CPD 3244

Basis of Design for

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

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1. DESCRIPTION OF THE DESIGN

This document presents the preliminary basis of design for a plant for the production of 200,000 mt/a fiber-grade MEG and 100,000 mt/a direct sales specification EO.

The design is done for two parties: Stork Engineers & Contractors, and the faculty of chemical engineering, Delft University of Technology. Stork would like to see a comparison between different designs currently commercially available, within the limits of licenses, proven technology, and commercial viability. While Stork is more interested in the comparison and analysis of design, the University is focused on the group employing their skills in order to design an installation which can operate from a practical point of view. While having two principals makes the task more interesting and challenging, the interests of both parties can occasionally conflict. We believe that a balance has been found, and a design been selected that should appeal to all parties.

The global capacity for ethylene oxide (EO, or oxirane) currently stands at about 14.5 Mmt/a, with 2.5 Mmt/a in western Europe, and 5.2 Mmt/a in North America. The balance is accounted for largely by the Middle East and Asia. Globally, the demand for oxirane is growing strongly, due to increased demand for polyesters. MEG accounts for some 60%[1] of the EO demand (split evenly between anti-freeze grade and fiber-grade[2]), although in Europe the trend is towards higher value EO derivatives. The total market for MEG has grown from 6.2 Mmt/a in 1990 to 10.1 Mmt/a in 1998. The growth in EO and MEG demand is expected to continue up to and past the year 2005.[1,3].

The aim of the design is to produce 100,000 mt/a purified EO and 200,000 mt/a fiber-grade MEG, at the highest economically feasible selectivity's. Details of the product streams are given in chapter 3, plant capacity. 70 % of the operational costs of existing plants consist of ethylene feed. Therefore, a higher selectivity markedly decreases these costs.

The production of MEG historically is accompanied by a substantial production of di-ethylene glycol (DEG), as well as higher glycols. It is the aim of this design to produce as little DEG as possible. A low energy usage and carbon dioxide production are also focal points.

Chapter 2 outlines the process concept chosen, and our motivation thereof. A simple block-scheme, as well as thermodynamic and pure component properties. Chapter 3, "Basic Assumptions", gives some of the details of the basis of design, such as capacity, location, battery limit, and streams.
Chapter 4 provides a number of commercial and financial reflections.

2. PROCESS DEFINITION

2.1 ETHYLENE OXIDE

Ethylene oxide has been commercially produced by two basic routes: the ethylene chlorohydrin process and direct oxidation. Other, commercially not viable, process include: arsenic-catalyzed liquid phase process, thallium-catalyzed epoxidation process, Lummus hypochlorite process, liquid-phase epoxidation with hydroperoxides, electrochemical process, unsteady-state direct oxidation process, fluid-bed direct oxidation process, and biological processes[4]. None of these processes has passed the pilot-plant phase.

The chlorohydrin process is based on the production of ethylene oxide from ethylene chlorohydrin by dehydrochlorination using either sodium of 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 environmental problems.

The direct oxidation of ethylene proceeds by the following reactions:

$$CH_2{=}CH_2 + \tfrac{1}{2}\,O_2 \rightarrow CH_2\text{-}CH_2 \quad (+\ 104.5\ \text{KJ/mol}) \qquad \textbf{[eqn 1]}$$

$$\backslash\ /$$

$$O$$

$$CH_2{=}CH_2 + 3\,O_2 \rightarrow 2\,CO_2 + 2\,H_2O \quad (+\ 1320\ \text{KJ/mol}) \qquad \textbf{[eqn 2]}$$

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[5]. 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 from air, or as a pure feed. Air based plants, eliminating an air-purification plant, are economically only feasible for a maximum production of ~25,000 t/a EO. Larger plants usually employ the oxygen based direct oxidation process.

The standard oxygen 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 ethylene glycol at the time of recovery. As much as 3-15% of all ethylene oxide can thus 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.
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[6].

Major licensors for the EO oxygen process are Scientific Design, Shell, Nippon Shokubai[7]. Over 70% of present world capacity is based on their processes. Union Carbide Corp. and

Dow Chemical use their own processes[7]. About 94% of U.S. EO capacity is located on the Gulf Coast, near secure and plentiful ethylene supplies.

Producer	Location	Capacity, 10^3 mt/a	Process oxidant	Technology
BASF[1]	Geismar, La.	218 → 425	oxygen	Shell
Dow	Plaquemine, La.	227	oxygen	Dow
Eastman	Longview, Tex.	91	oxygen	Shell
Formosa Plastic Corp.[2] source: 8	Point Comfort, Tex.	240 → 265	oxygen	unknown
Hoechst-Celanese	Clear Lake, Tex.	209	oxygen	Shell
Olin	Brandenburg, Ky.	50	oxygen	Shell
Oxy Petrochemicals	Bayport, Tex.	250	oxygen	Shell
PD Glycols	Beaumont, Tex.	202	oxygen	Scientific Design
Quantum	Morris, Ill.	113	oxygen	Scientific Design
Shell	Geismar, La.	364	oxygen	Shell
Sun Refining	Claymont, Del.	45	oxygen	Shell
Texaco	Port Neches, Tex.	332	oxygen	Scientific Design
Union Carbide	Seadrift, Tex.	349	air	Union Carbide
Union Carbide	Taft, La.	368	air, oxygen	Union Carbide
	Total	*3118*		

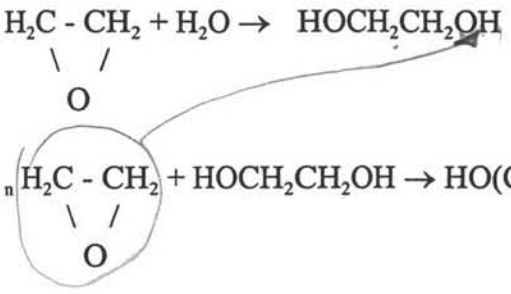
1. Being upgraded to ~425 mt/a, ready in 2000 (source:8)
2. Being upgraded to ~265 mt/a, ready in 2001 (source:8)

Table 2.1 U.S. Producers, capacities, process types, and technology for EO in 1992[7].

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[5].
Based on such commercial factors as number of plants already built/licensed, number of plants currently under construction/planned, catalyst quality (selectivity, activity, lifetime, costs), and reactor design, Shell/CRI, the absolute market leader, has been selected as the EO process for our basis of design. This choice also has a practical nature - many of the figures known to us apply to the Shell/CRI design[9]. Let it again be stressed that at the current time, based on the available literature, little variation between EO designs can be found.

2.2 MONO-ETHYLENE GLYCOL

Currently the method of ethylene glycol production is uncatalysed thermal hydrolysis of ethylene oxide. The ethylene oxide - water mixture is heated to ca. 200 °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[10]:



[eqn 3]

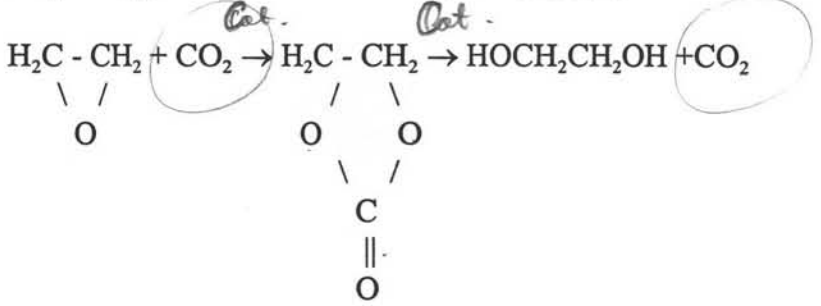
[eqn 4]

n = 1, 2, 3

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.
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 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 or replaced.

In the carbonate route process ethylene glycol is produced from ethylene oxide, with ethylene carbonate as an intermediate, the ethylene carbonate is subjected to hydrolysis to produce ethylene glycol. Both reaction steps are catalytic.



In table 1 a comparison of the different processes is presented. Five topics were used to compare the different processes. A more detailed version of table 1 can be found in appendix A

Process	Conversion[2]	Selectivity	H_2O/EO	number of main components	Implementation in industry
Conventional Thermal	=	-	-	+	++
Conventional catalytic[1]	=	+	+	++	-
Carbonate route process	=	++	++	-	-

notes:
1. catalyst dependent. Best found values are given
2. single pass conversion

Table 2.1. Comparison of different MEG processes

The two most important selection criteria are selectivity and the H_2O/EO ratio. The selectivity because ethylene efficiency (ethylene costs are a major part of the OPECS) is an important economic factor. The H_2O/EO ratio because of the large effect it has on the energy requirements of the separation section (separation of water and ethylene glycol).
The carbonate route process reaches a conversion and selectivity of nearly 100%, 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 two- to five-fold molar excess is added, as compared to a 20-fold excess for conventional processes. Therefore much less energy is needed in this process.
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[6]. 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, in a mini-plant[11].

2.3 PROCESS CONCEPT CHOSEN

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 will guarantee the workability of a plant. This makes the licensing of such design very attractive to potential customers.

The conceptual process design, however, is theoretically not hindered by the constraints of 'proven designs'. 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. 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 is illustrated in figure 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 1/3 for sales and 2/3 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 reabsorption in the MEG plant, albeit in a different absorbant. The absorption is therefore redundant. Several more elegant schemes are possible.

