	1.291
Heat of evap./cond.	[kJ/kg]	:	-	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

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HEAT EXCHANGER - SPECIFICATION SHEET

Equipment number	: E410	In series	: 1
Name	: Reboiler C405	In parallel	: 1

General Data

Service	: Reboiler
Type	: Thermosyphon
Position	: Vertical
Capacity	: 1583 kW
Heat exchange area	: 26 m^2
Reduced temperature	: 0.31
Heat flux	: 60000 W/m^2

Stork Engineers & Contractors

Process Conditions

			Shell side	Tube side
Medium		:	Condensing steam	Liquid
Temp. in	[°C]	:	160	-
Temp. out	[°C]	:	160	61
Pressure	[bara]	:	7.8	5
Density	[kg/m^3]	:	-	809
Mass flow	[kg/hr]	:	2739	13362.9
Vapor fraction		:	-	0
Average specific heat	[kJ/kg K]	:	-	2.644
Heat of evap./cond.	[kJ/kg]	:	2084.3	-
Material		:	cs	ss

Remarks:

Case identification : normal Overcapacity :	Design Revision : 3.00

Designers: F. Kleijn van Willigen M.J. Sorgedrager	R.M. Boucke F.A. Sheldon	Project ID: CPD3244 / P80511 CA30 Date : January 1, 2000

Appendix 13 - EO Reactor Design

Reactor Specifications

Inner tube diameter [m]	$d_{ti} := 0.04$
Reaction tube length [m]	$d_{to} := 0.044$
Outer tube diameter [m]	$L_{tube} := 8$
Tube pitch [m]	$p_t := 1.25 \cdot d_{to}$

Catalyst Specifications

Particle diameter [m]	$d_p := 0.008$
Reactivity [kg EO/m3 cat h]	$R_{cat} := 250$

Flow Specifications

EO massflow [kg/h]	IN	$M_{i.EO} := 0.299$
	OUT	$M_{o.EO} := 31494.129$
	PRODUCTION	$M_{p.EO} := M_{o.EO} - M_{i.EO}$
		$M_{p.EO} = 3.149 \cdot 10^4$
Total massflow [kg/h]		$F_{mt} := 802142.2$
Total Volume flow [m³/h]		$F_v := 76293.2$
Density [kg/m³]		$\rho := \frac{F_{mt}}{F_v}$ $\quad \rho = 10.514$

Calculations:

Single tube Area [m²]	$A_{tube} := \frac{1}{4} \cdot \pi \cdot d_{ti}^2$	$A_{tube} = 1.257 \cdot 10^{-3}$
Single tube Volume [m³]	$V_{tube} := \frac{1}{4} \cdot \pi \cdot d_{ti}^2 \cdot L_{tube}$	$V_{tube} = 0.01$
Total Reaction Volume [m³]	$V_{rt} := \frac{M_{p.EO}}{R_{cat}}$	$V_{rt} = 125.975$
Volume per reactor [m³]	$V_r := \left(\frac{1}{2}\right) V_{rt}$	$V_r = 62.988$
Number of tubes per reactor [-]	$N_t := \frac{V_r}{V_{tube}}$	$N_t = 6.265 \cdot 10^3$
Tube bundle diameter per reactor [m]	$D_{tb} := d_{to} \cdot \left(\frac{N_t}{0.215}\right)^{\frac{1}{2.207}}$	$D_{tb} = 4.638$
Shell or Reactor diameter [m]	$D_r := 0.5 \cdot p_t + D_{tb}$	$D_r = 4.666$
Retention time [sec]	$T_r := \frac{3600 \cdot V_{rt}}{F_v}$	$T_r = 5.944$
Total outer tube area per reactor[m²]	$A_{tot.tube} := \pi \cdot d_{to} \cdot N_t \cdot L_{tube}$	$A_{tot.tube} = 6.929 \cdot 10^3$

Pressure Drop:

Superficial velocity [m/s] $v_s := \frac{0.5 \cdot F_v}{A_{tube} \cdot N_t \cdot 3600}$ $v_s = 1.346$

Superficial mass velocity [kg/m²s] $v_{ms} := v_s \cdot \rho$ $v_{ms} = 14.15$

Bed Porosity [-] $\varepsilon_{bed} := 0.42$

Viscosity [Ns/m²] $\mu := 2.01 \cdot 10^{-5}$

Ergun

$$\Delta P := \frac{-v_{ms} \cdot (1 - \varepsilon_{bed})}{\rho \cdot d_p \cdot \varepsilon_{bed}^{3}} \cdot \left[\frac{150 \cdot (1 - \varepsilon_{bed}) \cdot \mu}{d_p} + 1.75 \cdot v_{ms} \right] \cdot L_{tube}$$

$\Delta P = -2.632 \cdot 10^{5}$ $P_0 := 22$

$P := \frac{\Delta P}{101300} + P_0$ $P = 19.402$

Heatsections

E102

Length tubes [m] $L_{102} := 2.2235$ $V_{102} := \left(\frac{1}{4} \cdot \pi \cdot d_{ti}^{2}\right) \cdot L_{102} \cdot N_t$ $V_{102} = 17.507$

E103

Length tubes [m] $L_{103} := 1.3125$ $V_{103} := \left(\frac{1}{4} \cdot \pi \cdot d_{ti}^{2}\right) \cdot L_{103} \cdot N_t$ $V_{103} = 10.334$

Total reactor volume [m³] $V_t := V_r + V_{102} + V_{103}$ $V_t = 90.828$

Total tube length [m] $L_t := L_{tube} + L_{102} + L_{103}$ $L_t = 11.536$

Appendix 13 - Carbonate Reactor Design

Reactor Specifications

Reactor Diameter [m] $D_r := 4$

Cross sectional Area [m²] $A_r := \frac{1}{4}\cdot\pi\cdot D_r^2$ $A_r = 12.566$

Bed Porosity [-] $\varepsilon_{bed} := 0.34$

Catalyst Specifications

Particle diameter [m] $d_p := 0.01$

Reactivity [kg EO/m3 cat h] $R_{cat} := 300$

Flow Specifications

EO massflow [kg/h] IN $M_{i.EO} := 18518.383$

OUT $M_{o.EO} := 0$

PRODUCTION $M_{p.EO} := M_{o.EO} - M_{i.EO}$

$M_{p.EO} = -1.852\cdot10^4$

Total massflow [kg/h] $F_{mt} := 695208.5$

Total Volume flow [m³/h] $F_v := 616.5$ $F_{v.out} := 639.9$

Density [kg/m³] $\rho := \frac{F_{mt}}{F_v}$ $\rho = 1.128\cdot10^3$

Calculations:

Total Reaction Volume [m³] $V_{rt} := \frac{-M_{p.EO}}{R_{cat}}$ $V_{rt} = 61.728$

Totaal Reactor Volume [m³] $V_{Rt} := \frac{V_{rt}}{\varepsilon_{bed}}$ $V_{Rt} = 181.553$

Corrected volume for gas volume [m³] $V_R := V_{Rt} + 20$ $V_R = 201.553$

Reactor height [m] $H_r := \frac{V_R}{A_r}$ $H_r = 16.039$

Retention Time [min] $T_r := 60\cdot\frac{V_R}{F_{v.out}}$ $T_r = 18.899$

Pressure Drop:

Superficial velocity [m/s] $v_s := \frac{F_v}{A_r \cdot 3600}$ $v_s = 0.014$

Superficial mass velocity [kg/m²s] $v_{ms} := v_s \cdot \rho$ $v_{ms} = 15.367$

Viscosity [Ns/m²] $\mu := 0.273$

Ergun

$$\Delta P := \frac{-v_{ms} \cdot \left(1 - \varepsilon_{bed}\right)}{\rho \cdot d_p \cdot \varepsilon_{bed}^{3}} \cdot \left[\frac{150 \cdot \left(1 - \varepsilon_{bed}\right) \cdot \mu}{d_p} + 1.75 \cdot v_{ms}\right] \cdot H_r \quad \text{Pa}$$

$\Delta P = -1.002 \cdot 10^{6}$ Pa $P_0 := 33$ Bar

$P := \frac{\Delta P}{101300} + P_0$ $P = 23.11$ bar

Appendix 13 - MEG Reactor1 Design

Reactor Specifications

Reactor Diameter [m] $D_r := 4$

Cross sectional Area [m²] $A_r := \frac{1}{4}\cdot\pi\cdot D_r^{\,2}$ $A_r = 12.566$

Bed Porosity [-] $\varepsilon_{bed} := 0.34$

Catalyst Specifications

Particle diameter [m] $d_p := 0.01$

Reactivity [kg EC/m3 cat./h] $R_{cat} := 750$

Flow Specifications

EC massflow [kg/h] IN $M_{i.EC} := 37028.237$

OUT $M_{o.EC} := 18514.119$

PRODUCTION $M_{p.EC} := M_{o.EC} - M_{i.EC}$

$M_{p.EC} = -1.851\cdot 10^{4}$

Total massflow [kg/h] $F_{mt} := 77472.1$

$F_v := 74.4$ $F_{v.out} := 922$

Total Volume flow [m³/h]

Density [kg/m³] $\rho := \frac{F_{mt}}{F_v}$ $\rho = 1.041\cdot 10^{3}$

Calculations:

Total Reaction Volume [m³] $V_{rt} := \frac{-M_{p.EC}}{R_{cat}}$ $V_{rt} = 24.685$

Totaal Reactor Volume [m³] $V_{Rt} := \frac{V_{rt}}{\varepsilon_{bed}}$ $V_{Rt} = 72.604$

Corrected volume for gas formation [m³] $V_R := V_{Rt} + 0.1\cdot V_{Rt}$ $V_R = 79.865$

Reactor height [m] $H_r := \frac{V_R}{A_r}$ $H_r = 6.355$

Retention Time [min] $T_r := 60\cdot\frac{V_{Rt}}{F_v}$ $T_r = 58.552$

Pressure Drop:

Superficial velocity [m/s] $v_s := \frac{F_v}{A_r \cdot 3600}$ $v_s = 1.645 \cdot 10^{-3}$

Superficial mass velocity [kg/m²s] $v_{ms} := v_s \cdot \rho$ $v_{ms} = 1.713$

Bed Porosity [-] $\varepsilon_{bed} := 0.34$

Viscosity [Ns/m²] $\mu := 0.485$

Ergun

$$\Delta P := \frac{-v_{ms} \cdot (1 - \varepsilon_{bed})}{\rho \cdot d_p \cdot \varepsilon_{bed}^3} \cdot \left[\frac{150 \cdot (1 - \varepsilon_{bed}) \cdot \mu}{d_p} + 1.75 \cdot v_{ms} \right] \cdot H_r$$

$\Delta P = -8.433 \cdot 10^4$ Pa $P_0 := 25$ Bar

$P := \frac{\Delta P}{101300} + P_0$ $P = 24.168$ Bar

Chimney tray calculation

Chimney area [m²] $A := 0.1 \cdot A_r$ $A = 1.257$

Chimney diameter (100 tubes) [m] $D := \sqrt{\frac{\frac{A}{100}}{0.25 \cdot \pi}}$ $D = 0.126$

Criterium [ρv²] $84 \cdot \left(\frac{F_{v.out}}{A \cdot 3600} \right)^2 = 3.489$

Free area [m²] $A_f := A_r - A$ $A_f = 11.31$

volume needed for a residence time of liquid of 3 min: $V_c := \frac{F_v}{60} \cdot 3$ $V_c = 3.72$

Chimney hoogte [m] $H := \frac{V_c}{A_f}$ $H = 0.329$

Total volume: $V := V_R + V_c$ $V = 83.585$

Total Height $H_t := H_r + H$ $H_t = 6.684$

Appendix 13 - MEG Reactor2 Design

Reactor Specifications

Reactor Diameter [m] $D_r := 4$

Cross sectional Area [m2] $A_r := \frac{1}{4} \cdot \pi \cdot D_r^2$ $A_r = 12.566$

Bed Porosity [-] $\varepsilon_{bed} := 0.34$

Catalyst Specifications

Particle diameter [m] $d_p := 0.01$

Reactivity [kg EC/m3 cat./h] $R_{cat} := 750$

Flow Specifications

EC massflow [kg/h] IN $M_{i.EC} := 18513.666$

OUT $M_{o.EC} := 18.514$

PRODUCTION $M_{p.EC} := M_{o.EC} - M_{i.EC}$

$M_{p.EC} = -1.85 \cdot 10^4$

Total massflow [kg/h] $F_{mt} := 67930.9$

$F_v := 66.7$ $F_{v.out} := 867.7$

Total Volume flow [m3/h]

Density [kg/m3] $\rho := \frac{F_{mt}}{F_v}$ $\rho = 1.018 \cdot 10^3$

Calculations:

Total Reaction Volume [m3] $V_{rt} := \frac{-M_{p.EC}}{R_{cat}}$ $V_{rt} = 24.66$

Totaal Reactor Volume [m3] $V_{Rt} := \frac{V_{rt}}{\varepsilon_{bed}}$ $V_{Rt} = 72.53$

Corrected volume for gas formation [m3] $V_R := V_{Rt} + 0.1 \cdot V_{Rt}$ $V_R = 79.783$

Reactor height [m] $H_r := \frac{V_R}{A_r}$ $H_r = 6.349$

Retention Time [min] $T_r := 60 \cdot \frac{V_{Rt}}{F_v}$ $T_r = 65.244$

Pressure Drop:

Superficial velocity [m/s] $v_s := \frac{0.5\cdot F_v}{A_r\cdot 3600}$ $v_s = 7.372\cdot 10^{-4}$

Superficial mass velocity [kg/m2s] $v_{ms} := v_s\cdot \rho$ $v_{ms} = 0.751$

Bed Porosity [-] $\varepsilon_{bed} := 0.34$

Viscosity [Ns/m2] $\mu := 0.485$

Ergun

$$\Delta P := \frac{-v_{ms}\cdot\left(1-\varepsilon_{bed}\right)}{\rho\cdot d_p\cdot \varepsilon_{bed}^{\,3}}\cdot\left[\frac{150\cdot\left(1-\varepsilon_{bed}\right)\cdot\mu}{d_p}+1.75\cdot v_{ms}\right]\cdot H_r$$

$\Delta P = -3.775\cdot 10^{4}$ Pa $P_0 := 25$ Bar

$P := \frac{\Delta P}{101300} + P_0$ $P = 24.627$ Bar

Chimney tray calculation

Chimney area [m2] $A := 0.1\cdot A_r$ $A = 1.257$

Chimney diameter (100 tubes) [m] $D := \sqrt{\frac{\frac{A}{100}}{0.25\cdot\pi}}$ $D = 0.126$

Criterium [ρv2] $84\cdot\left(\frac{F_{v.out}}{A\cdot 3600}\right)^2 = 3.09$

Free area [m2] $A_f := A_r - A$ $A_f = 11.31$

volume needed for a residence time of liquid of 3 min: $V_c := \frac{F_v}{60}\cdot 3$ $V_c = 3.335$

Chimney hoogte [m] $H := \frac{V_c}{A_f}$ $H = 0.295$

Total volume: $V := V_R + V_c$ $V = 83.118$

Total height: $H_t := H_r + H$ $H_t = 6.644$

Gas liquid vessel design

V101 steam drum

Hold up time for liquids [min]		$t_h := 5$
incoming steam from reactor [kg/h]		$F_{sr} := 106557$
outgoing BFW water [kg/h]		$F_{bfw} := 106557$
incoming water from equipment [kg/h]		$F_{ai} := 106557$
outgoing steam to equipment [kg/h]		$F_{ao} := 106557$
Specific volume saturated liquid [m³/kg]		$V_{sl} := 1.209 \cdot 10^{-3}$
Specific volume saturated vapor [m³/kg]		$V_{sv} := 71.45 \cdot 10^{-3}$
Densities [kg/m³]	$\rho_L := \frac{1}{V_{sl}}$	$\rho_L = 827.13$
	$\rho_v := \frac{1}{V_{sv}}$	$\rho_v = 13.996$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\frac{\rho_L}{\rho_v}}$	$u_v = 0.269$
Vapor volume flow [m³/s]	$F_v := \frac{F_{sr}}{\rho_v \cdot 3600}$	$F_v = 2.115$
Minimal vessel area [m²]	$A := \frac{F_v}{u_v}$	$A = 7.86$
Vessel area (chosen) [m²]		$A_v := A$
Diameter [m]	$D_v := \sqrt{4 \cdot \frac{A_v}{\pi}}$	$D_v = 3.164$
Disengagement space [m]	$H_d := D_v$	$H_d = 3.164$

Appendix 13

Liquid volume flow [m³/s]	$F_L := \frac{F_{bfw}}{\rho_L \cdot 3600}$	$F_L = 0.036$
Volume for holdup [m³]	$V_h := F_L \cdot t_h \cdot 60$	$V_h = 10.736$
Liquid depth [m]	$H_L := \frac{V_h}{A_v}$	$H_L = 1.366$
Vessel size:	$\text{Height} := H_L + H_d$	$\text{Height} = 4.529$ m
	$\text{Diameter} := D_v$	$\text{Diameter} = 3.164$ m
	$\text{Area} := A_v$	$\text{Area} = 7.86$ m^2
	$\text{Volume} := \text{Height} \cdot \text{Area}$	$\text{Volume} = 35.601$ m^3

V102 steam drum

Hold up time for liquids [min]		$t_h := 5$
incoming steam from reactor [kg/h]		$F_{sr} := 51117$
outgoing BFW water [kg/h]		$F_{bfw} := 51117$
incoming water from equipment [kg/h]		$F_{ai} := 51117$
outgoing steam to equipment [kg/h]		$F_{ao} := 51117$
Specific volume saturated liquid [m³/kg]		$V_{sl} := 1.102 \cdot 10^{-3}$
Specific volume saturated vapor [m³/kg]		$V_{sv} := 306.8 \cdot 10^{-3}$
Densities [kg/m³]	$\rho_L := \dfrac{1}{V_{sl}}$	$\rho_L = 907.441$
	$\rho_v := \dfrac{1}{V_{sv}}$	$\rho_v = 3.259$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\dfrac{\rho_L}{\rho_v}}$	$u_v = 0.584$
Vapor volume flow [m³/s]	$F_v := \dfrac{F_{sr}}{\rho_v \cdot 3600}$	$F_v = 4.356$
Minimal vessel area [m²]	$A := \dfrac{F_v}{u_v}$	$A = 7.46$
Vessel area (chosen) [m²]		$A_v := A$
Diameter [m]	$D_v := \sqrt{4 \cdot \dfrac{A_v}{\pi}}$	$D_v = 3.082$
Disengagement space [m]	$H_d := D_v$	$H_d = 3.082$

Liquid volume flow [m³/s] $F_L := \frac{F_{bfw}}{\rho_L \cdot 3600}$ $F_L = 0.016$