Scheme 2 involves splitting the EO reactor effluent into approximately 1/3 for EO production and 2/3 for MEG production. The EO production train is 'standard' Shell/CRI design, and the 2/3 MEG production is standard Mitsubishi design. However, this design contains 2 recycles for the EO reactor, since 2 absorbers are now operating parallel. This makes control of the reactor feed, critical because of explosive and flammability dangers, very difficult. A variation of this scheme involves a different absorbent better suited for the requirements than ethylene carbonate/ethylene glycol, propylene carbonate, in the Shell/CRI EO absorption train[12]. Absorption of EO in EC/EG is described in a patent to DOW Chemical Co[13].

The third scheme is the most preferable. In this design, the effluent of the Shell/CRI 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.

The heating and cooling of the large amounts of water as in the classical processes is not necessary in this design - much less absorbant, with a lower heat capacity (half that of water) is used. This design also eliminates a high-content ethylene oxide stream, which is very advantageous to process safety. Furthermore, equipment costs are lower, since only one absorber is required. This scheme takes full advantage of the experience of Shell/CRI in the EO design and Mitsubishi in the MEG design, while minimizing costs, enhancing operational safety, and maximizing reliability.

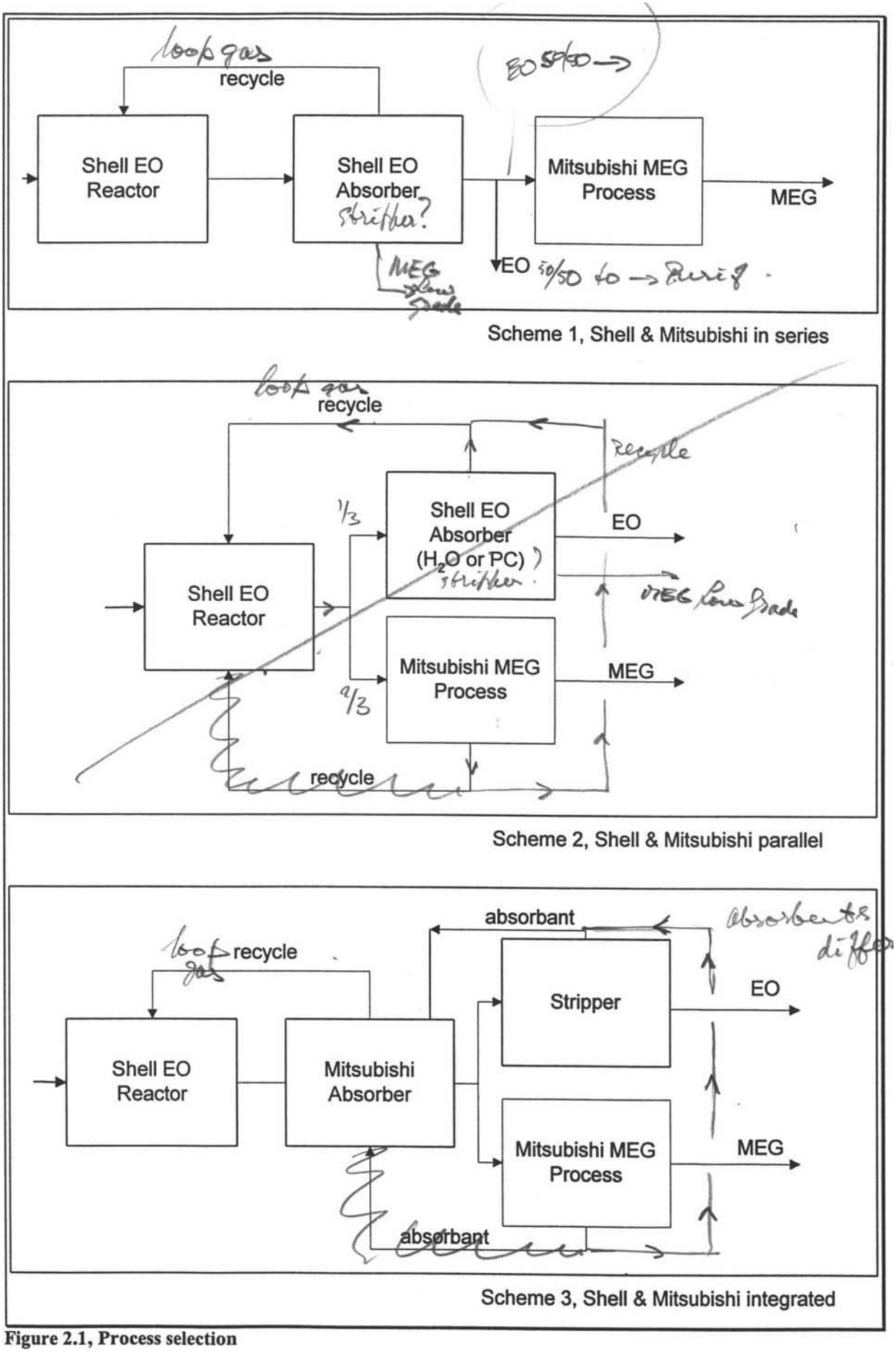
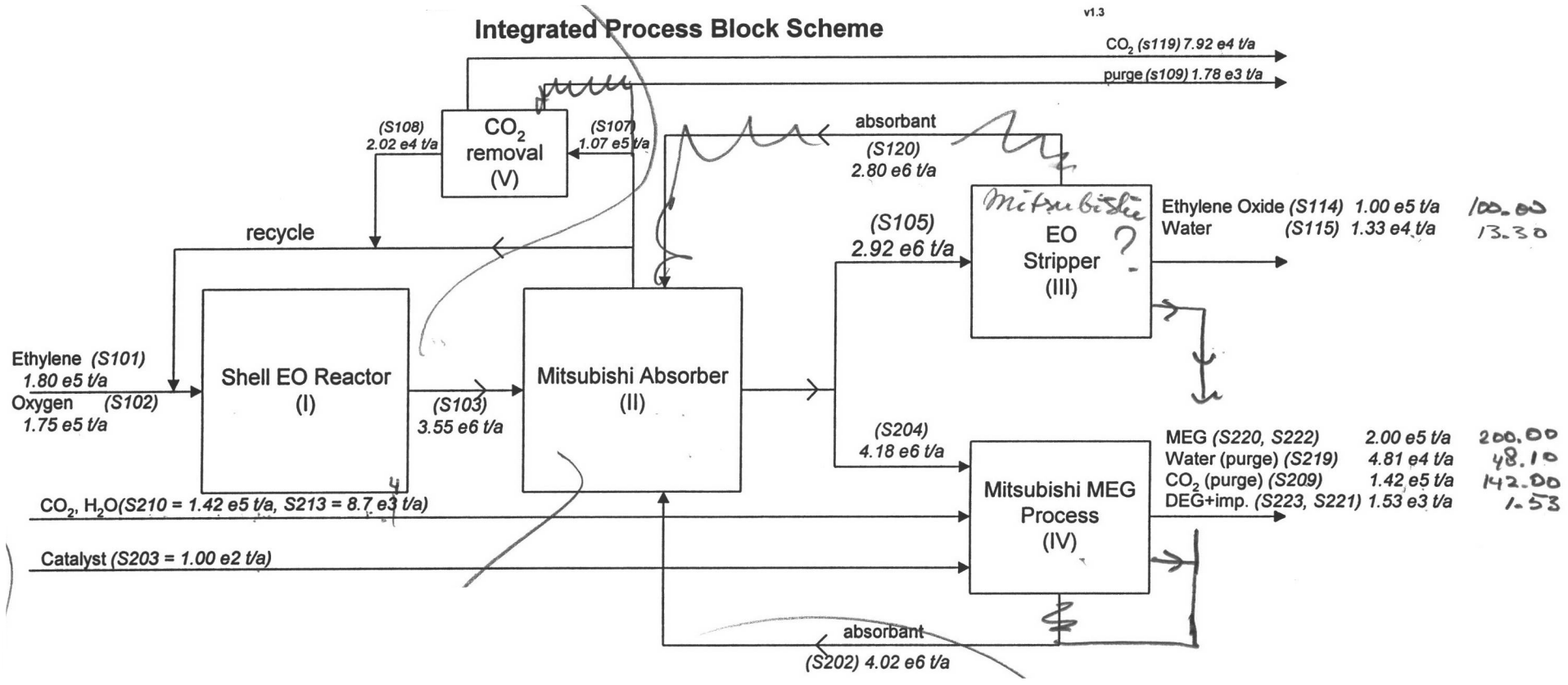


Figure 2.1, Process selection

2.4 BLOCKSCHEME OF THE C OSEN PROCESS



2.5 THERMODYNAMIC PROPERTIES

Experimental data has been gathered for the following systems[14]: water/EO, water/MEG, MEG/DEG, DEG/TEG, and water/MEG/DEG. Henry coefficients have been gathered for the solubility of EO and CO_2 in water[4], and vapor pressure for MEG/water mixtures.

RK-Soave model will be used for the calculation of the solubility's of small amounts of gas in an absorbant. The values as calculated by Aspen Plus have summarily been compared to ~~various~~ the literature values for the absorption of various components in water. It is known[1] that a glycol/carbonate mixture performs approximately 35% better as absorbent, and the Aspen Plus model is consistent with this fact. Varying the fraction of ethylene glycol in a H_2O/MEG/EO mixture has a profound influence on the $K_{VL,EO}$ value (see table 2.2).