Volume for holdup [m³] $V_h := F_L \cdot t_h \cdot 60$ $V_h = 4.694$

Liquid depth [m] $H_L := \frac{V_h}{A_v}$ $H_L = 0.629$

Vessel size: $\text{Height} := H_L + H_d$ $\text{Height} = 3.711$ m

$\text{Diameter} := D_v$ $\text{Diameter} = 3.082$ m

$\text{Area} := A_v$ $\text{Area} = 7.46$ m^2

$\text{Volume} := \text{Height} \cdot \text{Area}$ $\text{Volume} = 27.684\ m^3$

V301 V/L separator C301

Hold up time for liquids [min]		$t_h := 5$
outgoing vapor flow [kg/h]		$F_{ov} := 1152$
outgoing liquid flow [kg/h]		$F_{ol} := 3193$
Densities [kg/m³] liquid flow		$\rho_L := 978$
vapor flow		$\rho_v := 0.1156$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\dfrac{\rho_L}{\rho_v}}$	$u_v = 3.219$
Vapor volume flow [m³/s]	$F_v := \dfrac{F_{ov}}{\rho_v \cdot 3600}$	$F_v = 2.768$
Minimal vessel area [m²]	$A := \dfrac{F_v}{u_v}$	$A = 0.86$
Vessel area (chosen) [m²]		$A_v := A$
Diameter [m]	$D_v := \sqrt{4 \cdot \dfrac{A_v}{\pi}}$	$D_v = 1.046$
Disengagement space [m]	$H_d := D_v$	$H_d = 1.046$
Liquid volume flow [m³/s]	$F_L := \dfrac{F_{ol}}{\rho_L \cdot 3600}$	$F_L = 9.069 \cdot 10^{-4}$
Volume for holdup [m³]	$V_h := F_L \cdot t_h \cdot 60$	$V_h = 0.272$
Liquid depth [m]	$H_L := \dfrac{V_h}{A_v}$	$H_L = 0.316$
Vessel size:	$\text{Height} := H_L + H_d$	$\text{Height} = 1.363$ m
	$\text{Diameter} := D_v$	$\text{Diameter} = 1.046$ m
	$\text{Area} := A_v$	$\text{Area} = 0.86$ m^2
	$\text{Volume} := \text{Height} \cdot \text{Area}$	$\text{Volume} = 1.172$ m^3

V302 V/L separator C302

Hold up time for liquids [min]		$t_h := 5$
outgoing vapor flow [kg/h]		$F_{ov} := 100$
outgoing liquid flow [kg/h]		$F_{ol} := 25666$
Densities [kg/m³] liquid flow		$\rho_L := 1013$
vapor flow		$\rho_v := 0.1$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\dfrac{\rho_L}{\rho_v}}$	$u_v = 3.523$
Vapor volume flow [m³/s]	$F_v := \dfrac{F_{ov}}{\rho_v \cdot 3600}$	$F_v = 0.278$
Minimal vessel area [m²]	$A := \dfrac{F_v}{u_v}$	$A = 0.079$
Vessel area (chosen) [m²]		$A_v := 4.$
Diameter [m]	$D_v := \sqrt{4 \cdot \dfrac{A_v}{\pi}}$	$D_v = 2.257$
Disengagement space [m]	$H_d := D_v$	$H_d = 2.257$
Liquid volume flow [m³/s]	$F_L := \dfrac{F_{ol}}{\rho_L \cdot 3600}$	$F_L = 7.038 \cdot 10^{-3}$
Volume for holdup [m³]	$V_h := F_L \cdot t_h \cdot 60$	$V_h = 2.111$
Liquid depth [m]	$H_L := \dfrac{V_h}{A_v}$	$H_L = 0.528$
Vessel size:	$\text{Height} := H_L + H_d$	$\text{Height} = 2.785$ m
	$\text{Diameter} := D_v$	$\text{Diameter} = 2.257$ m
	$\text{Area} := A_v$	$\text{Area} = 4$ m^2
	$\text{Volume} := \text{Height} \cdot \text{Area}$	$\text{Volume} = 11.138$ m^3

V401 V/L separator C404

Hold up time for liquids [min] $t_h := 5$

outgoing liquid flow [kg/h] $F_{ol} := 12976$

Densities [kg/m³] liquid flow $\rho_L := 1013$

Vessel area (chosen) [m²] $A_v := 1.5$

Diameter [m] $D_v := \sqrt{4 \cdot \frac{A_v}{\pi}}$ $D_v = 1.382$

Liquid volume flow [m³/s] $F_L := \frac{F_{ol}}{\rho_L \cdot 3600}$ $F_L = 3.558 \cdot 10^{-3}$

Volume for holdup [m³] $V_h := F_L \cdot t_h \cdot 60$ $V_h = 1.067$

Liquid depth [m] $H_L := \frac{V_h}{A_v}$ $H_L = 0.712$

Vessel size:

$\text{Height} := H_L$ $\text{Height} = 0.712$ m

$\text{Diameter} := D_v$ $\text{Diameter} = 1.382$ m

$\text{Area} := A_v$ $\text{Area} = 1.5$ m^2

$\text{Volume} := \text{Height} \cdot \text{Area}$ $\text{Volume} = 1.067$ m^3

Appendix 13

S201 CO_2 flash drum

Hold up time for liquids [min]		$t_h := 5$
outgoing vapor flow [kg/h]		$F_{ov} := 1916$
outgoing liquid flow [kg/h]		$F_{ol} := 693292$
Densities [kg/m^3] liquid flow		$\rho_L := 871$
vapor flow		$\rho_v := 10$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\frac{\rho_L}{\rho_v}}$	$u_v = 0.327$
Vapor volume flow [m^3/s]	$F_v := \frac{F_{ov}}{\rho_v \cdot 3600}$	$F_v = 0.053$
Minimal vessel area [m^2]	$A := \frac{F_v}{u_v}$	$A = 0.163$
Vessel area (chosen) [m^2]		$A_v := 15$
Diameter [m]	$D_v := \sqrt{4 \cdot \frac{A_v}{\pi}}$	$D_v = 4.37$
Disengagement space [m]	$H_d := D_v$	$H_d = 4.37$
Liquid volume flow [m^3/s]	$F_L := \frac{F_{ol}}{\rho_L \cdot 3600}$	$F_L = 0.221$
Volume for holdup [m^3]	$V_h := F_L \cdot t_h \cdot 60$	$V_h = 66.331$
Liquid depth [m]	$H_L := \frac{V_h}{A_v}$	$H_L = 4.422$
Vessel size:	$\text{Height} := H_L + H_d$	$\text{Height} = 8.792$ m
	$\text{Diameter} := D_v$	$\text{Diameter} = 4.37$ m
	$\text{Area} := A_v$	$\text{Area} = 15$ m^2
	$\text{Volume} := \text{Height} \cdot \text{Area}$	$\text{Volume} = 131.884$ m^3