Mole fraction EG	$K_{VL,EO}$
0	75.60892
0.2	8.028834
0.4	5.059895
0.6	3.920233
0.8	3.345166
1	-

Table 2.2 EO solubility in water/EG mixtures, Aspen Plus RK-Soave model, T=100°C, P=10 bar.

Ethylene oxide/water mixtures at higher temperature/pressure are modeled by the NRTL model. Although certain sources state concerns that the NRTL does not approximate reality, other sources claim variations between -10% and +10% between NRTL model and experimental values. Aspen Plus bases its parameters on the Dechema databanks, therefore, the NRTL as given by Aspen will in principle be used. However, a more complete comparison between experimental data and the thermodynamic models is still required.

The following T-xy diagrams are included in the appendix D. These diagrams, however, must still be examined more closely for their value, but do give indications for the process conditions required for separations.

Comp. 1	Comp. 2	Pressure (bar)	Model
H_2O	MEG	0.1	RK-Soave
H_2O	EC	0.1	RK-Soave
CO_2	MEG	4	RK-Soave
CO_2	EC	4	RK-Soave
MEG	EC	0.1	RK-Soave
MEG	EO	1,10	NRTL
EC	EO	10	NRTL
EO	Ethylene	10	NRTL
H_2O	EO	1,10	NRTL

[1] Mitsubishi patent EP705826A1

2.6 LIST OF 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/m³	MAC value mg/m³
Acetaldehyde	Acetaldehyde	C_2H_4O	44.05	20.8	-121	783.4^{18}	100
Acetic acid	Ethanoic acid	$C_2H_4O_2$	60.05	117.9[7]	16.6	104.9^{20}	N.A.
Acetylene	Ethyne	C_2H_2	26.04	-84.0	-81.8	620.4^{-82}	N.A.
Argon	Argon	Ar	39.94	-185.7	-189.2	1400^{-186}	N.A.
Carbon dioxide	Carbon dioxide	CO_2	44.01	-78.5[4]	-56.6[5]	1101^{-37}	9000
Carbon monoxide	Carbon monoxide	CO	28.01	-191.3	-207.0	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	1119.7^{15}	N.A
Ethane	Ethane	C_2H_6	30.08	-88.6	-172.0	572.0^{-100}	N.A
Ethylene	Ethene	C_2H_4	28.05	-103.7	-1.69	567.8^{-104}	N.A.
Ethylene glycol	1,2-ehtanediol	$C_2H_6O_2$	62.07	197.3	-13	1108.8^{20}	26
Ethylene oxide	1,2-ethane epoxide	C_2H_4O	44.05	10.6	-111.7	882.1^{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	466.0^{-164}	N.A.
Methanol	Methanol	CH_4O	32.04	65.2[6]	-93.9	791.4^{20}	260
Nitrogen	Nitrogen	N_2	28.02	-195.8	-210.0	808.1^{-196}	N.A.
Oxygen	Oxygen	O_2	32.00	-183.0	-218.4	1149^{-183}	N.A.
Triethylene glycol	2,2'-(1,2-ethanediylbis (oxy))bisethanol	$C_6H_{14}O_4$	150.2	285.0	-7.0	1127.4^{15}	N.A.
Water	Water	H_2O	18.02	100.00	0.0	1.00[4]	N.A.

Notes:
1. At 101.3 kPa
2. Superscript refers to the reference temperature (°C) at which the density is measured compared to water at 4 °C
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

Table 2.3: List of pure component properties

2.6.1 Safety factors concerning Ethylene oxide process

Ethylene oxide must be classified as hazardous with regard to flammability, chemical instability, reactivity and toxicity. In table 2.4 -2.6 the fire and explosion data for EO, ethylene and methane can be found.

Table 2.4: Fire and explosion data for EO[15]

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, K	**2402**
auto-ignition temperature K	**718**
minimum ignition energy[2], J	**$2.7 \cdot 10^{-2}$**
minimum ignition energy[3] (10.4 mol %), J	**$6 \cdot 10^{-5}$**

Table 2.5: Fire and explosion data for ethylene[16]

Property	Value
lower flammable limit[1], mol %	**2.7**
upper flammable limit[1], mol %	**34**
auto-ignition temperature K	**698**
minimum ignition energy[2], J	**$0.07 \cdot 10^{-3}$**

Table 2.6: Fire and explosion data methane

Property	Value
lower flammable limit[1], mol %	**4.4**
upper flammable limit[1], mol %	**16.0**
auto-ignition temperature K	**810**
minimum ignition energy, J	**$0.28 \cdot 10^{-3}$**

1. in air
2. vapor
3. vapor-air mixture

In order to exclude completely a decomposition of EO in the gas phase during the reaction, methane is used as a carrier gas. For a fixed temperature and pressure a certain percentage of carrier gas is required for safe operation.
In appendix F a figure can be found which gives the "safe" carrier gas concentration at a certain pressure and temperature.

3. BASIC ASSUMPTIONS

The capacity of the plant is based on a production of 200,000 MT/a of MEG and 100,000 MT/a of 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 Stork.

On the following page a simplified scheme of the plant, inside the battery limit, is shown. This scheme defines what's inside the battery limit. All in- and outgoing streams pass the battery limit. The feedstock (ethylene and oxygen) is provided by a pipeline. The entering water is demineralized, which assumes a demineralisation unit outside the battery limit. The treatment of purge streams are also considered outside the battery limit as well as the treatment and storage of the products for further transportation.

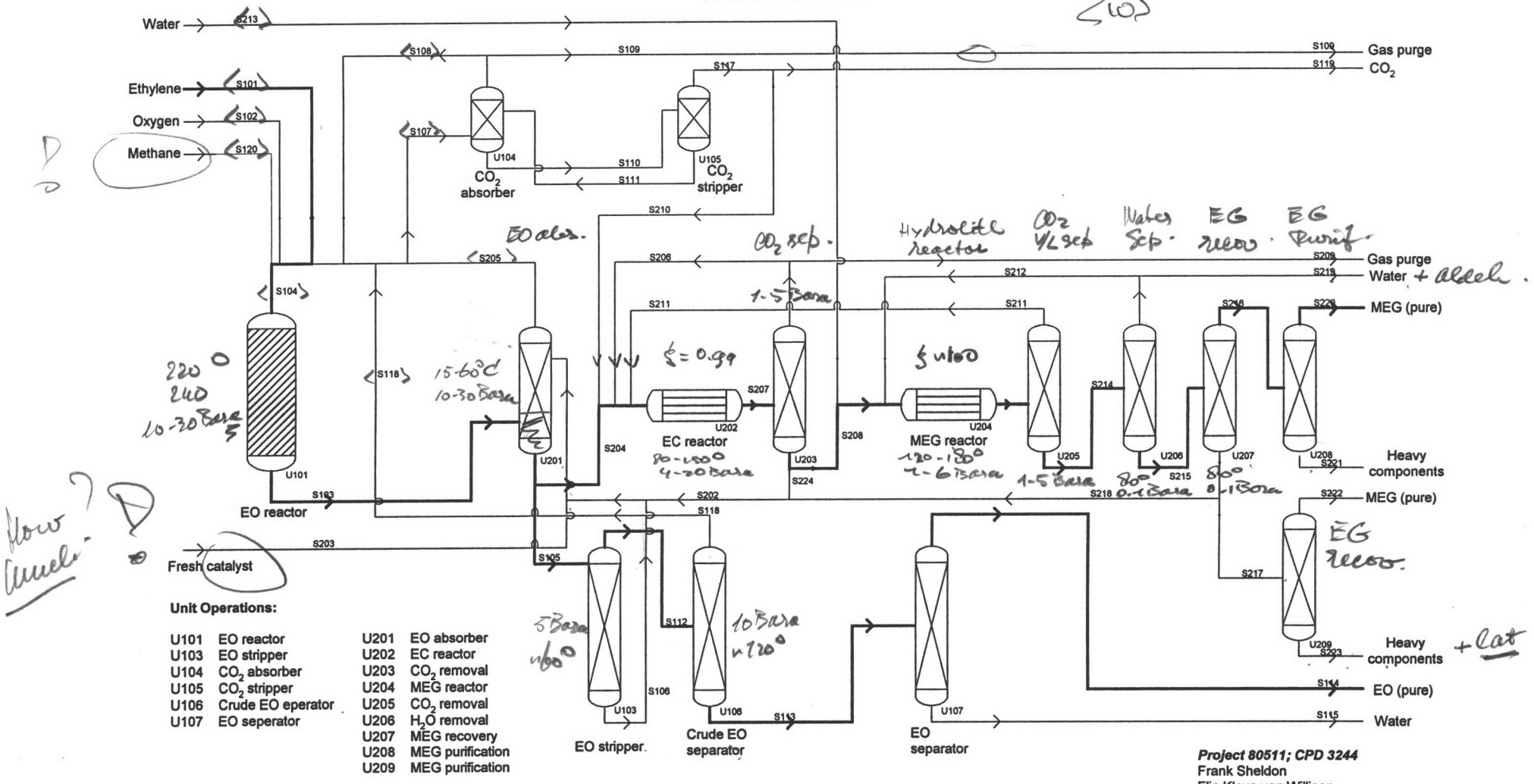
The ethylene, mixed with oxygen ,enter the EO reactor (U101) in which approximately 10 % of the ethylene is converted to ethylene oxide. This stream is fed to the absorber (U201) in which the EO is absorbed in a mixture containing ethylene glycol, ethylene carbonate and catalyst. The stream containing EO is split in two fractions. One third is fed to an EO stripper (U103) and two third is fed to the ethylene carbonate reactor (U202). The other stream leaving the absorber is recycled to the EO reactor. The stream fed to the EO stripper (U103) will be treated to ~~result in purified EO.~~ The stream fed to the EC reactor(U202) will be converted into ethylene carbonate with carbon dioxide. The excess carbon dioxide will be recycled. The reaction stream is split and partly recycled as absorbent and partly fed to the MEG reactor (U204). Water is added and the carbonate is converted to MEG. The water will be separated from the reaction mixture and recycled to the inlet of the MEG reactor. From the resulting stream carbon dioxide is separated and recycled to the EC reactor. In the column U207 MEG is partly coming over the top for further purification and partly over the bottom containing the catalyst and all heavy products. This stream is recycled to the absorber and partly purged to prevent heavy component build-up. This purge will be treated to separate the MEG from this stream.

Components	Temperature °C	Pressure kg/cm²g min	Quantity mt/a	State (V/L/S)	Source	Prices US$/mt
Feedstock						
Ethylene	ambient	25	180,260	V	Pipeline OSBL	520
Oxygen	ambient	27	174,773	V	Pipeline OSBL	50
Products						
EO	ambient	--	100,000	L	--	1,080
MEG	ambient	--	200,000	L	--	650
By-products						
DEG	ambient	--	1440	L	--	551
Aldehydes	ambient	--	2830	L	--	--
CO_2	ambient	--	79200	V	--	--
Other Chemicals						
Methane	ambient	25	2660	V	Pipeline OSBL	120
Water	ambient	5	87000	L	unit OSBL	2.5
Catalysts	Composition		Quantity	shape	Lifetime	Prices
EO-catalyst	silver on alumina		125 M^3/3y	cylinder	3 Years	8,800
MEG catalyst	an alkyl phosphonium salt		530	salt	---	10,000

The specifications of the feedstock, products, byproducts etc. are given in appendix B.

Ethylene Oxide and Ethylene Glycol from Ethylene

Process Scheme



Unit Operations:

U101	EO reactor	U201	EO absorber
U103	EO stripper	U202	EC reactor
U104	CO_2 absorber	U203	CO_2 removal
U105	CO_2 stripper	U204	MEG reactor
U106	Crude EO eperator	U205	CO_2 removal
U107	EO seperator	U206	H_2O removal
		U207	MEG recovery
		U208	MEG purification
		U209	MEG purification

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Note: More detailed specifications of the Unit operations are found in appendix C.

4. ECONOMIC MARGINS

The economic margin is defined as the difference between income from sales minus costs for feedstock, chemicals, etc.

Costs of used components in this process are summarized in table 4.1. The Economic margin is calculated as well.

The annual costs for the EO catalyst are calculated by dividing the price for one reactor fill by the catalyst lifetime, which is three years.

Component	Quantity mt/a	Price US$/mt	Costs US$/a	Sales US$/a
Ethylene	180000	$520	$93,600,000	
Oxide	175000	$50	$8,750,000	
Methane	2660	$120	$319,200	
EO catalyst	***		$432,540	
Water	87000	$3	$217,500	
MEG catalyst	530	$10,000	$5,300,000	
EO	100000	$1,080		$108,000,000
MEG	200000	$650		$130,000,000
			108,619,240	238,000,000
		MARGIN	$129,380,760	108,619,240

Table 4.1: Margin calculation for EO/MEG production

129,380,760

The discounted cash flow rate of return (DCFRR) is taken to be 10%. Some preliminary financial scenarios can be calculated now. Basic assumptions taken are that the initial investments are done during a period of 3 years proceeding plant operation. Division of investments over these 3 years is arbitrary. Annual operational costs are estimated as $ 10 mil. and 40% income tax is assumed

When a pay back time of 4 years is assumed, a financial scenario can be calculated. The investments done for this scenario can then be seen as maximal investments for this pay back period. For this scenario, Returns on investments are also calculated as well as Present Worth Indices. Results are demonstrated in table 4.2 below.

Year	-2	-1	0	1	2	3	4
Investments	-$ 5.00E+7	-$ 2.31E+8	-$ 1.00E+7	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0
Exploitation costs	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 1.00E+7	$ 1.00E+7	$ 1.00E+7	$ 1.00E+7
Total costs	$ 5.00E+7	$ 2.31E+8	$ 1.00E+7	$ 1.09E+8	$ 1.09E+8	$ 1.09E+8	$ 1.09E+8
Total Earnings	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 2.38E+8	$ 2.38E+8	$ 2.38E+8	$ 2.38E+8
Capital Costs	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 2.91E+7	$ 2.91E+7	$ 2.91E+7	$ 2.91E+7
Cash Flow	-$ 5.00E+7	-$ 2.31E+8	-$ 1.00E+7	$ 8.93E+7	$ 8.93E+7	$ 8.93E+7	$ 8.93E+7
Present value	-$ 6.11E+7	-$ 2.09E+8	-$ 1.00E+7	$ 8.08E+7	$ 7.31E+7	$ 6.61E+7	$ 5.98E+7
Net present value	-$ 6.11E+7	-$ 2.70E+8	-$ 2.80E+8	-$ 1.99E+8	-$ 1.26E+8	-$ 5.98E+7	$ 7.45E-9
Return on Investm.	-0.22	-0.75	-0.04	0.29	0.26	0.24	0.21
Present Worth Index	-0.22	-0.96	-1.00	-0.71	-0.45	-0.21	0.00

Year	5	6	7	8	9	10
Investments	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0	$ 0.00E+0
Exploitation costs	$ 1.00E+7	$ 1.00E+7	$ 1.00E+7	$ 1.00E+7	$ 1.00E+7	$ 1.00E+7
Total costs	$ 1.09E+8	$ 1.09E+8	$ 1.09E+8	$ 1.09E+8	$ 1.09E+8	$ 1.09E+8
Total Earnings	$ 2.38E+8	$ 2.38E+8	$ 2.38E+8	$ 2.38E+8	$ 2.38E+8	$ 2.38E+8
Capital Costs	$ 2.91E+7	$ 2.91E+7	$ 2.91E+7	$ 2.91E+7	$ 2.91E+7	$ 2.91E+7
Cash Flow	$ 8.93E+7	$ 8.93E+7	$ 8.93E+7	$ 8.93E+7	$ 8.93E+7	$ 8.93E+7
Present value	$ 5.41E+7	$ 4.90E+7	$ 4.43E+7	$ 4.01E+7	$ 3.63E+7	$ 3.28E+7
Net present value	$ 5.41E+7	$ 1.03E+8	$ 1.47E+8	$ 1.88E+8	$ 2.24E+8	$ 2.57E+8
Return on Investm.	0.19	0.18	0.16	0.14	0.13	0.12
Present Worth Index	0.19	0.37	0.53	0.67	0.80	0.92

Table 4.2: Financial scenario for the EO EG process

In this scenario the maximal investments are spread out over a 3 year period proceeding plant operation. $50 million is invested 3 years before plant operation, $231 2 years before, and $10 million 1 year before plant operation. Capital Costs are spread out over the economical lifetime of the plant which is assumed to be 10 years.
What can be seen is that the net present value after 4 years of operation is nearly zero. The present worth index after 10 years of operation is 92% which means that after 10 years of operation, 92% of initial investments is earned on top of paying back all investments.

List of abbreviations

BOD	Basis of design
CAPEX	Capital expenses
DCFRR	Discounted- cash- flow rate of return
DEG	Diethylene glycol
EC	Ethylene carbonate
EG	Ethylene glycol
EO	Ethylene oxide
EP	End point
IBP	Initial boiling point
MAINTEX	Maintenance expenses
MEG	Monoethylene glycol
Mmt	Mega metric tons
NFW	Net future worth
NPW	Net present worth
OPEX	Operational expenses
OSBL	Outside battery limit
PC	Propylene carbonate
POT	Pay out time
ROI	Return on investment
ROR	Rate of return
TEG	Triethylene glycol

Appendix A: MEG Process comparing from basic principles

Explained topics

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

Notes:

1. catalyst dependent. Best found values are given here.
2. Single pass conversion
3.

Abbreviations:

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

Compared Topics

Process	Conversion[2]	Selectivity	H_2O/EO	number of main components	Implementation in industry
Conventional Thermal	=	-	-	+	++
Conventional catalytic[1]	=	+	+	++	-
Carbonate route process	=	++	++	-	-

The two most important topics are selectivity and H_2O/EO ratio. The selectivity because ethylene efficiency (Ethylene costs are a major part of the OPECS) is an important economic factor. The H_2O/EO ratio because the large effect it has on the energy consumption in the separation section.