S202 Flash Drum

Hold up time for liquids [min]		$t_h := 5$
outgoing vapor flow [kg/h]		$F_{ov} := 9541$
outgoing liquid flow [kg/h]		$F_{ol} := 271$
Densities [kg/m³] liquid flow		$\rho_L := 750$
vapor flow		$\rho_v := 11$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\frac{\rho_L}{\rho_v}}$	$u_v = 0.289$
Vapor volume flow [m³/s]	$F_v := \frac{F_{ov}}{\rho_v \cdot 3600}$	$F_v = 0.241$
Minimal vessel area [m²]	$A := \frac{F_v}{u_v}$	$A = 0.834$
Vessel area (chosen) [m²]		$A_v := A$
Diameter [m]	$D_v := \sqrt{4 \cdot \frac{A_v}{\pi}}$	$D_v = 1.03$
Disengagement space [m]	$H_d := D_v$	$H_d = 1.03$
Liquid volume flow [m³/s]	$F_L := \frac{F_{ol}}{\rho_L \cdot 3600}$	$F_L = 1.004 \cdot 10^{-4}$
Volume for holdup [m³]	$V_h := F_L \cdot t_h \cdot 60$	$V_h = 0.03$
Liquid depth [m]	$H_L := \frac{V_h}{A_v}$	$H_L = 0.036$
Vessel size:	$\mathrm{Height} := H_L + H_d$	$\mathrm{Height} = 1.066$ m
	$\mathrm{Diameter} := D_v$	$\mathrm{Diameter} = 1.03$ m
	$\mathrm{Area} := A_v$	$\mathrm{Area} = 0.834$ m^2
	$\mathrm{Volume} := \mathrm{Height} \cdot \mathrm{Area}$	$\mathrm{Volume} = 0.889$ m^3

S301 CO_2 Flash Drum

Hold up time for liquids [min]		$t_h := 5$
outgoing vapor flow [kg/h]		$F_{ov} := 9732$
outgoing liquid flow [kg/h]		$F_{ol} := 45$
Densities [kg/m³] liquid flow		$\rho_L := 823$
vapor flow		$\rho_v := 12$
maximum design vapor velocity from C&R p410 [m/s]	$u_v := 0.035\sqrt{\frac{\rho_L}{\rho_v}}$	$u_v = 0.29$
Vapor volume flow [m³/s]	$F_v := \frac{F_{ov}}{\rho_v \cdot 3600}$	$F_v = 0.225$
Minimal vessel area [m²]	$A := \frac{F_v}{u_v}$	$A = 0.777$
Vessel area (chosen) [m²]		$A_v := A$
Diameter [m]	$D_v := \sqrt{4 \cdot \frac{A_v}{\pi}}$	$D_v = 0.995$
Disengagement space [m]	$H_d := D_v$	$H_d = 0.995$
Liquid volume flow [m³/s]	$F_L := \frac{F_{ol}}{\rho_L \cdot 3600}$	$F_L = 1.519 \cdot 10^{-5}$
Volume for holdup [m³]	$V_h := F_L \cdot t_h \cdot 60$	$V_h = 4.557 \cdot 10^{-3}$
Liquid depth [m]	$H_L := \frac{V_h}{A_v}$	$H_L = 5.863 \cdot 10^{-3}$
Vessel size:	$\mathrm{Height} := H_L + H_d$	$\mathrm{Height} = 1.001$ m
	$\mathrm{Diameter} := D_v$	$\mathrm{Diameter} = 0.995$ m
	$\mathrm{Area} := A_v$	$\mathrm{Area} = 0.777$ m^2
	$\mathrm{Volume} := \mathrm{Height} \cdot \mathrm{Area}$	$\mathrm{Volume} = 0.778$ m^3

Appendix 13 - Heat exchanger calculations

Name				Heat dut	Inlet T	Outlet T		IN			OUT			mass flow		OPEX
																$/hr
			kW	GJ/hr	°C	°C		°C	bara		°C	bara		kg/hr		
R101 Reactor	**R101**	EO reactor	-53625	-193.05		240	hot	230	27.6	liq	230	27.6	vap	-106557	generated HPS	$ -988.85
E101	**E101**	Exchanger	31901	114.84	180	110	cold	34			110				E101	
S201: Flash drum	**S201**	Heater	-13694	-49.30		100	hot	33		liq	43		liq	1179426	Cooling water	$ 58.97
S202: Flash drum	**S302**	Heater	-222	-0.80		100	hot	33		liq	43		liq	19139	Cooling water	$ 0.96
S301: Flash drum	**S301**	Heater	-150	-0.54		125	hot	33		liq	43		liq	12919	Cooling water	$ 0.65
C301: Distillation column	**E302**	Condenser	-4413	-15.89		14	hot	4		liq	10		liq	630687	Chilled water	$ 132.44
C404: Heavy ends separator	**E407**	Condenser	-2300	-8.28		29	hot	4		liq	10		liq	328702	Chilled water	$ 69.03
C302: Distillation column	**E304**	Condenser	-8944	-32.20		130	hot	120	1.9	liq	120	1.9	vap	-14621	generated LPS	$ -73.11
C301: Distillation column	**E301**	Reboiler	3500	12.60		132	cold	160	7.8	vap	160	7.8	liq	6054	MPS	$ 48.43
C302: Distillation column	**E303**	Reboiler	8889	32.00		164	cold	230	27.6	vap	230	27.6	liq	17663	HPS	$ 163.91
C401: Raw EO stripper	**E401**	Reboiler	29806	107.30		136	cold	230	27.6	vap	230	27.6	liq	59226	HPS	$ 549.62
C403: EO secondary Stripper	**E405**	Reboiler	5000	18.00		109	cold	120	1.9	vap	120	1.9	liq	8173	LPS	$ 40.87
C404: Heavy ends separator	**E406**	Reboiler	2031	7.31		121	cold	160	7.8	vap	160	7.8	liq	3512	MPS	$ 28.10
C405: Light End separator	**E410**	Reboiler	1583	5.70		61	cold	120	1.9	vap	120	1.9	liq	2588	LPS	$ 12.94
E102	**E102**	Exchanger	51290	184.64	110	220	cold	230	27.6	vap	230	27.6	liq	101918	HPS	$ 945.80
E103	**E103**	Exchanger	-29551	-106.38	240	180	hot	160	7.8	liq	160	7.8	vap	-51117	Generated MPS	$ -408.93
E104	**E104**	Exchanger	-29665	-106.79	110	43	hot	33		liq	43		liq	2554898	Cooling water	$ 127.74
E105a	**E105a**	Exchanger	-15466	-55.68	43	20	hot	4		liq	10		liq	2210340	Chilled water	$ 464.17
E105b	**E105b**		-23787	-85.63	78	43	hot	33		liq	43		liq	2048650	Cooling water	$ 102.43
E201 tubes	**E201 tubes**		-26108	-93.99	117	78	hot	105		liq	42		liq		E201	
		total	**-65362**	**-235.30**												
E201 shell	**E201 shell**	Exchanger	26108	93.99	42	105	cold	78		liq	117		liq		E105	
E202	**E202**	Exchanger	3473	12.50	89	150	cold	160	7.8	vap	160	7.8	liq	6008	MPS	$ 48.06
E203	**E203**	Exchanger	406	1.46	133	150	cold	160	7.8	vap	160	7.8	liq	703	MPS	$ 5.62
E408	**E408**	Exchanger	-744	-2.68	138	57	hot	33		liq	43		liq	64053	Cooling water	$ 3.20
E402	**E402**	Exchanger	-542	-1.95	134	43	hot	33		liq	43		liq	46711	Cooling water	$ 2.34
E403	**E403**	Exchanger	-4636	-16.69	131	43	hot	33		liq	43		liq	399253	Cooling water	$ 19.96
E404	**E404**	Exchanger	-1359	-4.89	100	43	hot	33		liq	43		liq	117079	Cooling water	$ 5.85
E409	**E409**	Exchanger	-4135	-14.89	108	43	hot	33		liq	43		liq	356150	Cooling water	$ 17.81