Appendix B: Specifications

FEEDSTOCK SPECIFICATIONS

- **Ethylene**

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
Sulfur (H_2S), vol ppm max.	1
Methanol, vol ppm max.	10
water, vol ppm max.	2
Chlorine (Cl), vol ppm max.	1

- **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
Sulfur compounds, wt ppm, max.	5
Halogen compounds, vol ppm, max	5

PRODUCT SPECIFICATIONS

- **Fiber-grade 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 acetic), 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
Distillation range at 760 mm. Hg	
IBP, °C min	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.	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
Distillation range at 760 mm. Hg	
IBP to DP	4°C including 246°C

UTILITY SPECIFICATIONS

- **Steam**

Price, US$/mt	9.28
High pressure (for start up), kg/cm²g min	32
High pressure, kg/cm²g min	25
Medium pressure, kg/cm²g min	14
Low pressure, kg/cm²g min	4

- **Power**

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

- **Cooling water**

Price, US$/m³	0.02
Supply pressure at grade, kg/cm²g	7
Return pressure at grade, kg/cm²g	2
Inlet temperature, °C	33
Temperature difference, °C	10

- **Process water**

Price, US$/m³	2.5
Quality	Demineralized
Pressure, kg/cm²g	5
Temperature	Ambient
pH, units	7.5 to 8.5
Conductivity, μΩ/cm	0.5 to 1.0
SiO_2, wt ppm max.	0.1

- **Nitrogen**

Price, US$/nm³	0.02
Pressure, kg/cm²g	7
Temperature	Ambient
Purity, vol% min.	99.9
Oxygen, vol ppm max.	10

- **Instrument air**

Price, US$/nm³	0.02
Pressure, kg/cm²g	5
Temperature	Ambient
Quality	Dry, Oil-free
Dew point, °C	-40

CATALYST SPECIFICATIONS

- **EO catalyst**

Price, US$/mt	8,800
composition	silver on alumina catalyst
shape	cylinder
bulk density kg/m³	850

- **MEG catalyst**

Price, US$/mt	10,000
composition	Tributylmethylphosphonium iodide
State	dissolved

Appendix C: Information about the unit operations

U101 Ethylene oxide reactor: (Type: multi-tubular)

Reaction	Ethylene to ethylene oxide	
catalyst	heterogeneous	
	silver cat supported on alpha alumina	
Conversion	10% per pass	
Process conditions	Temperature	Pressure
	220 - 240 °C	10-30 bar

U103 Ethylene oxide stripper:

Process	ethylene oxide stripped from EC/MEG with steam	
Process conditions	Temperature	Pressure
	>60 °C	5 bar
Thermodynamic model	NRTL	

U106 Crude ethylene oxide separator:

Process	light impurities distilled	
Process conditions	Temperature	Pressure
	120 °C	10 bar
Thermodynamic model	NRTL	

U107 Ethylene oxide separator:

Process	high purity EO distilled	
Process conditions	Temperature	Pressure
	120 °C	5 bar
Thermodynamic model	NRTL	

U201 Ethylene oxide absorber:

Absorbing solution	50 wt%	50 wt%
	Ethylene glycol / ethylene carbonate (ratio 0.3 to 4)	mainly: water (3-15%), catalyst (1.5-3.5%)
Process conditions	Temperature	Pressure
	15-60 °C	10-30 bar
Gas/liquid ratio	0.2-1.0	
Unit operation Type	Packed- or plate column	
Thermodynamic model	RK-Soave	Parameter: solubility

U202 Ethylene Carbonate reactor (Type: multi-tubular)

Reaction	Ethylene oxide to ethylene carbonate	
catalyst	homogeneous	heterogeneous
	halogenated organic phosponiumsalts	quaternary ammonium salts
Conversion	99%	
Process conditions	Temperature	Pressure
	80-150 °C	4-20 bar
Retention time	12-120 min	

U203 Carbon dioxide separator

Top	unused carbon dioxide, (oxygen, ethylene, etc. from EO plant?)
Bottom	ethylene glycol and ethylene carbonate + catalyst
Pressure	lowered to 1-5 bar
Thermodynamic model	RK-Soave

U204 Hydrolytic reactor (Type: multi-tubular)

Reaction	ethylene carbonate to ethylene glycol (MEG +DEG)	
Catalyst	idem as U202 or none	
Conversion	nearly 100%	
Selectivity	MEG	DEG
	99%	1%
Process conditions	Temperature	Pressure
	120-180 °C	1-6 bar
Retention time	10-180 min	

U205 Carbon dioxide gas/liquid separator

Top	carbon dioxide
Bottom	ethylene glycol, ethylene carbonate, water, catalyst
Pressure	lowered to 1-5 bar

U206 Water separation

Top	water	
Bottom	ethylene glycol, ethylene carbonate, catalyst	
Process conditions	Temperature	Pressure
	80 °C	~0.1 bar
Thermodynamic model	RK-Soave	

U207 Ethylene glycol recovery

Top	ethylene glycol (raw)	
Bottom	diethylene glycol, ethylene glycol(+catalyst), ethylene carbonate	
Process conditions	Temperature	Pressure
	80 °C	<0.1 bar
Thermodynamic model	RK-Soave	

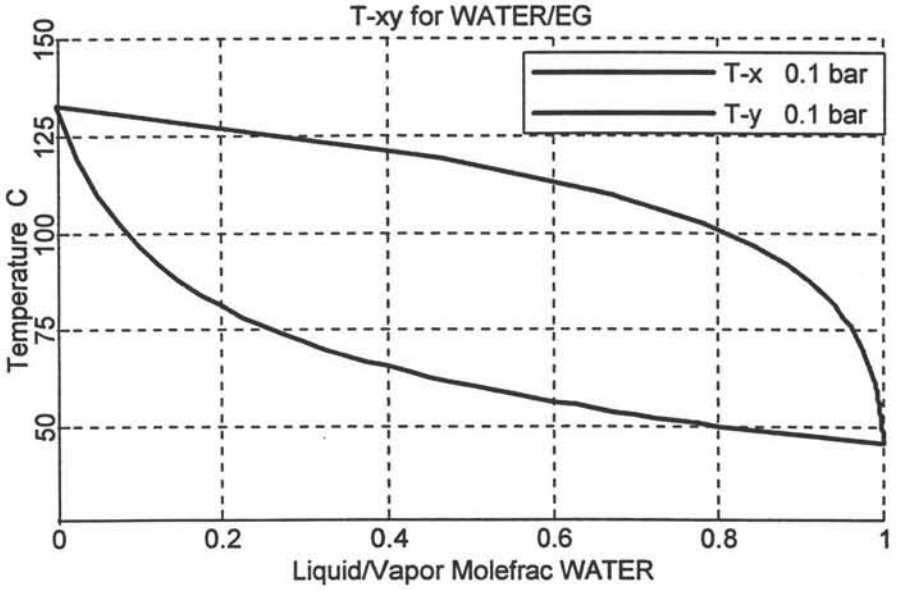
U208 Ethylene glycol purification

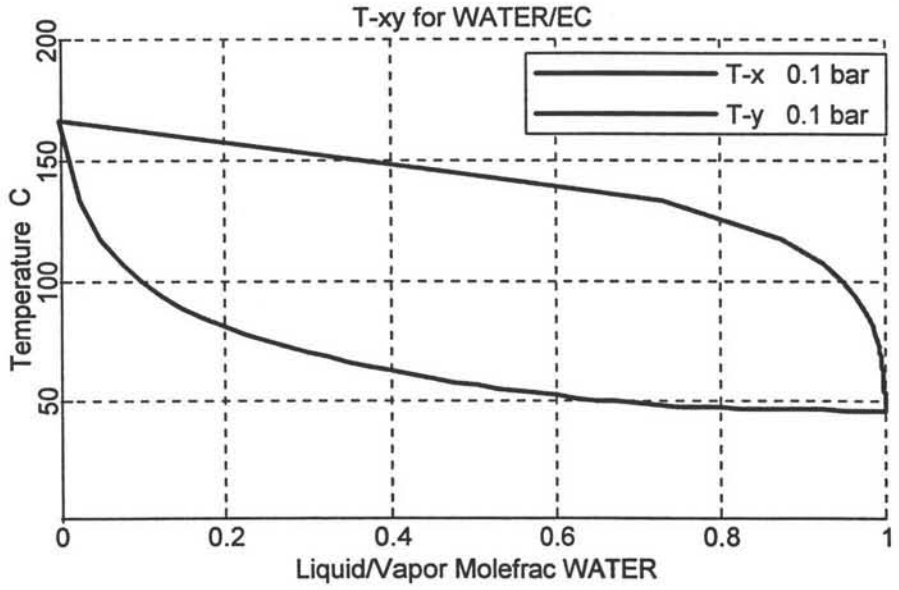
Top	ethylene glycol (purified)
Bottom	diethylene glycol, traces of other components