Appendix 13 - Heat exchanger calculations

Name	720 K									
					Tube lengt	Tube dia	One tube	Tubes	Shell dia	outlet pipe
	LMTD		U	A	m	m	m2		m	m
R101	20			6929	8	0.044	1.11	6265	4.66	1.55
E101	73		110	3973						
			4 maal	993	4.88	0.025	0.38	2589	2.13	1.00
S201	Coils are not sized, heat transfer is presumed not to be a problem									
S302	Coils are not sized, heat transfer is presumed not to be a problem									
S301	Coils are not sized, heat transfer is presumed not to be a problem									
	LMTD		U	A	Tube lengt	Tube dia	One tube	Tubes	Shell dia	outlet pipe
	K		W/m2 K	m2	m	m	m2		m	m
E302	7		2000	337	2.44	0.025	0.19	1756	1.79	0.82
E407	22		2000	52	2.44	0.025	0.19	272	0.78	0.32
E304	10		1000	894	2.44	0.025	0.19	4662	2.78	1.34
	MTD	Tr	heat flux	A	Tube lengt	Tube dia	one tube	Tubes	Shell dia	outlet pipe
	K	-	W/m2	m2	m	m	m2		m	m
E301	27.70	0.41	35000	100.00	2.44	0.025	0.19	521	0.87	0.45
E303	65.60	0.55	60000	148.15	2.44	0.025	0.19	772	1.03	0.55
E401	94.20	0.51	60000	496.76	2.44	0.025	0.19	2589	1.78	1.00
E405	11.00	0.32	35000	142.86	2.44	0.025	0.19	745	1.02	0.54
E406	39.40	0.39	40000	50.76	2.44	0.025	0.19	265	0.64	0.32
E410	58.90	0.25	60000	26.39	2.44	0.025	0.19	138	0.48	0.23
	LMTD		U	A	Tube lengt	Tube dia	One tube	Tubes	Shell dia	outlet pipe
	K		W/m2 K	m2	m	m	m2		m	m
E102	44		300	3856						
			2 maal	1928	2.22	0.044	0.31	6265	4.66	1.55
E103	43		300	2276						
			2 maal	1138	1.31	0.044	0.18	6265	4.66	1.55
E104	30		300	3300						
			4 maal	825	4.88	0.025	0.38	2150	1.64	0.91
E105a	36		900	477	2.44	0.025	0.19	2484	1.75	0.98
E105b	32		900	825	2.44	0.025	0.19	4301	2.23	1.29
E201 Tubes	22		400	2950						
			3 maal	983	4.88	0.025	0.38	2562	1.77	0.99
E201 Shell	22		400	2949						
			3 maal	983	4.88	0.025	0.38	2562	1.77	0.99
E202	31		1000	111	2.44	0.025	0.19	581	0.91	0.47
E203	17		1000	24	2.44	0.025	0.19	123	0.46	0.22
E408	51		300	48	2.44	0.025	0.19	251	0.63	0.31
E402	37		300	49	2.44	0.025	0.19	257	0.63	0.32
E403	36		300	431	2.44	0.025	0.19	2249	1.67	0.93
E404	27		750	67	2.44	0.025	0.19	350	0.73	0.37
E409	29		750	188	2.44	0.025	0.19	979	1.15	0.61

Appendix 14 - Fire and Explosive Index

Process Unit	EO reactor, including recycle compressor		EO absorber	
Material	ethylene (presence of EO also considered)		EC/MEG	
Material factor	24		5	
General process hazards F_1				
	0.5	moderate exotherm	0.5	diking
	1	critical-to-control exotherm		
F_1 =	1.5		0.5	
Special process hazards F_2	4	health ranking	1	health ranking
	0.8	penalty	0.2	penalty
	0.8	flammable range penalty	1	pressure penalty
	1	pressure penalty	10.45	X
	0.12	X	2.39	quantity penalty
	0.20	quantity penalty	0.10	corrosion penalty
	0.20	corrosion penalty		
	0.50	compressor penalty		
F_2 =	3.50		3.69	
F_3=F_1*F_2	5.25		1.84	
F&EI	126		9	

Process Unit	EO purification C403		MEG purification (C301+C302	
Material	EO		MEG	
Material factor	29		5	
General process hazards F1				
	0.3	liquid above flashpoint	0.5	diking
	0.5	diking		
F1 =	0.8		0.5	
Special process hazards F2	4	health ranking	1	health ranking
	0.8	penalty	0.2	penalty
	0.5	subatmospheric penalty	0.5	subatmospheric pe
	0.78	X	3.62	X
	1.32081	quantity penalty	2.107062	quantity penalty
	0.2	corrosion penalty	0.1	corrosion penalty
	0.8	flammable range penalty	0.1	leakage likely
			0.5	vacuum compress
F2 =	3.62081		3.507062	
F3=F1*F2	2.896648		1.753531	
F&EI	84.0028		8.767654	

Appendix 15 – Economy

The method used for the cost-estimation of heat-exchangers, compressors and pumps is to consider the size of the equipment and to relate that to a cost using the formula below.

$C_e = C * S^n$

wherein:

C_e = purchased equipment cost
S = characteristic size parameter
C = cost constant
n = index for that type of equipment.

These calculations are done on a spreadsheet and attached. The C value , the S parameter and the n index are all given for each piece of equipment.

Exchangers

Data is gathered from conceptual cost estimation manual. Prices are given in 1984 US dollars. Prices are all based on carbon steel (CS) shell and tube, and no tubesheet. When stainless steel (SS) is used, the cost is multiplied by 1.2 for SS tubes, 1.2 for SS shells, and by 1.1 for a CS or 1.3 for a SS tubesheet.
These values are time corrected by multiplying them with 1.04^{15} to account for the 15 year time difference.

Reactors

Data is gathered from conceptual cost estimation manual. Prices are given in 1984 US dollars. EO reactor, is calculated as if this is a huge heat exchanger. The costs of the MEG and EC reactors are gathered from price information per reactor volume..
These values are time corrected by multiplying them with 1.04^{15} to account for the 15 year time difference.

Columns

Data is gathered from Coulson and Richardsons fig 6.4. The prices are given in 1992 pound sterling. These are converted to US dollars by multiplying with 1.6.
These values are time then corrected by multiplying them with 1.04^{7} to account for the 15 year time difference.

Pumps, Compressors and vessels

Data is gathered from Coulson and Richardsons table 6.2. The price is given in 1992 pound sterling. These are converted to US dollars by multiplying with 1.6.

These values are then time corrected by multiplying them with 1.04^7 to account for the 15 year time difference.

CO_2 removal unit

The price of the CO2 removal unit is based on the price of the CO2 removal unit in the Scientific design license, correcting for the different EO production. This leads to a price of 7 million dollars.

Profitability measures

For the calculation of the discounted cash flow the following formula has been used:

$$DCF = NCF / (1 + r)^n$$

DCF = discounted cash flow
NCF = net cash flow
r = discount rate in % /100
n = year for which the DCF is calculated.

Economy

Process Equipment

Columns: data from C&R fig 6.4. in pound sterling mid 1992

	H	D	steel	packing	trays	Costs		time base
C101 EO absorber	32	4.7	ss	mella	15	k£	462	1992
C301 Water removal	6	2.8	cs	sieve cs	6	k£	30	1992
C302 MEG purification	17	2.9	cs	mella	11	k£	105	1992
C401 Raw EO stripper	20	3.3	ss	mella	8	k£	140	1992
C402 Sec EO absorber	23	0.8	ss	mella	9	k£	53	1992
C403 Sec EO stripper	12	1.8	ss	mella	5	k£	45	1992
C404 Heavy end separator	10	0.7	ss	sieve	15	k£	24	1992
C405 Light end separator	3	0.9	ss	sieve	20	k£	13 +	1992
Total columns costs						k£	872	1992

Exchangers: data from conceptual cost estimation manual in dollars mid 1984

$ 180 per m2; SS tubes *1.2; tubesheet: CS*1.1 SS*1.3 C*1.2

	Area (m2)	Tubes	Shell	Sheet			
E101 Feed preheater	3973	SS	SS	SS	k$	1,116	1984
E102 Feed Heater	3856	SS	CS	CS	k$	916	1984
E103 Effluent cooler	2276	SS	CS	CS	k$	541	1984
E104 Effluent secondary cooler	3300	SS	CS	CS	k$	784	1984
E105A Absorbent cooler	477	SS	CS	CS	k$	113	1984
E105B Absorbent cooler	825	SS	CS	CS	k$	196	1984
E201 Preheater R201	2950	SS	SS	SS	k$	828	1984
E202 Preheater R202	111	CS	CS	CS	k$	22	1984
E203 Preheater R203	24	CS	CS	CS	k$	5	1984
E301 Reboiler C301	100	CS	CS	CS	k$	20	1984
E302 Condensor C301	253	CS	CS	CS	k$	50	1984
E303 Reboiler C302	148	CS	CS	CS	k$	29	1984
E304 Condenser C302	256	CS	CS	CS	k$	51	1984
E401 Reboiler C401	496	SS	CS	CS	k$	118	1984
E402 Post - cooler K401	49	SS	CS	CS	k$	12	1984
E403 Post Cooler K402/403	431	SS	CS	CS	k$	102	1984
E404 Absorbent cooler	67	SS	CS	CS	k$	16	1984
E405 Reboiler C403	143	SS	CS	CS	k$	34	1984
E406 Reboiler C404	51	SS	CS	CS	k$	12	1984
E407 Condenser C404	52	SS	CS	CS	k$	12	1984
E408 Post cooler K404	48	SS	CS	CS	k$	11	1984
E409 Reboiler C405	26	SS	CS	CS	k$	6 +	1984
Total heat exchanger costs					k$	4,995	1984

compressors Coulson and Richardsons tab

cost = $C * S^n$	power S (kW)	C (£)	n	spared			
K101 Loop gas compressor	3058.6	500	0.8	1	k£	307	1992
K102 Recycle compressor	745.4	500	0.8	1	k£	99	1992
K103 CO_2 compressor	286.9	500	0.8	1	k£	46	1992
K301 vacuum compressor C301	680.0	700	0.8	1	k£	129	1992
K302 vacuum compressor C302	7.0	700	0.8	1	k£	3	1992
K401 EO primary compressor	448.5	500	0.8	1	k£	66	1992
K402 EO sec Compressor	644.0	500	0.8	1	k£	88	1992
K403 EO recycle Compressor	205.6	500	0.8	1	k£	35	1992
K404 Raw pri. EO compressor	1067.6	500	0.8	1	k£	132	1992
K405 Raw sec. EO compressor	451.1	500	0.8	1	k£	66 +	1992
Total compressor costs					k£	974	1992