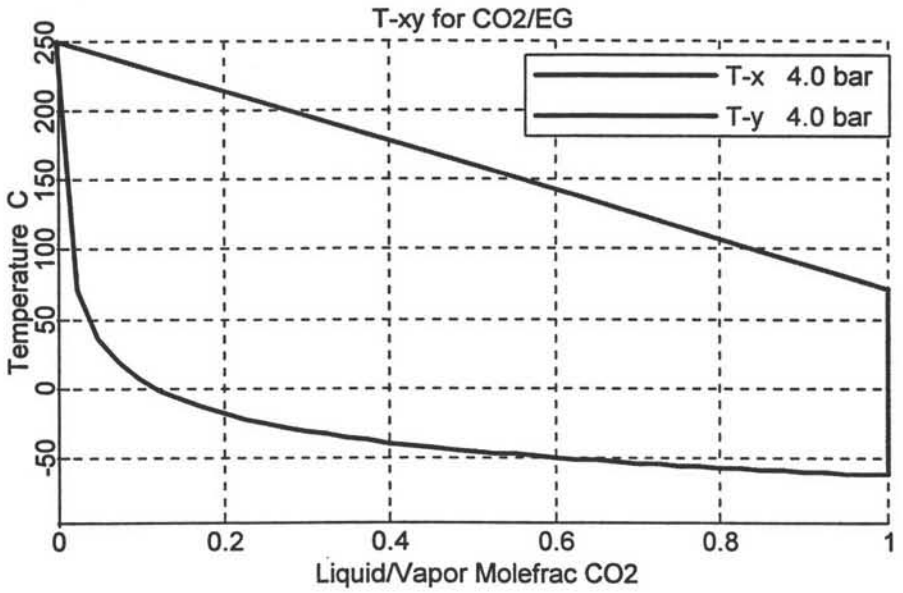
U209 Ethylene glycol recovery

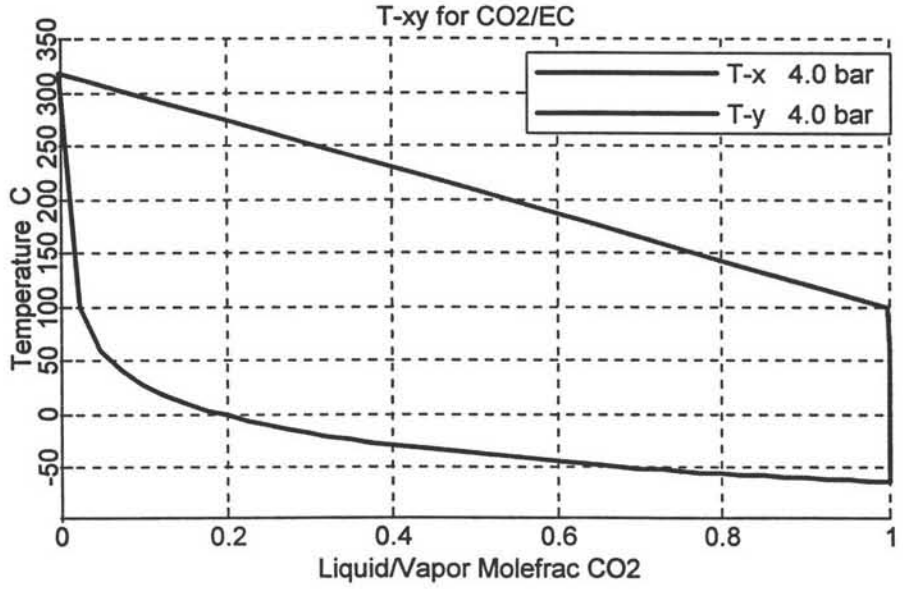
Top	ethylene glycol (purified)
Bottom	concentrated catalyst and other heavy products

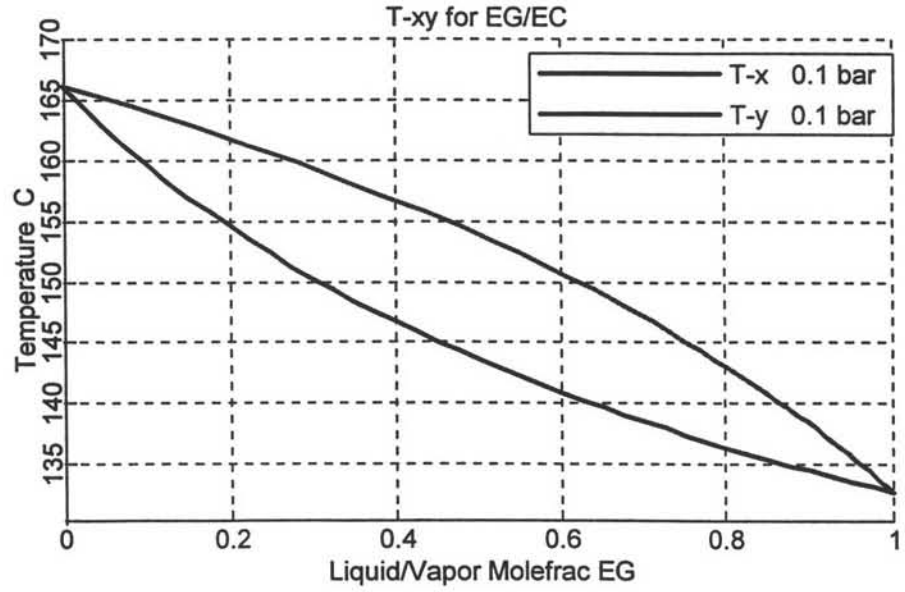
Appendix D: T-xy diagrams

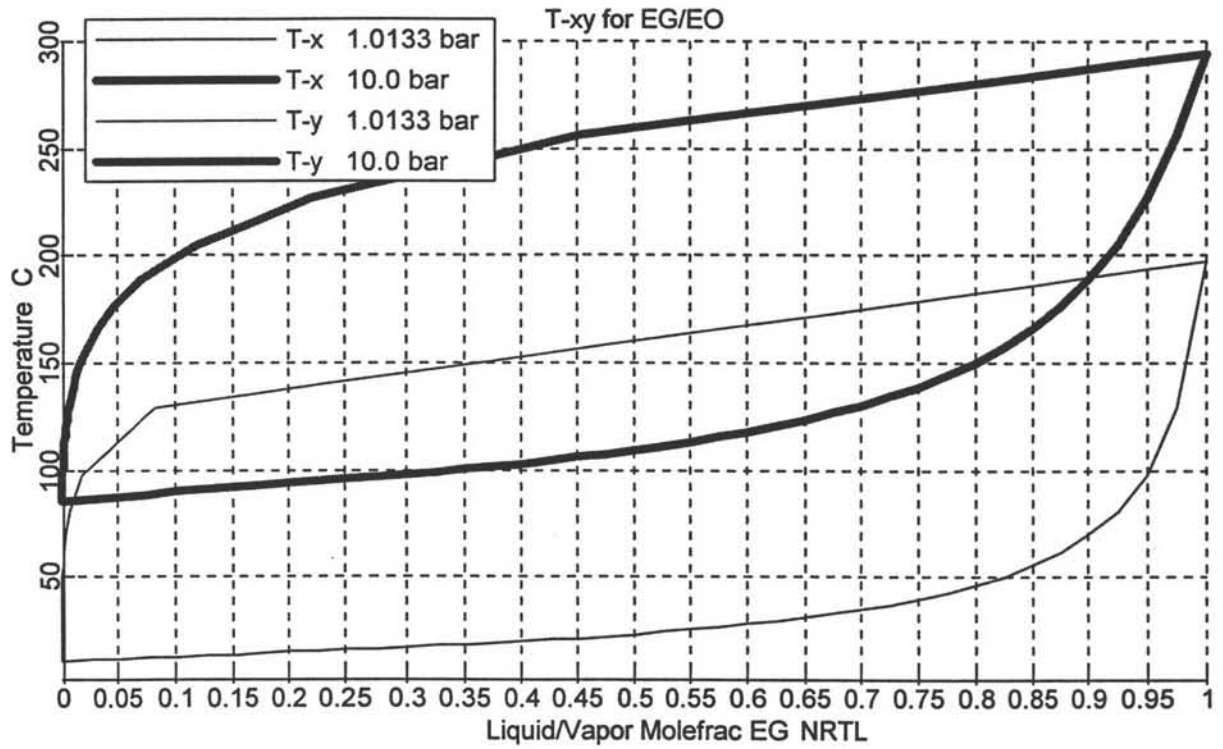


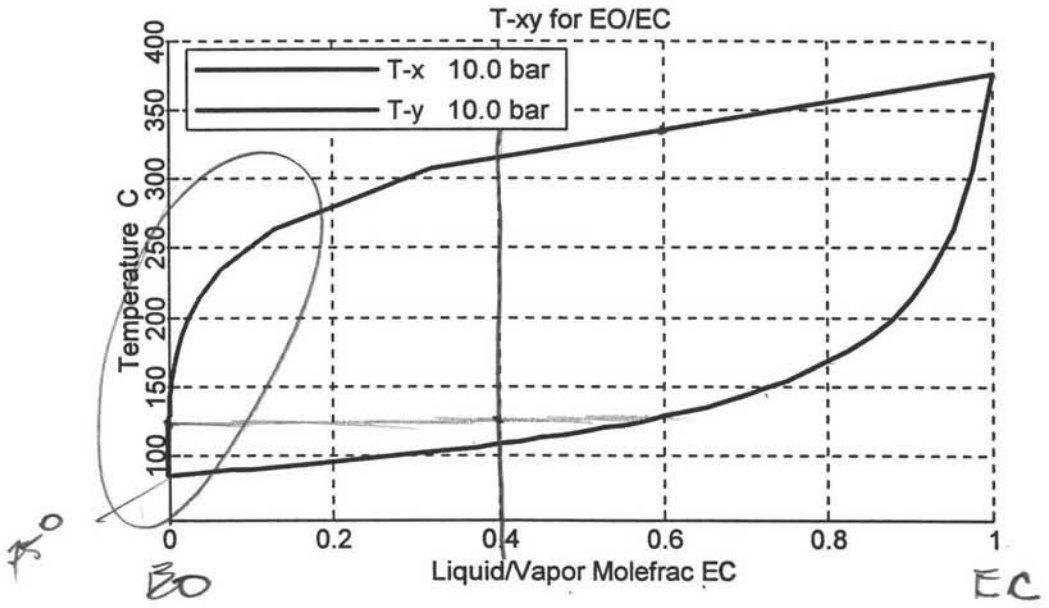


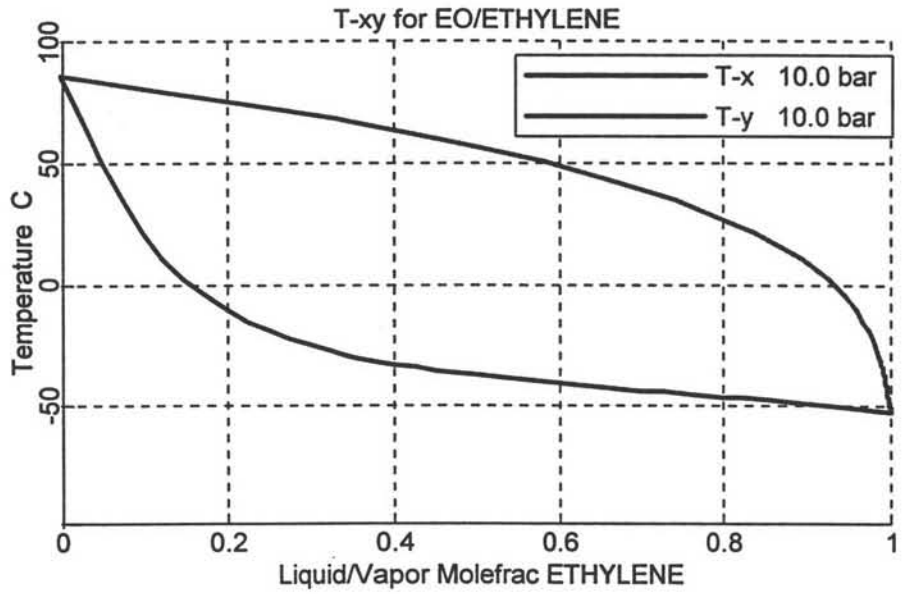

T-xy for CO2/EG
T-x 4.0 bar
T-y 4.0 bar
Temperature C
250
200
150
100
50
0
-50
0
0.2
0.4
0.6
0.8
1
Liquid/Vapor Molefrac CO2

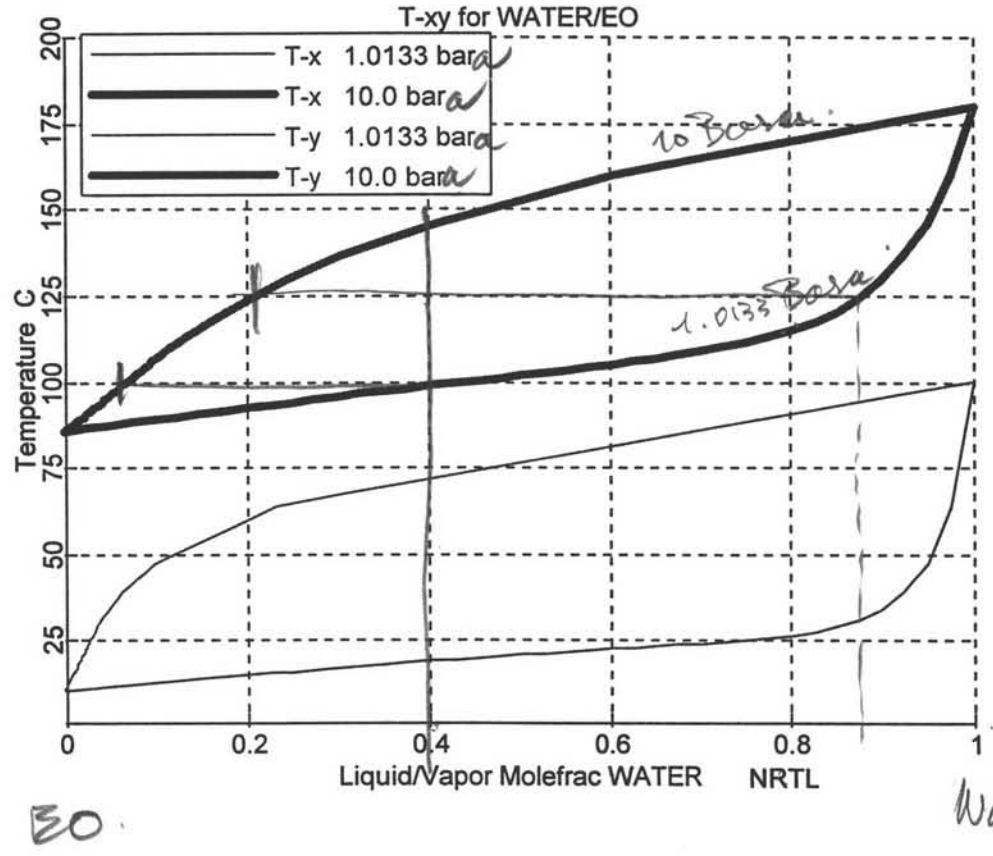

T-xy for CO2/EC
T-x 4.0 bar
T-y 4.0 bar
Temperature C
350
300
250
200
150
100
50
0
-50
0
0.2
0.4
0.6
0.8
1
Liquid/Vapor Molefrac CO2


T-xy for EG/EC
T-x 0.1 bar
T-y 0.1 bar
Temperature C
170
165
160
155
150
145
140
135
0
0.2
0.4
0.6
0.8
1
Liquid/Vapor Molefrac EG


T-xy for EG/EO
T-x 1.0133 bar
T-x 10.0 bar
T-y 1.0133 bar
T-y 10.0 bar
Temperature C
300
250
200
150
100
50
0 0.05 0.1 0.15 0.2 0.25 0.3 0.35 0.4 0.45 0.5 0.55 0.6 0.65 0.7 0.75 0.8 0.85 0.9 0.95 1
Liquid/Vapor Molefrac EG NRTL


T-xy for EO/EC
T-x 10.0 bar
T-y 10.0 bar
Temperature C
400
350
300
250
200
150
100
0
0.2
0.4
0.6
0.8
1
Liquid/Vapor Molefrac EC
EO
EC


T-xy for EO/ETHYLENE
T-x 10.0 bar
T-y 10.0 bar
Temperature C
100
50
0
-50
0
0.2
0.4
0.6
0.8
1
Liquid/Vapor Molefrac ETHYLENE


T-xy for WATER/EO
T-x 1.0133 bara
T-x 10.0 bara
T-y 1.0133 bara
T-y 10.0 bara
10 Bara
1.0133 Bara
Temperature C
200
175
150
125
100
75
50
25
0
0.2
0.4
0.6
0.8
1
Liquid/Vapor Molefrac WATER
NRTL
EO
Water

Appendix E:

EO Catalyst Comparison

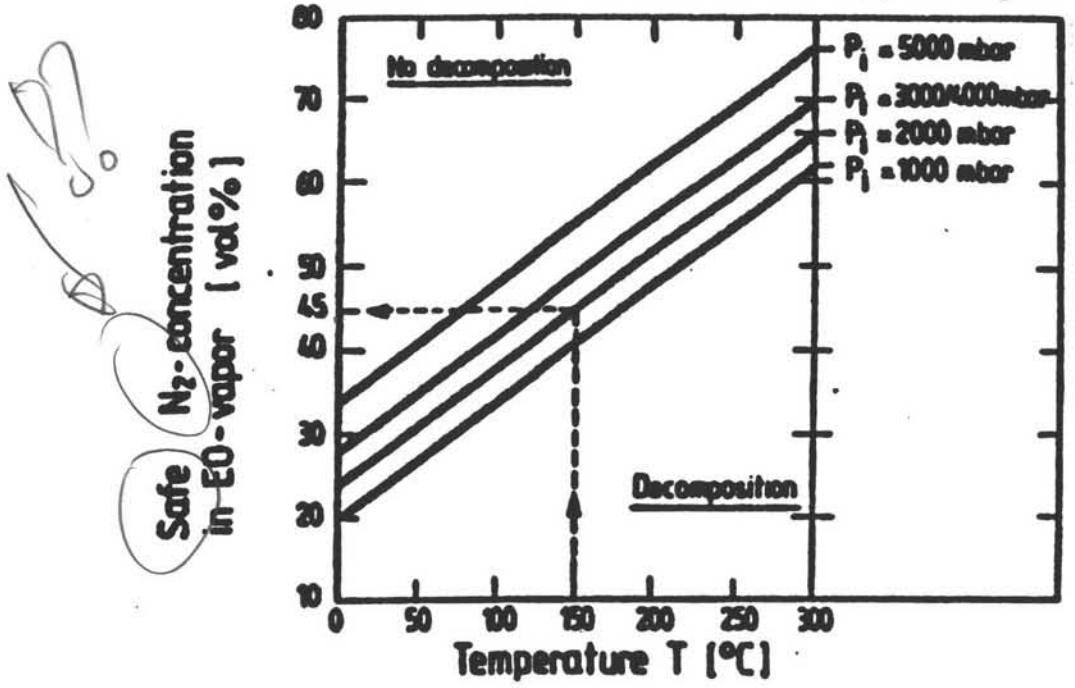
Company	date	patent #	selectivity	conversion ethylene	conversion oxygen	modifier=inhibitor	promoter	lifetime	diluent/inhibitor ratio	surface area m^2/g B.E.T.	porosimetry	support
ICI	16/8/89	357292	Discarded - ethylene concentration must be higher than 35 vol%, which is higher than our anticipated process 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 A	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	12/4/94	624398 A	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	8/5/87	247414 A	82.1			ethylene dichloride	Na, K, Rb, Vs			.6-2	H2O: 20-50%	alpha alumina, .5-12% Si
Nippon Shokubai	25/8/99	937498 A	81	10			Cs only					alpha alumina
Scientific Design	11/4/99	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	11/4/99	99/11370	84.6,83.8	160kg/hr/m^3 kat			Cs, F, S, Ge, Sn			.5-1.3	Hg: 20-70 %	
Scientific Design	11/4/99	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	2/3/99	95/05896	Only startup			ethyl chloride		10 hrs			H2O: 46%, 6.2 mu	alpha alumina, .8% Si
Shell	9/5/96	96/13329	85		40%	ethyl chloride	Cs, Re, P, B	2-3 days		.1-3	H2O: 75%	alpha alumina,
Shell	11/5/94	625370 A	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 A	86 (ex2)			ethyl chloride		100 days, dec. to 80%				alpha alumina, 1.3% Si
Shell	30/10/8	266015	86			VC 4.5-5.5 ppm		60, sel. const!		0.42	H2O: 36%, Hg:2.7 mu	alpha alumina, 2% Si
UCC	4/10/82	76504 A1	71-75			ethyl chloride						alpha alumina, .74% Si