pumps Coulson and Richardsons table 6.2

cost = $C * S^n$		power S (kW)	API-610 C (£)	n	spared			
P101	Main absorbent pump	465.1	500	0.8	2	£	136,154	1992
P102	Absorbent pump	15.9	500	0.8	1	£	4,579	1992
P103	Coolant pump	6.3	500	0.8	2	£	4,339	1992
P104	Coolant pump	3.0	500	0.8	2	£	2,410	1992
P201	EO feed pump	287.4	500	0.8	1	£	46,322	1992
P202	liquid phase pump S202	0.2	500	0.8	1	£	119	1992
P203	Water pump	2.1	500	0.8	1	£	901	1992
P204	liquid phase pump S201	46.8	500	0.8	1	£	10,835	1992
P205	liquid phase pump R202	6.4	500	0.8	1	£	2,221	1992
P301	Reflux pump C301	2.2	500	0.8	1	£	941	1992
P302	Reflux pump C302	0.7	500	0.8	1	£	363	1992
P303	MEG pump	2.4	500	0.8	1	£	1,000	1992
P304	Heavy component pump	0.0	500	0.8	1	£	14	1992
P305	Raw MEG pump	3.5	500	0.8	1	£	1,348	1992
P401	Absorbent pump	257.0	500	0.8	1	£	42,357	1992
P402	Reflux pump C404	0.1	500	0.8	1	£	94	1992
P403	Bottom pump C402	9.1	500	0.8	1	£	2,930	1992
P404	Absorbent pump	23.1	500	0.8	1	£	6,173	1992
P405	Bottom pump C405	1.9	500	0.8	1	£	845 +	1992
Total pump costs						k£	264	1992

Appendix 15 - Economy

Reactors from Conceptual cost estimating manual

R101 EO reactor	based on 2 heat exchangers with A=7000	k$	3,326		1984
R201 Carbonate reactor	data from Conceptual cost estimating manual	k$	600		1984
R202 Primary MEG reactor	data from Conceptual cost estimating manual	k$	500		1984
R203 Secondary MEG Reactor	data from Conceptual cost estimating manual	k$	500	+	1984
Total reactor costs		**k$**	**4,926**		1984

vessels from Coulson and Richardsons table 6.2

		volume m3				
S201 CO2 flash drum	horizontal CS	132	k£	28		1992
S202 Flash drum	horizontal CS	1	k£	2		1992
S301 CO2 flash drum	horizontal CS	1	k£	1		1992
V101 Steam drum	horizontal CS	35.6	k£	13		1992
V102 Steam drum	horizontal CS	27.7	k£	11		1992
V301 V/L separator C301	horizontal CS	1	k£	2		1992
V302 V/L separator C302	horizontal CS	11	k£	6		1992
V401 V/L Separator C404	horizontal SS	1	k£	2	+	1992
Total vessel costs			**k£**	**64**		1992

Special equipment				
C102 CO2 removal system	fixed price, source Scientific Design	k$	7,000	1999
cost index for 1 year	1.04			

Summary

Total columns costs	k£	872	1992
converted to dollars (£ = 1.6$)	k$	1,395	1992
time corrected	k$	1,836	1999
Total heat exchanger costs	k$	4,995	1984
time corrected	k$	8,995	1999
Total compressor costs	k£	974	1992
converted to dollars (£ = 1.5$)	k$	1,558	1992
time corrected	k$	2,050	1999
Total pump costs	k£	264	1992
converted to dollars (£ = 1.5$)	k$	422	1992
time corrected	k$	556	1999
Total reactor costs	k$	4,926	1984
time corrected	k$	8,872	1999
Total vessel costs	k£	64	1992
converted to dollars (£ = 1.5$)	k$	103	1992
time corrected	k$	136	1999
C102 CO2 removal system	k$	7,000	1999
Total purchased equipment costs (PCE)	**k$**	**29,310**	1999
Total physical plant costs (PPC)	k$	95,256	
Fixed capital (FC)	**k$**	**138,121**	

Appendix 15 - Economy

1 Fixed capital (FC)					**k$**	**138,121**
Operating Costs					Cost per Year	
variable costs	price		use			
Ethylene	520.00	$/ton	193,198	ton	k$	100,463
Oxygen	50.00	$/ton	184,584	ton	k$	9,229
Methane	120.00	$/ton	11,592	ton	k$	1,391
Methylene chloride	750.00	$/ton	12	ton	k$	9
Water	2.50	$/ton	65,143	ton	k$	163
Catalyst EO reaction, alumina carrier	10352.94	$/m3	125,975 m3 per 3 yr		k$	435
Silver for catalyst	5.18	$/ounce	176,056	=5 ton	k$	912
Catalyst based on DOW MSA-1	11764.71	$/m3	327	m3	k$	3,843
Chilled water	0.21	$/ton			k$	5,323
Cooling water	0.02	$/ton			k$	1,090
Steam	9.28	$/ton			k$	2,595
Electricity	0.04	$/kWh	69,823,689	kWh	k$	2,793
2 Total raw materials					**k$**	**116,445**
3 Total Utilities					**k$**	**11,801**
4 Miscellaneous materials	10% of 5				k$	1,381
5 Maintenance	10% of 1				k$	13,812
6 Operating labour					k$	2,700
7 Lab costs	20% of 6				k$	540
8 Supervision	20% of 6				k$	540
9 Plant overheads	50% of 6				k$	1,350
10 Capital charges	15% of 1				k$	20,718
11 Insurance	1% of 1				k$	1,381
12 Local taxes	2% of 1				k$	2,762
13 Royalties	1% of 1				k$	1,381 +
Annual production costs					**k$**	**174,812**
Annual income		Price	Production			
EO sales		$ 1,080.00	103,800		k$	112,104
MEG sales		$ 650.00	205,328		k$	133,463 +
Total sales					**k$**	**245,567**
Economic margin					**k$**	**70,755**

Appendix 16 - Lovenstein calculations

Column number	Tray spacing	Density liq	Density vap	Vapor rate	Vapor velocity	Column diameter		Aspen comparison
	lt	rhol	rhov	Vw	uv	Dc	height	Aspen Dc
	[m]	[kg/m3]	[kg/m3]	[kg/s]	[m/s]	[m]	[m]	[m]
C301	0.6096	1011	1	9.04	1.718	2.589	6.1	2.77
C404	0.6096	896	3	3.604	0.932	1.281	11.5824	1.22
C405	0.6096	809	8	3.211	0.541	0.972	60.96	0.91