Appendix E:

EO Catalyst Comparison

wt% Ag	Li	Cs	Others	ethylene/O2 ratio mole %	diluent	T Celsius	P bar		
6-28				.05-100	40-70% methane, .3-5% ethane	201-207			
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
5-50%		250-2000 ppm	W:5-700 ppm, Na: 92 ppm	3.50	60% N2	235-255	7	rings? Saddle's? High conversio	Other metals, see source
			Na: 0.8-2%			215	15	Carrier (Si) and Na analysis	
				2.69		240	20	Carrier must be washed with H2O	
8-15%		600 ppm	S: 34 ppm, Best with Sb: 130 ppm		78% N2+CO2	240	20.5	Equivalent diam. 4-8mm for supports	
		640,580	F:180,75, S:34,34, Ge:75,0, Sn:0,125		N2, CO2	240	20.4		
		600-800	F:75, S: ~40, Lan:175		N2, CO2	240	20.4		
		340,460	Re: 280, P:10,31;B:4,42	3.53	5% CO2, 54.5 % N2	230	15.5		
		430,460	Re:280, S:48, Rare earth:80	3.53	5% CO2, 54.5 % N2		15.5		
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	
13.5		500	Re: 260, S:35	3.53	7% CO2, 54.5% N2	250-256	14.5	ex 10	
13.1		.00906 wt%	K:.00268wt%	0.75	N2	250-260			

Appendix F: "safe" carrier gas concentration at a certain pressure and temperature



Literature (Odd place !!).

[1] European Chemical News, 11-17 October 1999, pg 20.
[2] Kirk-Othmer, 4th edition, vol. 9, pg 942.
[3] Leaflet: The Shell Ethylene Oxide/Ethylene Glycols Process, Shell Chemicals.
[4] Kirk-Othmer, 4th edition, vol. 9, pg 939-940.
[5] See appendix E for a comparison between different catalysts
[6] EP 207490, issued to Atochem, 01/07/86, describes the purification using water as absorbent.
[7] Kirk-Othmer, 4th edition, vol. 9, pg 924
[8] C.J. van Tiggelen.
[9] EP 339748, EP 352849, WO 95/08545, EP 352850 are some of the patents that describe the Shell process.
[10] Ullman's Volume A10, pg 105
[11] EP 0776890A2, Mitsibushi patent, 29/11/1999
[12] Mitsubishi patent EP705826A1
[13] US patent 4233221, to DOW Chemical Co., 11/11/1980.
[14] Dechema
[15] Prevention of Ethylene oxide decomposition, R. Siwek, E. Rosenberg, 1989
[16] Chemiekaarten, 11e editie, 1996

Planning CPD 3244
Project nr. 80511

Start or the project: 10/8/99
End of the project (based on 12 SP) 1/7/99

week	Date	finished work
44	2-Nov	BOD review
45		
46	19-Nov	Concept
47		
48		
49		
	15-Dec	Facilities Stork no longer available
50	17-Dec	Detailed
51		
52	25/12-2/1	KERST
1	7-Jan	CPD-document
2	14-Jan	CPD-Final assesment

Concept
Aspen flowsheet (chemical)
Equipment design
Mass + Heat balances
Heat integration development
Process control development

Detailed
Process control
Heat integration
Economy
wastes
SHES
Aspen flowsheet (total)

Optional
Modeling serial process and comparing with integrated process
(see fig 2.1 scheme 1 and 3 of the BOD)

Addendum to fig 4.2 Financial scdenario of the BOD

Year	-2	-1	0	1	2	3	4	5	6	7	8	9	10
Investments	M$ - 50.00	M$ - 183.51	M$ - 10.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00	M$ 0.00
Exploitation costs	M$ 0.00	M$ 0.00	M$ 0.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00	M$ 10.00
Total costs	M$ 50.00	M$ 183.51	M$ 10.00	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62	M$ 108.62
Total earnings	M$ 0.00	M$ 0.00	M$ 0.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00	M$ 238.00
Capital costs	M$ 0.00	M$ 0.00	M$ 0.00	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35	M$ 24.35
Cash Flow	M$ - 50.00	M$ - 183.51	M$ - 10.00	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37	M$ 87.37
Present Value	M$ - 61.07	M$ - 202.81	M$ - 10.00	M$ 79.05	M$ 71.53	M$ 64.72	M$ 58.56	M$ 52.99	M$ 47.95	M$ 43.39	M$ 39.26	M$ 35.52	M$ 32.14
Net Present Value	M$ - 61.07	M$ - 263.88	M$ - 273.88	M$ - 194.82	M$ - 123.29	M$ - 58.56	M$ - 0.00	M$ 52.99	M$ 100.94	M$ 144.33	M$ 183.58	M$ 219.11	M$ 251.25
Return on Investm.	-0.22	-0.74	-0.04	0.29	0.26	0.24	0.21	0.19	0.18	0.16	0.14	0.13	0.12
Present Worth Index	-0.22	-0.96	-1.00	-0.71	-0.45	-0.21	0.00	0.19	0.37	0.53	0.67	0.80	0.92

Addendum to chapter 2 of the BOD

Reaction Heats of the occurring reactions

Combustion enthalpy E	-1323 kJ/mol
E +1/2 O2 -> EO	-106.7 kJ/mol
EO + H2O -> MEG	-93.1 kJ/mol
EO + CO2 -> EC	-60.8 kJ/mol
EC + H2O -> MEG + CO2	-32.3 kJ/mol

Thermal process (Heat integration) Energy produced compared with the energy needed

Heat produced from EO combustion	1.19E+12	kJ/a
Reaction Heat from EO Process	5.90E+11	kJ/a
Total Heat produced in the EO process	1.78E+12	kJ/a

Heat required for water removal	710.6	kJ/mol MEG	patent: Ep0776890 A2
steam required for water removal	5.5	t steam/t MEG	
Heat required for water removal per year	2.29E+12	kJ/a	

The extra heat requirement for removing a 20 fold excess of water using a multistage evaporator after MEG formation is then as stated below:
Note: These calculations are made on simple 'heat of reaction' and 'heat of evaporation' data

required	2.29E+12	kj/a	steam per year	1.10E+06	ton
produced	1.78E+12	kj/a	Energy per t steam	2.08E+06	kJ/ton
extra heat:	5.12E+11	kj/a	**extra steam/year**	2.46E+05	t steam/a

EO reactor (I)

In		Out	
S104	3.55E+06 t/a	S103	3.55E+06 t/a

Absorber (II)

In		Out	
S103	3.55E+06 t/a	S205	3.20E+06 t/a
S202	4.02E+06 t/a	S204	4.18E+06 t/a
S120	2.80E+06 t/a	S105	2.92E+06 t/a
total	1.04E+07 t/a	total	1.03E+07 t/a

CO2 removal (V)

In		Out	
S107	1.07E+05 t/a	S108	2.20E+04 t/a
		S109	5.33E+03 t/a
		S119	7.92E+04 t/a
total	1.07E+05 t/a	total	1.07E+05 t/a

EO Stripper (III)

In		Out	
S105	2.92E+06 t/a	S114	1.00E+05 t/a
		S115	1.33E+04 t/a
		S120	2.80E+06 t/a
total	2.92E+06 t/a	total	2.92E+06 t/a

Mitsubishi MEG process (IV)

In		Out	
S204	4.18E+06 t/a	S202	4.02E+06 t/a
S210	1.42E+05 t/a	S220,222	2.00E+05 t/a
S213	8.70E+04 t/a	S219	4.81E+04 t/a
S203	1.00E+02 t/a	S209	1.42E+05 t/a
		S223,221	1.53E+03 t/a
total	4.41E+06 t/a	total	4.41E+06 t/a

Total process

In		Out	
S101	1.80E+05 t/a	S109	5.33E+03 t/a
S102	1.75E+05 t/a	S114	1.00E+05 t/a
S210	1.42E+05 t/a	S115	1.33E+04 t/a
S213	8.70E+04 t/a	S220,222	2.00E+05 t/a
S203	1.00E+02 t/a	S219	4.81E+04 t/a
		S209	1.42E+05 t/a
		S223,221	1.53E+03 t/a
		S119	7.92E+04 t/a
total	5.84E+05 t/a	total	5.89E+05 t/a